11 Advanced Wastewater Treatment

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Advanced wastewater treatment is defined as the additional treatment needed to remove suspended, colloidal, and dissolved constituents remaining after conventional secondary treatment. Dissolved constituents may range from relatively simple inorganic ions, such as calcium, potassium, sulfate, nitrate, and phosphate, to an ever-increasing number of highly complex synthetic organic compounds. In recent years, the effects of many of these substances on the environment have become understood more clearly. Research is ongoing to determine (1) the environmental effects of potential toxic and biologically active substances found in wastewater and (2) how these substances can be removed by both conventional and advanced wastewater-treatment processes. As a result, wastewater-treatment requirements are becoming more stringent in terms of both limiting concentrations of many of these substances in the treatment plant effluent and establishing whole effluent toxicity limits, as outlined in Chap. 2. To meet these new requirements, many of the existing secondary treatment facilities will have to be retrofitted and new advanced wastewater-treatment facilities will have to be constructed. Therefore, the purpose of this chapter is to present an introduction to the subject of advanced wastewater treatment. The chapter contains an expanded discussion of the need for advanced wastewater treatment, an overview of the available technologies used for the removal of the constituents of concern, identified previously in Chap. 2, and an introduction to the more important of these technologies as applied to the removal of specific constituents found in wastewater. The ultimate disposal of residuals from advanced wastewater treatment is considered in Chap. 14.

11-1 NEED FOR ADVANCED WASTEWATER TREATMENT

The need for advanced wastewater treatment is based on a consideration of one or more of the following factors.

1. The need to remove organic matter and total suspended solids beyond what can be accomplished by conventional secondary treatment processes to meet more stringent discharge and reuse requirements.
2. The need to remove residual total suspended solids to condition the treated wastewater for more effective disinfection.
3. The need to remove nutrients beyond what can be accomplished by conventional secondary treatment processes to limit eutrophication of sensitive water bodies.
4. The need to remove specific inorganic (e.g., heavy metals) and organic constituents (e.g., MBTE and NDMA) to meet more stringent discharge and reuse requirements for both surface water and land-based effluent dispersal and for indirect potable reuse applications (e.g., groundwater recharge).
5. The need to remove specific inorganic (e.g., heavy metals, silica) and organic constituents for industrial reuse (e.g., cooling water, process water, low-pressure boiler makeup water, and high-pressure boiler water).

With increased scientific knowledge derived from laboratory studies and environmental monitoring concerning the impacts of the residual constituents found in secondary effluent, it is anticipated that many of the methods now classified as advanced will become conventional within the next 5 to 10 years.

Compounds containing available nitrogen and phosphorus have received considerable attention since the mid-1960s. Initially, nitrogen and phosphorus in wastewater

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discharges became important because of their effects in accelerating eutrophication of lakes and promoting aquatic growths. More recently, nutrient control has become a routine part of treating wastewaters used for the recharge of groundwater supplies. Nitrification of wastewater discharges is also required in many cases to reduce ammonia toxicity or to lessen the impact on the oxygen resources in flowing streams or estuaries. As a result of the many concerns over nutrients, nutrient removal has become, for all practical purposes, an integral part of conventional wastewater treatment. Thus, the removal of nitrogen and phosphorus by biological processes is considered in Chaps. 8 and 9. The removal of phosphorus by chemical methods is discussed in Chap. 6.

11–2 TECHNOLOGIES USED FOR ADVANCED TREATMENT

Over the past 20 years, a wide variety of treatment technologies have been studied, developed, and applied for the removal of the residual constituents found in treated effluent. The residual constituents found in treated wastewater are reviewed and the technologies used for the advanced treatment of wastewater are introduced in this section.

Residual Constituents in Treated Wastewater

The typical composition of domestic wastewater was reported previously in Table 3–15. In addition to the constituents reported in Table 3–15, most domestic wastewaters also contain a wide variety of trace compounds and elements, although they are not measured routinely. The potential effects of residual constituents contained in treated effluents may vary considerably. Some of the effects of specific constituents are listed in Table 11–1. The list presented in Table 11–1 is not meant to be exhaustive; rather, it is meant to highlight that a wide variety of substances must be considered in establishing and meeting discharge requirements.

Classification of Technologies

Advanced wastewater-treatment systems may be classified by the type of unit operation or process or by the principal removal function performed. To facilitate a general comparison of the various operations and processes, information on (1) the principal residual constituent removal function and (2) the types of operations or processes that can be used to perform this function are presented in Table 11–2. As will be noted in reviewing Table 11–2, many of the operations and processes can be used for the removal of a number of constituents. The individual constituents listed in Table 11–2 can be grouped into four broad categories requiring removal: (1) residual organic and inorganic colloidal and suspended solids, (2) dissolved organic constituents, (3) dissolved inorganic constituents, and (4) biological constituents. Typical advanced treatment process flow diagrams incorporating many of the technologies listed in Table 11–2 are illustrated on Fig. 11–1.

Removal of Organic and Inorganic Colloidal and Suspended Solids

Removal of organic and inorganic colloidal and suspended solids is typically accomplished by filtration. A general classification of the filtration processes commonly used in wastewater engineering is presented on Fig. 11–2. As illustrated on Fig. 11–2, the fil-

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Table 11-1
Typical residual constituents found in treated wastewater effluents and their impacts

Residual constituent	Effect
Inorganic and organic colloidal and suspended solids	
Suspended solids	• May cause sludge deposits or interfere with receiving water clarity
	• Can impact disinfection by shielding organisms
Colloidal solids	• May affect effluent turbidity
Organic matter (particulate)	• May shield bacteria during disinfection, may deplete oxygen resources
Dissolved organic matter	
Total organic carbon	• May deplete oxygen resources
Refractory organics	• Toxic to humans; carcinogenic
Volatile organic compounds	• Toxic to humans; carcinogenic; form photochemical oxidants
Pharmaceutical compounds	• Impact aquatic species (e.g., endocrine disruption, sex reversal)
Surfactants	• Cause foaming and may interfere with coagulation
Dissolved inorganic matter	
Ammonia	• Increases chlorine demand
	• Can be converted to nitrates and, in the process, can deplete oxygen resources
	• With phosphorus, can lead to the development of undesirable aquatic growths
	• Toxic to fish
Nitrate	• Stimulates algal and aquatic growth
	• Can cause methemoglobinemia in infants (blue babies)
Phosphorus	• Stimulates algal and aquatic growth
	• Interferes with coagulation
	• Interferes with lime-soda softening
Calcium and magnesium	• Increase hardness and total dissolved solids
Chloride	• Imparts salty taste
Total dissolved solids	• Interfere with agricultural and industrial processes
Biological	
Bacteria	• May cause disease
Protozoan cysts and oocysts	• May cause disease
Viruses	• May cause disease

ters used for wastewater fall into three general classifications: (1) depth filtration, (2) surface filtration, and (3) membrane filtration. In depth filtration, the removal of suspended material occurs within and on the surface of the filter bed (see Fig. 11-3*a*). In surface and membrane filtration, the suspended material is removed by straining through a straining surface (e.g., filter cloth) or a thin supported membrane (see Fig. 11-3*b*). An automobile oil filter and a kitchen colander are diverse examples of surface filters.

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Table 11-2
Unit operations and processes for the removal of residual constituents found in treated wastewater effluents[a]

Residual constituent	Unit operation or process (section discussed)				
	Depth filtration (11-4)	Surface filtration (11-5)	Micro- and ultra-filtration (11-6)	Reverse osmosis (11-6)	Electro-dialysis (11-6)
Inorganic and organic colloidal and suspended solids					
Suspended solids	✔	✔	✔	✔	✔
Colloidal solids	✔	✔	✔	✔	✔
Organic matter (particulate)				✔	✔
Dissolved organic matter					
Total organic carbon				✔	✔
Refractory organics				✔	✔
Volatile organic compounds				✔	✔
Dissolved inorganic matter					
Ammonia[a]				✔	✔
Nitrate[a]				✔	✔
Phosphorus[a]	✔[b]			✔	✔
Total dissolved solids				✔	✔
Biological					
Bacteria			✔	✔	✔
Protozoan cysts and oocysts	✔		✔	✔	✔
Viruses				✔	✔

[a] The biological removal of nitrogen and phosphorus is considered in Chaps. 7 through 10.

[b] Phosphorous removal is accomplished in a two-stage filtration process.

[c] Some carryover can occur.

Removal of Dissolved Organic Constituents

Many treatment methods can be used for the removal of dissolved organic constituents. Because of the complex nature of the dissolved organic constituents, the treatment methods must consider the specific characteristics of the wastewater and the nature of the constituents. Treatment processes used to remove some of the specific dissolved organic constituents include (1) carbon adsorption, (2) reverse osmosis, (3) chemical precipitation, (4) chemical oxidation, (5) advanced chemical oxidation, (6) electrodialysis, and (7) distillation. Chemical precipitation and chemical oxidation are discussed in Chap. 6. Biological treatment for nutrient removal is discussed in Chaps. 7 through 10.

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Table 11-2 (*Continued*)

Unit operation or process (section discussed)						
Adsorption (11-7)	**Air stripping (11-8)**	**Ion exchange (11-9)**	**Advanced oxidation processes (11-10)**	**Distillation (11-11)**	**Chemical precipitation (6-3, 4, 5)**	**Chemical oxidation (6-7)**
✔		✔		✔	✔	
✔		✔		✔	✔	
				✔		✔
✔		✔	✔	✔	✔	✔
✔			✔	✔		
✔	✔		✔	✔[c]		
	✔	✔		✔		
		✔		✔		
				✔	✔	
		✔		✔		
				✔		
✔		✔		✔	✔	
				✔		

Removal of Dissolved Inorganic Constituents

As reported in Table 11-2, a number of different unit operations and processes have been investigated in various advanced wastewater-treatment applications. Although many of them have proved to be technically feasible, other factors, such as cost, operational requirements, and aesthetic considerations, have not been favorable in some cases. Nevertheless, it is important that environmental engineers be familiar with the more important operations and processes so that in any given situation they can consider all treatment possibilities. Removal of dissolved inorganic constituents is accomplished by chemical processes or membrane filtration. The principal unit operations and processes are (1) chemical precipitation (discussed in Chap. 6), (2) ion exchange, (3) ultrafiltration, (4) reverse osmosis, (5) electrodialysis, and (6) distillation. As noted previously, chemical precipitation is discussed in Chap. 6.

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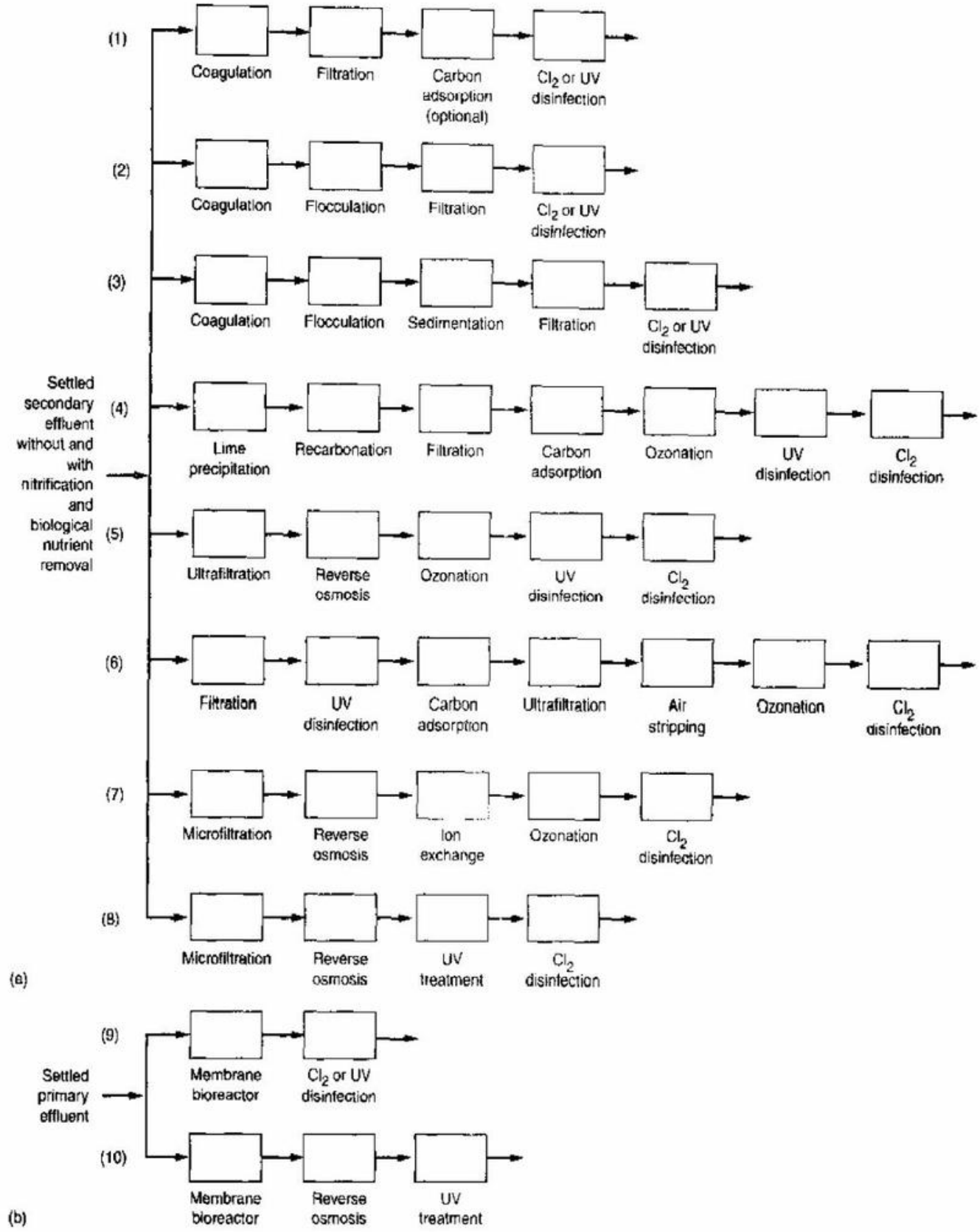

Figure 11–1

Typical process flow diagrams for wastewater treatment employing advanced treatment processes with (a) settled secondary effluent and (b) settled primary effluent. All of the flow diagrams have been used at one time or another. For example, in flow diagram 7, ion exchange is used for the removal of nitrate. In flow diagram 10, UV can be used for both disinfection and the destruction of NDMA.

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Figure 11-2

Classification of filtration processes used in wastewater management. Note: Intermittent and recirculating porous medium filters are used for small systems and are not considered in this text.

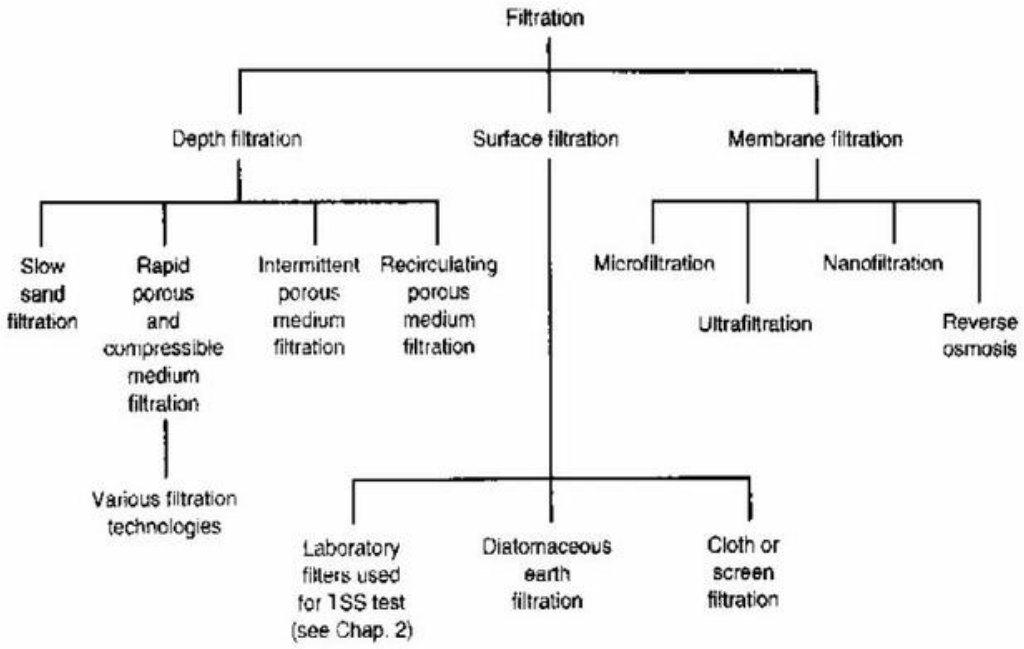
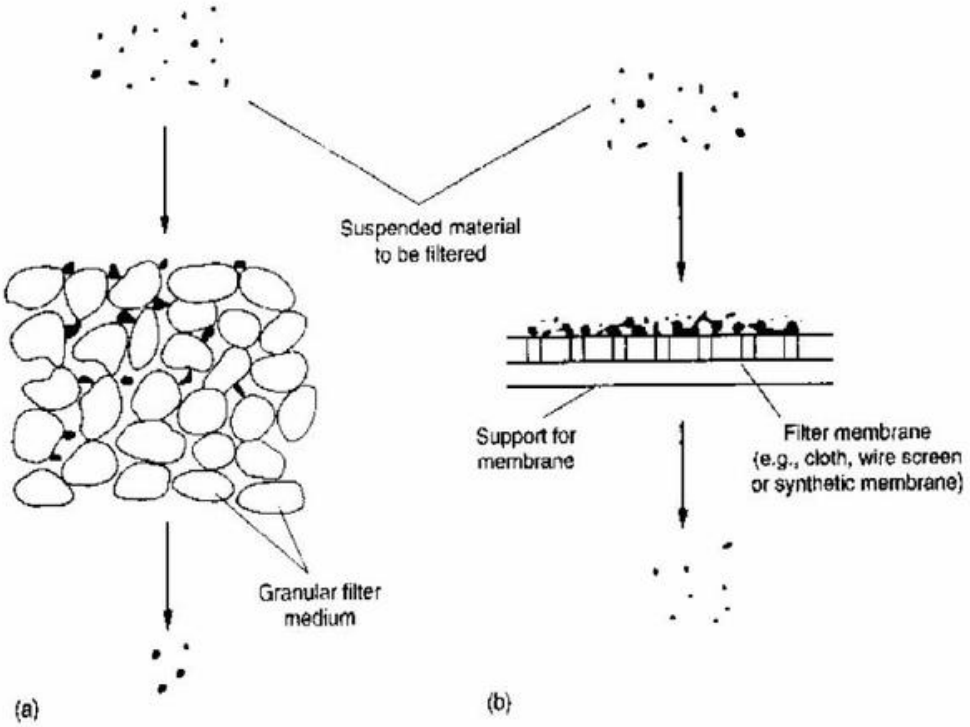


Figure 11-3

Definition sketch for filtration processes: (a) depth filtration and (b) surface filtration.



Removal of Biological Constituents

In addition to the constituents discussed above, the removal of biological constituents is also of interest. The unit operations and processes that are useful for the removal of biological constituents including bacteria, protozoan cysts and oocysts, and viruses are reported in Table 11-2. Because the effectiveness of the unit operations and processes listed in Table 11-2 is variable, disinfection of the treated effluent is required for most applications.

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Process Selection and Performance Data

Selection of a given operation, process, or combination thereof depends on (1) the use to be made of the treated effluent, (2) the nature of the wastewater, (3) the compatibility of the various operations and processes, (4) the available means to dispose of the ultimate contaminants, and (5) the environmental and economic feasibility of the various systems. Specific factors that should be considered in the selection of treatment processes may be found in Table 4–11 in Chap. 4. It should be noted that in some situations economic feasibility may not be a controlling factor in the design of advanced wastewater-treatment systems, especially where specific constituents must be removed to protect the environment. Based on the variations in performance observed in the field, pilot-plant testing is recommended for the development of treatment performance data and design criteria. Representative performance data are presented in the discussion of the individual technologies.

11–3 INTRODUCTION TO DEPTH FILTRATION

Depth filtration involves the removal of particulate material suspended in a liquid by passing the liquid through a filter bed comprised of a granular or compressible filter medium (see Fig. 11–3*a*). Although depth filtration is one of the principal unit operations used in the treatment of potable water, the filtration of effluents from wastewater-treatment processes is becoming more common. Depth filtration is now used to achieve supplemental removals of suspended solids (including particulate BOD) from wastewater effluents of biological and chemical treatment processes to reduce the mass discharge of solids and, perhaps more importantly, as a conditioning step that will allow for the effective disinfection of the filtered effluent. Depth filtration is also used as a pretreatment step for membrane filtration (see Sec. 11–6). Single- and two-stage filtration is also used to remove chemically precipitated phosphorus.

Historically, the first depth filtration process developed for the treatment of wastewater was the slow sand filter [typical filtration rates of 30 to 60 $L/m^2 \cdot d$ (0.75 to 1.5 $gal/ft^2 \cdot d$), Frankland (1870), and Dunbar (1908)]. The rapid sand filter (typical filtration rates of 80 to 200 $L/m^2 \cdot min$ (2.0 to 5.0 $gal/ft^2 \cdot min$), the subject of this section, was developed to treat larger volumes of water in a facility with a smaller footprint. To introduce the subject of depth filtration, the purpose of this section is to present (1) a general introduction to the depth filtration process, (2) an introduction to filter clean-water hydraulics, and (3) an analysis of the filtration process. The types of filters that are available and issues associated with their selection and design, including a discussion of the need for pilot-plant studies, are considered in the following section.

Description of the Filtration Process

Before discussing the available filter technologies, it will be useful to first describe the basics of the depth filtration including (1) the physical features of a conventional granular medium-depth filter, (2) filter-medium characteristics, (3) the *filtration process* in which suspended material is removed from the liquid, (4) the operative particle-removal mechanisms that bring about the removal of suspended material within the filter, and (5) the *backwash process*, in which the material that has been retained within the filter is removed.

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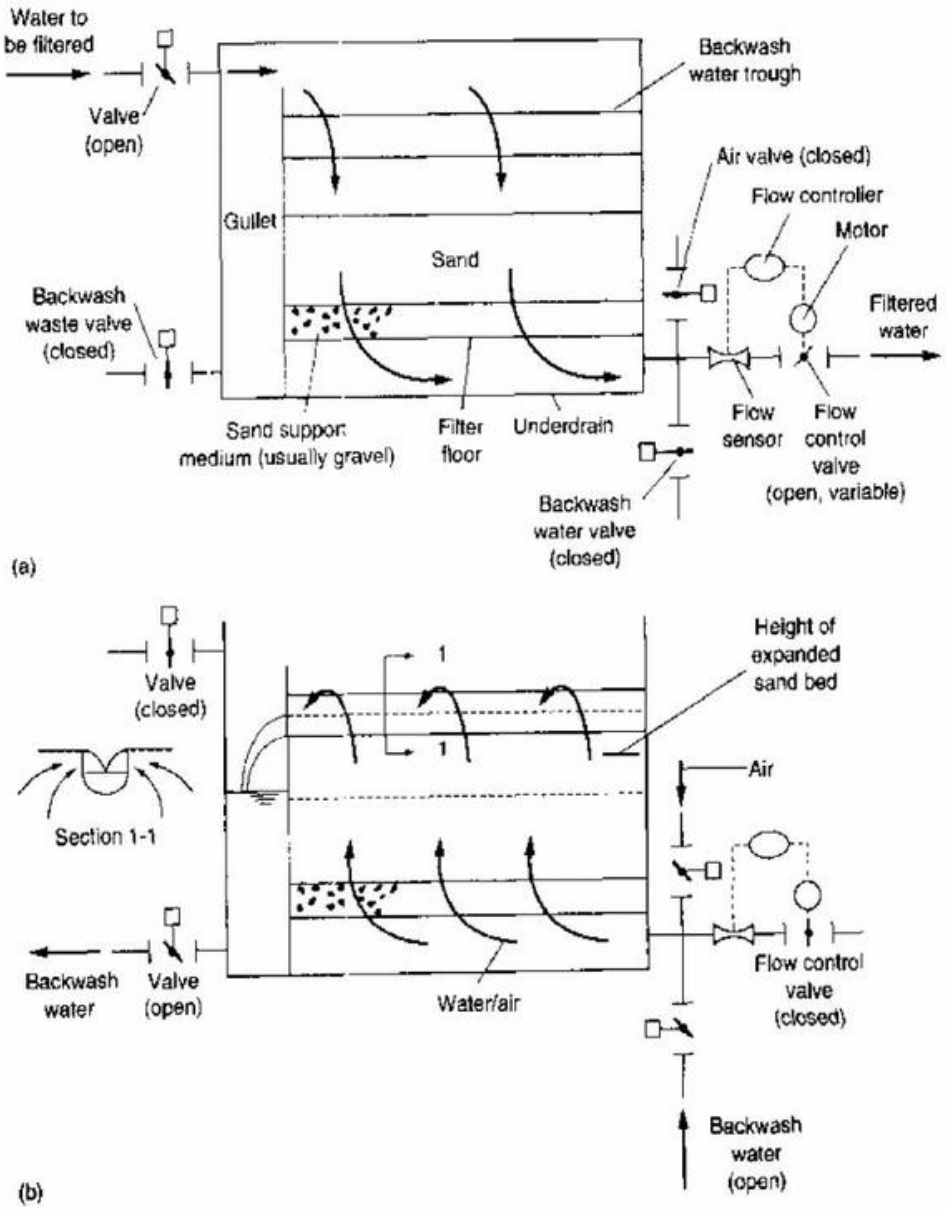
Physical Features of a Depth Filter. The general features of a conventional rapid granular medium-depth filter are illustrated on Fig. 11–4. As shown, the filtering medium (sand in this case) is supported on a gravel layer, which, in turn, rests on the filter underdrain system. The water to be filtered enters the filter from an inlet channel. Filtered water is collected in the underdrain system, which is also used to reverse the flow to backwash the filter. Filtered water typically is disinfected before being discharged to the environment. If the filtered water is to be reused, it can be discharged to a storage reservoir or to the reclaimed water distribution system. The hydraulic control of the filter is described in a subsequent section.

Filter-Medium Characteristics. Grain size is the principal filter-medium characteristic that affects the filtration operation. Grain size affects both the clear-water headloss and the buildup of headloss during the filter run. If too small a filtering medium is

Figure 11–4

General features and operation of a conventional rapid granular medium-depth filter: (a) flow during filtration cycle, and (b) flow during backwash cycle. *(From Tchobanoglous and Schroeder, 1985.)*



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selected, much of the driving force will be wasted in overcoming the frictional resistance of the filter bed. On the other hand, if the size of the medium is too large, many of the small particles in the influent will pass directly through the bed. The size distribution of the filter material is usually determined by sieve analysis using a series of decreasing sieve sizes. The designation and size of opening for U.S. sieve sizes are given in Table 11–3. The results of a sieve analysis are usually analyzed by plotting the cumulative percent passing a given sieve size on arithmetic-log or probability-log paper (see Example 11–1 and Fig. 11–23).

The effective size of a filtering medium is defined as the 10 percent size based on mass and is designated as d_{10}. For sand, it has been found that the 10 percent size by weight corresponds to the 50 percent size by count. The uniformity coefficient (UC) is defined as the ratio of the 60 percent size to the 10 percent size ($UC = d_{60}/d_{10}$). Sometimes it is advantageous to specify the 99 percent passing size and the 1 percent passing size to define the gradation curve for each filter medium more accurately. Addi-

Table 11–3 Designation and size of opening of U.S. sieve sizes

Sieve size or number	Size of opening in	Size of opening mm
3/8 in	0.375[a]	9.51[a]
1/4 in	0.250[a]	6.35[a]
4	0.187	4.76
6	0.132	3.36
8	0.0937	2.38
10	0.0787[a]	2.00[a]
12	0.0661	1.68
14	0.0555[a]	1.41[a]
16	0.0469	1.19
18	0.0394[a]	1.00[a]
20	0.0331	0.841
25	0.0280[a]	0.710[a]
30	0.0234	0.595
35	0.0197[a]	0.500[a]
40	0.0165	0.420
45	0.0138[a]	0.350[a]
50	0.0117	0.297
60	0.0098[a]	0.250[a]
70	0.0083	0.210
80	0.0070[a]	0.177[a]
100	0.0059	0.149

[a] Size does not follow the ratio $(2)^{0.5}$.

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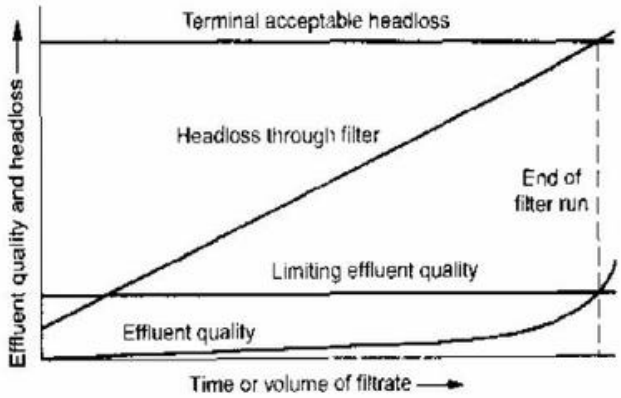

Figure 11-5

Definition sketch for length of filter run based on headloss and effluent turbidity.

tional information on filter-medium characteristics is presented in the following section dealing with the design of filters.

The Filtration Process. During filtration in a conventional downflow depth filter, wastewater containing suspended matter is applied to the top of the filter bed (Fig. 11–4*a*). As the water passes through the filter bed, the suspended matter in the wastewater is removed by a variety of removal mechanisms as described below. With the passage of time, as material accumulates within the interstices of the granular medium, the headloss through the filter starts to build up beyond the initial value, as shown on Fig. 11–5. After some period of time, the operating headloss or effluent turbidity reaches a predetermined headloss or turbidity value, and the filter must be cleaned. Under ideal conditions, the time required for the headloss buildup to reach the preselected terminal value should correspond to the time when the suspended solids in the effluent reach the preselected terminal value for acceptable quality. In actual practice, one or the other event will govern the backwash cycle.

Particle Removal Mechanisms. The principal particle removal mechanisms, believed to contribute to the removal of material within a granular-medium filter, are identified and described in Table 11–4. The major removal mechanisms (the first six listed in Table 11–4) are illustrated pictorially on Fig. 11–6. Straining has been identified as the principal mechanism that is operative in the removal of suspended solids during the filtration of settled secondary effluent from biological treatment processes (Tchobanoglous and Eliassen, 1970). Other mechanisms including impaction, interception, and adhesion are also operative even though their effects are small and, for the most part, masked by the straining action.

The removal of the smaller particles found in wastewater (see Fig. 11–6) must be accomplished in two steps involving (1) the transport of the particles to or near the surface where they will be removed and (2) the removal of particles by one or more of the operative removal mechanisms. This two-step process has been identified as transport and attachment (O'Melia and Stumm, 1967).

Backwash Process. The end of the filter run (filtration phase) is reached when the suspended solids in the effluent start to increase (break through) beyond an acceptable level, or when a limiting headloss occurs across the filter bed (see Fig. 11–5). Once

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Table 11–4 Principal mechanisms and phenomena contributing to removal of material within a granular medium-depth filter

Mechanism/ phenomenon	Description
1. Straining	
a. Mechanical	Particles larger than the pore space of the filtering medium are strained out mechanically
b. Chance contact	Particles smaller than the pore space are trapped within the filter by chance contact
2. Sedimentation	Particles settle on the filtering medium within the filter
3. Impaction	Heavy particles will not follow the flow streamlines
4. Interception	Many particles that move along in the streamline are removed when they come in contact with the surface of the filtering medium
5. Adhesion	Particles become attached to the surface of the filtering medium as they pass by. Because of the force of the flowing water, some material is sheared away before it becomes firmly attached and is pushed deeper into the filter bed. As the bed becomes clogged, the surface shear force increases to a point at which no additional material can be removed. Some material may break through the bottom of the filter, causing the sudden appearance of turbidity in the effluent
6. Flocculation	Flocculation can occur within the interstices of the filter medium. The larger particles formed by the velocity gradients within the filter are then removed by one or more of the above removal mechanisms
7. Chemical adsorption a. Bonding b. Chemical interaction 8. Physical adsorption a. Electrostatic forces b. Electrokinetic forces c. van der Waals forces	Once a particle has been brought in contact with the surface of the filtering medium or with other particles, either one of these mechanisms, chemical or physical adsorption or both, may be responsible for holding it there
9. Biological growth	Biological growth within the filter will reduce the pore volume and may enhance the removal of particles with any of the above removal mechanisms (1 through 5)

either of these conditions is reached, the filtration phase is terminated, and the filter must be cleaned (backwashed) to remove the material (suspended solids) that has accumulated within the granular filter bed. Backwashing is accomplished by reversing the flow through the filter (see Fig. 11–4*b*). A sufficient flow of washwater is applied until the granular filtering medium is fluidized (expanded), causing the particles of the filtering medium to abrade against each other.

The suspended matter arrested within the filter is removed by the shear forces created by backwash water as it moves up through the expanded bed. The material that has accumulated within the bed is then washed away. Surface washing with water and air

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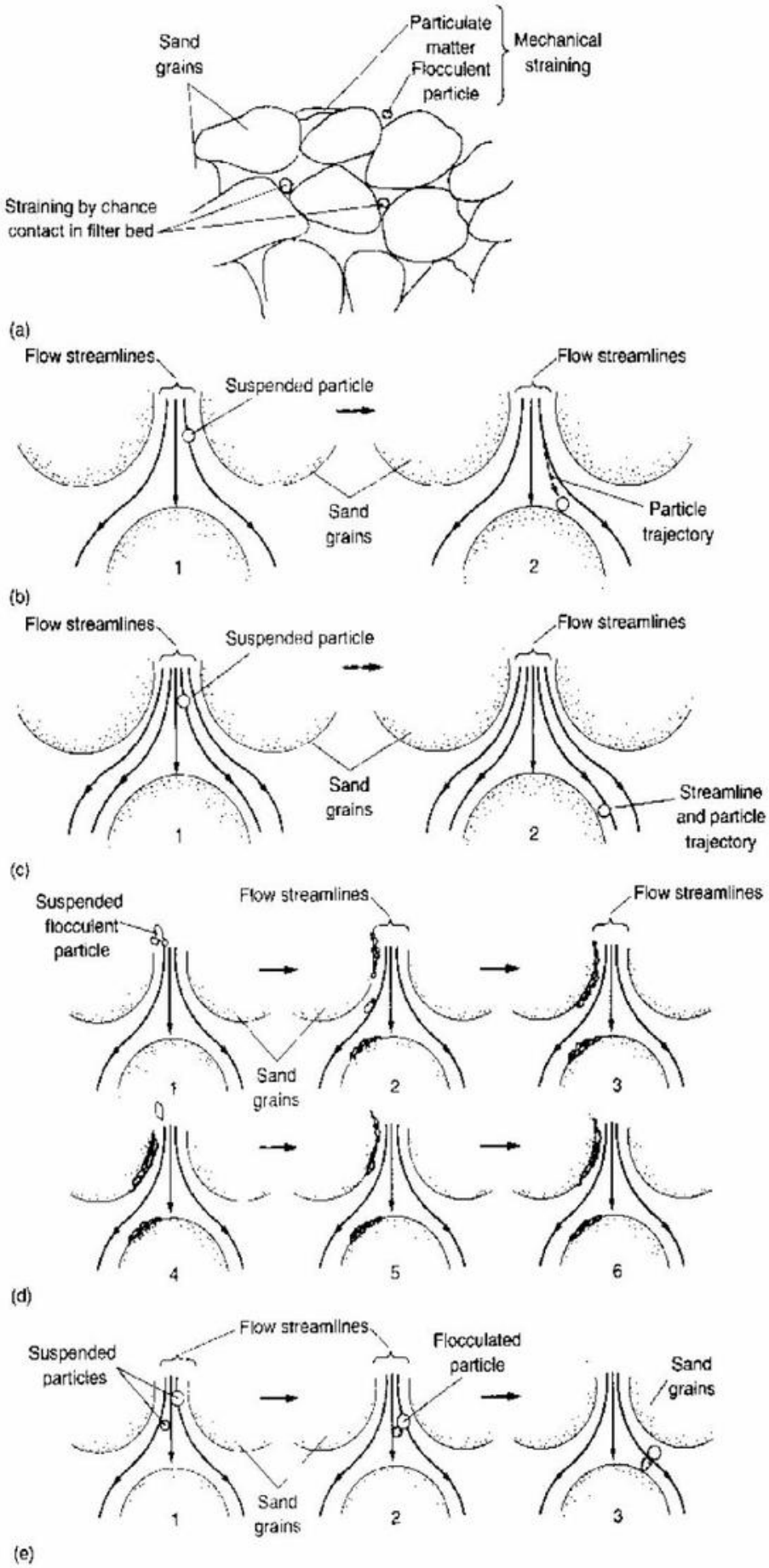
Figure 11-6

Removal of suspended particulate matter within a granular filter (a) by straining, (b) by sedimentation or inertial impaction, (c) by interception, (d) by adhesion, (e) by flocculation. (*Adapted from Tchobanoglous and Schroeder, 1985.*)



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scouring are often used in conjunction with the water backwash to enhance the cleaning of the filter bed. In most wastewater-treatment plant flow diagrams, the washwater containing the suspended solids that are removed from the filter is returned either to the primary settling facilities or to the biological treatment process. Backwash hydraulics are considered in the section dealing with filter hydraulics.

Filter Hydraulics

During the past 60 years considerable effort has been devoted to the modeling of the filtration process. The models fall into two general categories: those models used to predict the clean-water headloss through a filter bed and the filter backwash expansion and those models used to predict the performance of filters for the removal of suspended solids. Filter hydraulics are considered in the following discussion.

Clean-Water Headloss. Over the years several equations have been developed to describe the flow of clean water through a porous medium (Carman, 1937; Fair and Hatch, 1933; Hazen, 1905; Kozeny, 1927; Rose, 1945); these equations are summarized in Table 11–5. In most cases, the equations for the flow of clean water through a porous medium are derived from a consideration of the Darcy-Weisbach equation for flow in a closed conduit and dimensional analysis. The summation term in Eqs. (11–2), (11–6), and (11–8) is included to account for the stratification that occurs in filters. To account for stratification, the mean size of the material retained between successive sieve sizes is assumed to correspond to the mean size of the successive sieves (see Table 11–9 in the following section for sieve sizes), assuming that the particles retained between sieve sizes are substantially uniform (Fair and Hatch, 1933).

In applying the equations given in Table 11–5, some confusion exists over the definition of the shape factor. The definition of the shape factor will depend on whether particle surface, volume, or a linear dimension is of importance in the application (Trussell and Chang, 1999). In Chap. 5, the shape factor for a particle ϕ_p is defined as the ratio of the surface area of an equivalent sphere to the surface area of the particle, for particles of the same volume. For spherical-shaped particles the specific surface area is

$$\frac{A_p}{v_p} = \frac{\pi d^2}{(\pi d^3/6)} = \frac{6}{d} \tag{11–11}$$

where A_p = area of filter-medium particle, m^2, mm^2
v_p = volume of filter-medium particle, m^3, mm^3
d – diameter of filter-medium particle, m, mm

For irregularly shaped particles of the same volume the specific surface area is:

$$\frac{A_p}{v_p} = \frac{6}{\phi d} = \frac{S}{d} \tag{11–12}$$

where ϕ = approximate shape factor
S = approximate area to volume shape factor

In the literature S has been identified as a shape factor (Fair et al., 1968). The area to volume shape factor is used in the Fair-Hatch equation [(Eq. 11–5)]. Computation of the clean-water headloss through a filter is illustrated in Example 11–1.

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Table 11–5
Formulas used to compute the clean-water headloss through a granular porous medium

Equation	No.
Carman-Kozeny (Carman, 1937)	
$h = \frac{f}{\phi}\frac{1-\alpha}{\alpha^3}\frac{L}{d}\frac{v_s^2}{g}$	(11–1)
$h = \frac{1}{\phi}\frac{1-\alpha}{\alpha^3}\frac{Lv_s^2}{g}\sum f\frac{p}{d_g}$	(11–2)
$f = 150\frac{1-\alpha}{N_R} + 1.75$	(11–3)
$N_R = \frac{\phi d v_s \rho}{\mu}$	(11–4)
Fair-Hatch (Fair and Hatch, 1933)	
$h = k\nu S^2\frac{(1-\alpha)^2}{\alpha^3}\frac{L}{d^2}\frac{v_s}{g}$	(11–5)
$h = k\nu\frac{(1-\alpha)^2}{\alpha^3}\frac{Lv_s}{g}\left(\frac{6}{\phi}\right)^2\sum\frac{p}{d_g^2}$	(11–6)
Rose (Rose, 1945)	
$h = \frac{1.067}{\phi}C_d\frac{1}{\alpha^4}\frac{L}{d}\frac{v_s^2}{g}$	(11–7)
$h = \frac{1.067}{\phi}\frac{Lv_s^2}{\alpha^4 g}\sum C_d\frac{p}{d_g}$	(11–8)
$C_d = \frac{24}{N_R} + \frac{3}{\sqrt{N_R}} + 0.34$	(11–9)
Hazen (Hazen, 1905)	
$h = \frac{1}{C}\frac{60}{T+10}\frac{L}{(d_{10})^2}v_h$	(11–10)

Definition of terms

C = coefficient of compactness (varies from 600 for very closely packed sands that are not quite clean to 1200 for very uniform clean sand)
C_d = coefficient of drag
d = grain size diameter, m (ft)
d_g = geometric mean diameter between sieve sizes d_1 and d_2, $\sqrt{d_1 d_2}$, mm (in)
d_{10} = effective grain size diameter, mm
f = friction factor
g = acceleration due to gravity, 9.81 m/s² (32.2 ft/s²)
h = headloss, m (ft)
k = filtration constant, 5 based on sieve openings, 6 based on size of separation
L = depth of filter bed or layer, m (ft)
N_R = Reynolds number
p = fraction of particles (based on mass) within adjacent sieve sizes
S = shape factor (varies between 6.0 for spherical particles and 8.5 for crushed materials)
T = temperature, °C (°F)
v_h = superficial (approach) filtration velocity, m/d (ft/d)
v_s = superficial (approach) filtration velocity, m/s (ft/d)
α = porosity
μ = viscosity, N·s/m² (lb·s/ft²)
ν = kinematic viscosity, m²/s (ft²/s)
ρ = density = kg/m³ (lb·s²/ft⁴)
ϕ = particle shape factor (1.0 for spheres, 0.82 for rounded sand, 0.75 for average sand, 0.73 for crushed coal and angular sand)

EXAMPLE 11–1 Determination of Clean-Water Headloss in a Granular-Medium Filter Determine the effective size, the uniformity coefficient, and the clean-water headloss in a filter bed composed of 0.75 m of uniform sand with the size distribution given below for a filtration rate of 160 L/m²·min. Assume that the operating temperature is 20°C. Use the Rose equation [Eq. (11–8)] given in Table 11–5 for computing the headloss. Assume the porosity of the sand in the various layers is 0.40 and use a value of 0.85 for the shape factor for sand.

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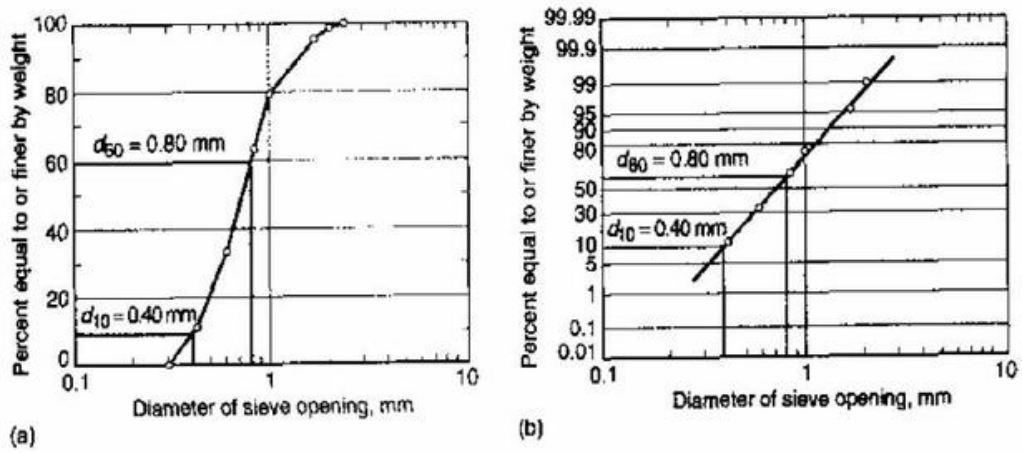
Sieve size or number	Percent of sand retained	Cumulative percent passing	Geometric mean size[a], mm
6–8	0	100	
8–10	1	99	2.18
10–12	3	96	1.83
12–18	16	80	1.30
18–20	16	64	0.92
20–30	30	34	0.71
30–40	22	12	0.50
40–50	12		0.35

[a]Using sieve size data from Table 11–3, the geometric mean size $d_g = \sqrt{d_1 d_2}$.

Solution

1. Determine the effective size and the uniformity coefficient of the sand. Plot the cumulative percent passing versus the corresponding sieve size. Two different methods of plotting the data are presented below.



a. The effective size d_{10} read from the graphs is 0.40 mm.

b. The uniformity coefficient is

$$\text{UC} = \frac{d_{60}}{d_{10}} = \frac{0.80\ \text{mm}}{0.40\ \text{mm}} = 2.0$$

2. Determine the clean-water headloss using Eq. (11–8).

$$h = \frac{1.067}{\phi}\,\frac{Lv_s^2}{\alpha^4 g}\,\Sigma C_d \frac{p}{d_g}$$

a. Set up computation table to determine the summation term in Eq. (11–8).

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Sieve size or number	Fraction of sand retained	Geometric mean size, $d_g \times 10^3$ m	Reynolds number, N_R	C_d	$C_d(p/d)$, m^{-1}
8–10	0.01	2.18	4.93	6.56	30
10–12	0.03	1.83	4.15	7.60	124
12–18	0.16	1.30	2.93	10.28	1,268
18–20	0.16	0.92	2.08	13.99	2,441
20–30	0.30	0.71	1.60	17.71	7,509
30–40	0.22	0.50	1.13	24.38	10,729
40–50	0.12	0.35	0.80	33.73	11,459
Sum					33,560

i. Determine the C_d value for each geometric mean as illustrated below.

$$C_d = \frac{24}{N_R} + \frac{3}{\sqrt{N_R}} + 0.34$$

$$N_R = \frac{\phi d v_s \rho}{\mu} = \frac{\phi d v_s}{\nu}$$

$$d = 2.18 \text{ mm}$$

$$v_s = \frac{(60 \text{ L/m}^2\cdot\text{min})}{(10^3 \text{ L/m}^3)(60 \text{ s/min})} = 0.00267 \text{ m/s}$$

$$\nu = 1.003 \times 10^{-6} \text{ m}^2/\text{s (see Appendix C)}$$

$$N_R = \frac{(0.85)(0.00218 \text{ m})(0.00267 \text{ m/s})}{(1.003 \times 10^{-6} \text{ m}^2/\text{s})}$$

$$N_R = 4.93$$

ii. Determine C_d.

$$C_d = \frac{24}{4.93} + \frac{3}{\sqrt{4.93}} + 0.34 = 6.56$$

b. Determine the headloss through the stratified filter bed using Eq. (11–8).

$$L = 0.75 \text{ m}$$

$$v_s = 0.00267 \text{ m/s}$$

$$\phi = 0.85$$

$$\alpha = 0.40$$

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$$g = 9.81 \text{ m/s}^2$$

$$h = \frac{1.067}{(0.85)} \frac{(0.75 \text{ m})(0.00267 \text{ m/s})^2}{(0.40)^4(9.81 \text{ m/s}^2)} (33{,}500/\text{m})$$

$$h = 0.90 \text{ m}$$

Backwash Hydraulics. To understand what happens during the backwash operation it will be helpful to refer to Fig. 11–7 in which the pressure drop across a packed bed is illustrated as the backwash velocity through it increases. Between points *A* and *B*, the bed is stable, and the pressure drop and Reynolds number N_R are related linearly. At point *B*, the pressure drop essentially balances the weight of the filter. Between points *B* and *C* the filter bed is unstable, and the particles comprising the filter medium adjust their position to present as little resistance to flow as possible. At point *C*, the loosest possible arrangement is obtained in which the particles are still in contact. Beyond point *C*, the particles begin to move freely but collide frequently so that the motion is similar to that of particles in hindered settling. Point *C* is referred to as the "point of fluidization." By the time point *D* is reached, the particles are all in motion, and, beyond this point, increases in N_R result in very small increases in ΔP as the bed continues to expand and the particles move in more rapid and more independent motion. Ultimately, the particles will stream with the fluid, and the bed will cease to exist at point *E*.

To expand a filter bed comprised of a uniform filter medium hydraulically, the headloss must equal the buoyant mass of the granular medium in the fluid. Mathematically this relationship can be expressed as

$$h = L_e(1 - \alpha_e)\left(\frac{\rho_s - \rho_w}{\rho_w}\right) \tag{11-13}$$

where h = headloss required to expand the bed
L_e = depth of the expanded bed
α_e = expanded porosity

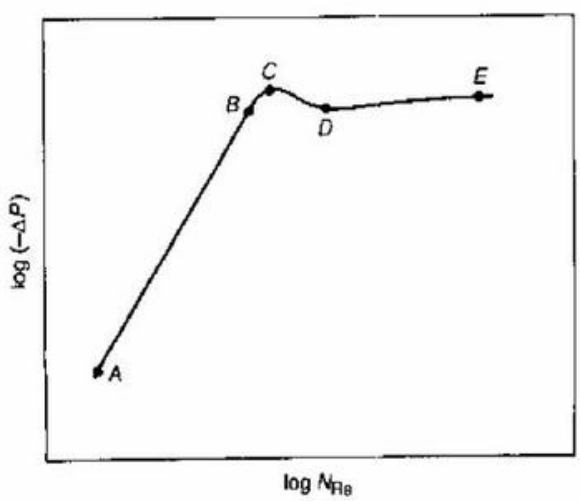

Figure 11–7

Schematic diagram illustrating the fluidization of a filter bed. *(Adapted from Foust et al. 1960.)*

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ρ_s = density of the medium
ρ_w = density of water

Because the individual particles are kept in suspension by the drag force exerted by the rising fluid, it can be shown from settling theory (see Sec. 5–5) that

$$C_D A_p \rho_w \frac{v^2}{2} \phi(\alpha_e) = (\rho_s - \rho_w) g v_p \tag{11-14}$$

where v = face velocity of backwash water, m/s
$\phi(\alpha_e)$ = correction factor to account for the fact that v is the velocity of the backwash water and not the particle-settling velocity v_p

and other terms are as defined previously.

From experimental studies (Fair, 1951; Richardson and Zaki, 1954) it has been found that the expanded-bed porosity can be approximated using the following relationships, assuming the Reynolds number is approximately 1.

$$\phi(\alpha_e) = \left(\frac{v_s}{v}\right)^2 = \left(\frac{1}{\alpha_e}\right)^9 \tag{11-15}$$

Thus

$$\alpha_e = \left(\frac{v}{v_s}\right)^{0.22} \tag{11-16}$$

or

$$v = v_s \alpha_e^{4.5} \tag{11-17}$$

where v_s = settling velocity of particle, m/s

However, because the volume of the filtering medium per unit area remains constant, $(1 - \alpha)L$ must be equal to $(1 - \alpha_e)L_e$ so that

$$\frac{L_e}{L} = \frac{1 - \alpha}{1 - \alpha_e} = \frac{1 - \alpha}{1 - (v/v_s)^{0.22}} \tag{11-18}$$

Where the filter medium is stratified, the smaller particles in the upper layers expand first. To expand the entire bed, the backwash velocity must be sufficient to lift the largest particle. To account for filter-bed stratification, Eq. (11–18) is modified assuming that particles retained between sieve sizes are substantially uniform (Fair and Hatch, 1933).

$$\frac{L_e}{L} = (1 - \alpha)\Sigma \frac{p}{(1 - \alpha_e)} \tag{11-19}$$

where p = fraction of filter medium retained between sieve sizes.

Thus, the required backwash velocity and expanded depth can be estimated using Eqs. (11–18) and (11–19), respectively, as illustrated in Example 11–2. Additional details on filter-bed expansion may be found in Amirtharajah (1978), Cleasby and Fan (1982), and Dharmarajah and Cleasby (1986); Kawamura (2000), Leva (1959); and Richardson and Zaki (1954).

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EXAMPLE 11-2 Determination of Required Backwash Velocities for Filter Cleaning A stratified sand bed with the size distribution given below is to be backwashed at a rate of 0.75 m/min. Determine the degree of expansion and whether the proposed backwash rate will expand all of the bed. Assume the following data are applicable:

Sieve size or number	Percent of sand retained	Geometric mean size[a], mm
6–8	0	2.83
8–10	1	2.18[b]
10–12	3	1.83
12–18	16	1.30
18–20	16	0.92
20–30	30	0.71
30–40	22	0.50
40–50	12	0.35

[a]Based on sieve sizes given in Table 11-3.
[b]2.18 $\sqrt{2.38 \times 2.0}$

1. Granular medium = sand
2. Specific gravity of sand = 2.65
3. Depth of filter bed = 0.90 m
4. Temperature = 20°C

Solution

1. Set up computation table to determine the summation term in Eq. (11-19).

$$\frac{L_e}{L} = (1 - \alpha)\Sigma \frac{p}{(1 - \alpha_e)}$$

Sieve size or number	Fraction of sand retained	Geometric mean size d_g, mm	v_s, m/s	v/v_s	α_e	$p/(1 - \alpha_e)$
8–10	0.01	2.18	0.304	0.041	0.496	0.020
10–12	0.03	1.83	0.270	0.046	0.509	0.061
12–18	0.16	1.30	0.210	0.060	0.538	0.346
18–20	0.16	0.92	0.157	0.080	0.573	0.351
20–30	0.30	0.71	0.123	0.102	0.605	0.760
30–40	0.22	0.50	0.085	0.146	0.655	0.638
40–50	0.12	0.35	0.055	0.227	0.722	0.432
Summation						2.632

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a. Determine the particle settling velocity as illustrated in Example 5–6. Alternatively, the settling velocities can be estimated using Fig. 5–21 in Chap. 5. The computed settling velocity values are entered in the computation table.
b. Determine the values of v/v_s and enter the computed values in the computation table. The backwash velocity is $v = 0.75$ m/min $= 0.0125$ m/s.
c. Determine the values of α_e and enter the computed values in the computation table.

$$\alpha_e = \left(\frac{v}{v_s}\right)^{0.22} = \left(\frac{0.0125}{0.304}\right)^{0.22} = 0.496$$

d. Determine the values for column 7 and enter the computed values in the computation table.

$$\frac{p}{1-\alpha_e} = \frac{1}{1-0.496} = 1.98$$

2. Determine the expanded-bed depth using Eq. (11–19).

$$\frac{L_e}{L} = (1-\alpha)\Sigma\frac{p}{(1-\alpha_e)}$$

$$L_e = (0.9\text{ m})(1-0.4)(2.63) = 1.42\text{ m}$$

3. Because the expanded porosity of the largest size fraction (0.496) is greater than the normal porosity of the filter material, the entire filter bed will be expanded.

Comment The expanded depth needs to be known to establish the minimum height of the wash-water troughs above the surface of the filter bed. In practice, the bottom of the backwash-water troughs is set from 50 to 150 mm (2 to 6 in) above the expanded filter bed. The width and depth of the troughs should be sufficient to handle the volume of backwash water used to clean the bed, with a minimum freeboard of 600 mm (24 in) at the upper end of the trough.

Analysis of the Filtration Process

In general, the mathematical characterization of the time-space removal of particulate matter within the filter is based on a consideration of the equation of continuity, together with an auxiliary rate equation.

Equation of Continuity. The equation of continuity for the filtration operation may be developed by considering a suspended-solids mass balance for a section of filter of cross-sectional area A and of thickness Δz, measured in the direction of flow (see Fig. 11–8). Following the approach delineated in Chap. 4, the mass balance would be as follows:

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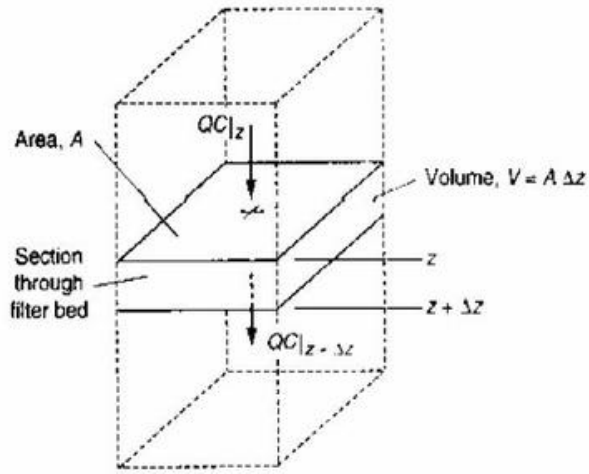
Figure 11–8

Definition sketch for the analysis of the filtration process. *(From Tchobanoglous and Schroeder, 1985.)*



1. General word statement:

$$\text{Rate of accumulation of solids within the volume element} = \text{rate of flow into the volume element} - \text{rate of flow out of the volume element} \tag{11–20}$$

2. Simplified word statement:

Accumulation = inflow − outflow

3. Symbolic representation:

$$\left(\frac{\partial q}{\partial t} + \alpha(t)\frac{\partial \bar{C}}{\partial t}\right)\Delta V = QC + Q\left(C + \frac{\Delta C}{\Delta z}\Delta z\right) \tag{11–21}$$

where $\partial q/\partial t$ = change in quantity of solids deposited within the filter time, g/m^3·min
$\alpha(t)$ = average porosity as a function of time
$\partial \bar{C}/\partial t$ = change in average concentration of solids in pore space with time, g/m^3·min
ΔV = differential volume, m^3
Q = volumetric flowrate, L/min (the unit L/min is used for convenience)
C = concentration of suspended solids, g/m^3
$\Delta C/\Delta z$ = change in concentration of suspended solids in fluid stream with distance, g/m^3·m

Substituting $A\,\Delta z$ for ΔV and Av for Q where v is the filtration velocity, simplifying, and taking the limit as Δz approaches zero, Eq. (11–21) becomes

$$-v\frac{\partial C}{\partial z} = \frac{\partial q}{\partial t} + \alpha(t)\frac{\partial \bar{C}}{\partial t} \tag{11–22}$$

In Eq. (11–22), the first term represents the difference between the mass of suspended solids entering and leaving the section; the second term represents the time rate of change in the mass of suspended solids accumulated within the interstices of the filter medium; and the third term represents the time rate of change in the suspended-solids concentration in the pore space within the filter volume.

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In a flowing process, the quantity of fluid contained within the bed is usually small compared with the volume of liquid passing through the bed. In this case, the materials balance equation can be written as

$$-v\frac{\partial C}{\partial z} = \frac{\partial q}{\partial t} \tag{11-23}$$

Equation (11–23) is the one most commonly found in the literature dealing with filtration theory.

If the shape of the removal curve within the filter does not vary with time, the equation of continuity [Eq. (11–23)] may be written as an ordinary differential equation:

$$-v\frac{dC}{dz} = \frac{dq}{dt} \tag{11-24}$$

Rate Equation. To solve Eq. (11–24), an additional independent equation is required. The most direct approach is to derive a relationship that can be used to describe the change in concentration of suspended matter with distance, such as

$$\frac{\partial C}{\partial z} = \phi(V_1, V_2, V_3, \ldots) \tag{11-25}$$

in which V_1, V_2, and V_3 are the variables governing the removal of suspended matter from solution.

An alternative approach is to develop a complementary equation in which the pertinent process variables are related to the amount of material retained (accumulated) within the filter at various depths. In equation form, this may be written as

$$\frac{\partial q}{\partial t} = \phi(V_1, V_2, V_3, \ldots) \tag{11-26}$$

As noted previously, the principal removal mechanism for the residual suspended solids found in settled secondary effluent from biological treatment processes is straining. The removal of suspended material by straining can be identified by noting (1) the variation in the normalized concentration-removal curves through the filter as a function of time, and (2) the shape of the headloss curve for the entire filter or an individual layer within the filter. If straining is the principal removal mechanism, the shape of the normalized removal curve will not vary significantly with time (see Fig. 11–9*a*), and the headloss curves will be curvilinear (Tchobanoglous and Eliassen, 1970). If the floc particles that are being filtered are easily broken apart, the normalized removal curves will vary with time as shown on Fig. 11–9*b*.

From the size and distribution of the influent particles (see Fig. 2–9, Chap. 2) and the shape of the normalized curves (see Fig. 11–9*a*), it can be concluded that the rate of change of concentration with distance must be proportional to some removal coefficient that is changing with the degree of treatment or removal achieved in the filter. For example, the entire particle-size distribution in the influent is passed through the first layer. The probability of removing particles from the waste stream is p_1. In the second layer, the probability of removing particles is p_2; p_2 is less than p_1, assuming that some of the larger particles will be removed by the first layer. Continuing this argument, it

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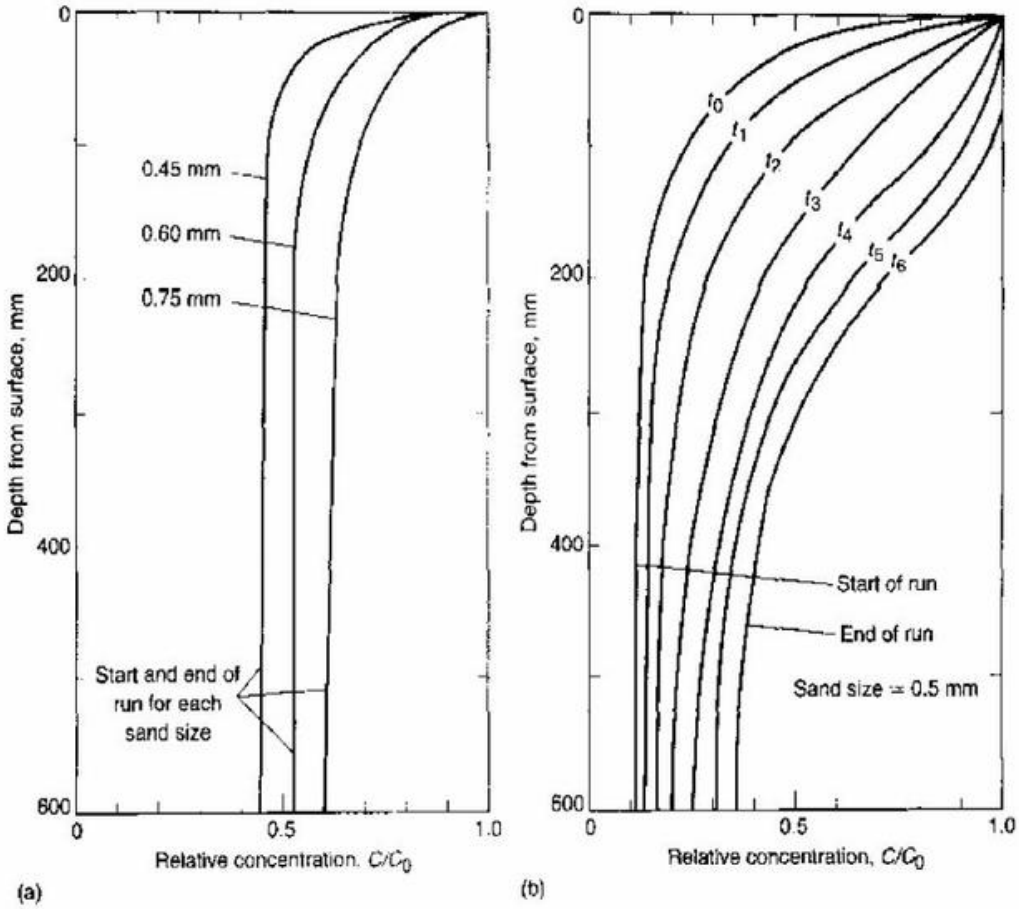
Figure 11-9

Normalized suspended-solids removal curves as a function of depth and time as observed during (a) the filtration of settled effluent from an activated-sludge treatment process and (b) the filtration of pretreated wastewater (coagulation/flocculation/sedimentation) with a weak floc. When the removal curves do not vary significantly with time as shown on Fig. 11-9a, straining is the predominant removal mechanism.



can be reasoned that the rate of removal must always be changing as a function of the degree of treatment. This phenomenon can be expressed mathematically using the following equation:

$$\frac{dC}{dz} = \left[\frac{1}{(1 + az)^n}\right] r_o C \qquad (11\text{–}27)$$

where C = concentration, mg/L
z = distance from top of filter bed, cm
r_o = initial removal rate, cm^{-1}
a, n = constants

In Eq. (11–27), the term within brackets is a retardation factor as described in Chap. 4. When the exponent n is equal to zero, the term within the brackets is equal to 1; under these conditions, Eq. (11–27) represents a logarithmic curve. When n equals 1, the value of the term within brackets drops off rapidly in the first 125 mm (5 in) and then more gradually as a function of distance. Therefore, it appears that the exponent n may be related to the distribution of particle sizes in the influent. For example, when dealing with a uniform filter medium and filtering particles of one size, it would be expected that the value of the exponent n would be equal to zero and that the initial removal could be described as a first-order removal function. It should be noted that this equation was verified only for filtration rates up to 400 $L/m^2 \cdot min$ (10 $gal/ft^2 \cdot min$).

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The value of r_o, the initial removal rate constant, is obtained by computing the slope of the removal curve at near zero depth, because $[1/(1 + az)^n] \approx 1$. The easiest way to determine the slope is to rewrite Eq. (11–27) as follows:

$$\left(\frac{Cr_o}{dC/dz}\right)^{1/n} = 1 + az \tag{11–28}$$

If Eq. (11–28) is plotted functionally, the value of n is equal to the value that results in a straight-line plot. The slope of the line describing the experimental data will be equal to the constant a.

Generalized Rate Equation. On the basis of experimental results derived from pilot testing and data reported in the literature, there appear to be five major factors that affect the time-space removal of the residual suspended matter from a biological treatment process within a granular filter for a given temperature. These factors are the size of the filter medium, the rate of filtration, the influent particle size and size distribution, the floc strength, and the amount of material removed within the filter. Although a number of different formulations are possible, a generalized rate equation in which all five factors are considered can be developed by multiplying Eq. (11–27) by a factor that takes into account the effect of the material accumulated in the filter. The proposed equation is

$$\frac{dC}{dz} = -\frac{1}{(1 + az)^n} r_o C\left(1 - \frac{q}{q_u}\right)^m \tag{11–29}$$

where q = quantity of suspended solids deposited in the filter
q_u = ultimate quantity of solids that can be deposited in the filter
m = a constant related to floc strength

Initially, when the amount of material removed by the filter is low, $q = 0$; $(1 - q/q_u)^m = 1$, and Eq. (11–29) is equivalent to Eq. (11–27). As the upper layers begin to clog, the term $(1 - q/q_u)^m$ approaches zero, and the rate of change in concentration with distance is equal to zero. At the lower depths, the amount of material removed is essentially zero, and the previous analysis applies.

Headloss Development. In the past, the most commonly used approach to determine headloss in a clogged filter was to compute it with a modified form of the equations used to evaluate the clear-water headloss (see Table 11–5). In all cases, the difficulty encountered in using these equations is that the porosity must be estimated for various degrees of clogging. Unfortunately, the complexity of this approach renders most of these formulations useless or, at best, extremely difficult to use.

An alternative approach is to relate the development of headloss to the amount of material removed by the filter. The headloss would then be computed using the expression

$$H_t = H_o + \sum_{i=1}^{n} (h_i)_t \tag{11–30}$$

where H_t = total headloss at time t, m (ft)
H_o = total initial clear-water headloss, m (ft)
$(h_i)_t$ = headloss in the ith layer of the filter at time t, m (ft)

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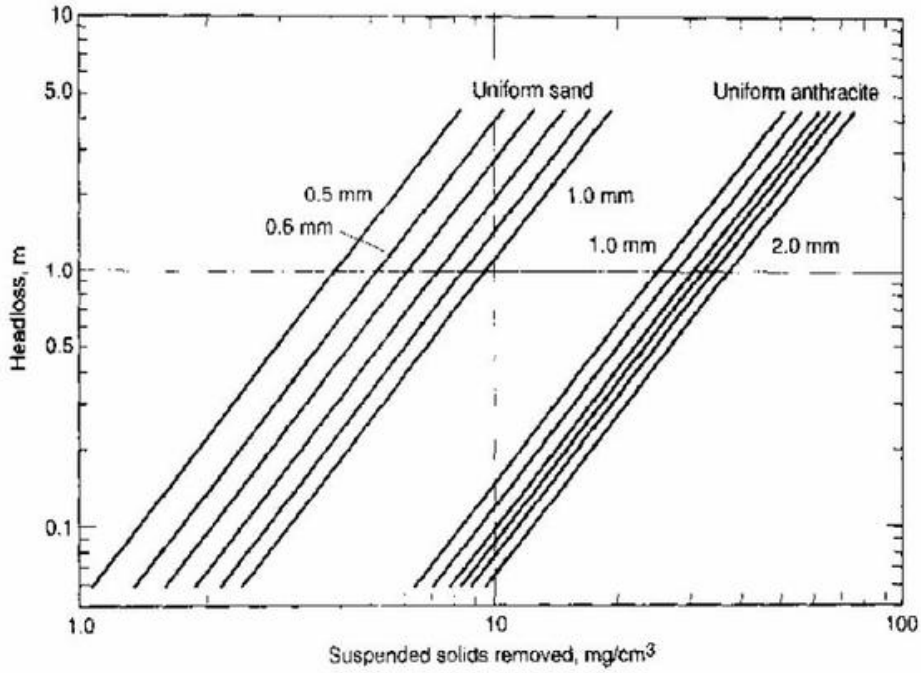
Figure 11-10

Headloss versus suspended solids removed for various sizes of uniform sand and anthracite. *(Adapted from Tchobanoglous and Eliassen, 1970.)*



From an evaluation of the incremental headloss curves for uniform sand and anthracite, the buildup of headloss in an individual layer of the filter was found to be related to the amount of material contained within the layer (see Fig. 11–10). The form of the resulting equation for headloss in the ith layer is

$$(h_i)_t = a(q_i)_t^b \tag{11-31}$$

where $(q_i)_t$ = amount of material deposited in the ith layer at time t, mg/cm^3
a, b – constants

In Eq. (11–31), it is assumed that the buildup of headloss is only a function of the amount of material removed. Determination of the buildup of headloss during the filtration process using the data presented on Fig. 11–10 is illustrated in Example 11–3.

EXAMPLE 11–3 **Analysis of Filtration Data from a Pilot Plant used to Filter Settled Effluent from an Activated-Sludge Process** The normalized suspended-solids-removal-ratio curves shown below were derived from a filtration pilot-plant study conducted at an activated-sludge wastewater-treatment plant. Using these curves and the following data, develop curves that can be used to estimate (1) the headloss buildup as a function of the length of run and (2) the length of run to a terminal headloss of 3.0 m (10 ft) as a function of the filtration rate.

Biological treatment process

1. Solids retention time SRT = 15 d
2. Average suspended-solids concentration in effluent from secondary settling tank = 20 mg/L

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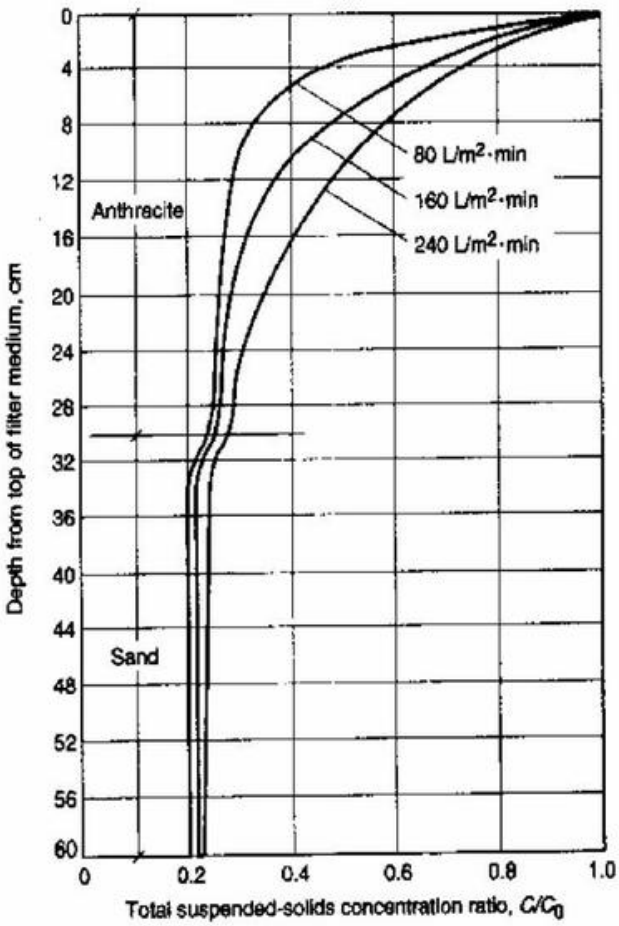
Pilot plant

1. Type of filter bed = dual-medium
2. Filter media = graded anthracite and sand
3. Filter-medium characteristics
   a. Anthracite, d_{10} = 1.4 mm, $UC \approx 1.4$
   b. Sand, d_{10} = 0.6 mm, $UC \approx 1.4$
4. Filter-bed depth = 0.3 m
   a. Anthracite = 0.3 m
   b. Sand = 0.3 m
5. Filtration rates = 80, 160, and 240 L/m²·min
6. Temperature = 20°C
7. General observation: average concentration-ratio curves given in the following figure did not vary significantly with time



Average suspended-solids concentration-ratio curves versus filter depth for various filtration rates.

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Solution

1. To analyze the concentration-ratio curves, rewrite Eq. (11–24) in a form suitable for numerical analysis:

$$-v\frac{\Delta C}{\Delta z} = \frac{\Delta q}{\Delta t}$$

$$-v\frac{C_{z+1} - C_z}{z_{z+1} - z_z} = \frac{q_2 - q_1}{t_2 - t_1}$$

2. Set up a computation table and determine the values of ΔC $(C_{z+1} - C_z)$ and the mean grain size for various depth intervals throughout the filter. The required computations are summarized in the following table. The entries in the table are explained below the table.

Depth, cm (1)	Geometric mean grain size, d_g, mm (2)	Filtration rate, L/m²·min: 80			160			240		
		C_z/C_o (3)	C_z (4)	ΔC (5)	C_z/C_o (6)	C_z (7)	ΔC (8)	C_z/C_o (9)	C_z (10)	ΔC (11)
0		1.0	20.0		1.0	20.0		1.0	20.0	
	1.18			11.4			7.4			5.6
4		0.43	8.6		0.63	12.6		0.72	14.4	
	1.42			2.2			3.2			2.8
8		0.32	6.4		0.47	9.4		0.58	11.6	
	1.56			0.8			2.2			2.0
12		0.28	5.6		0.36	7.2		0.48	9.6	
	1.69			0.2			1.2			1.6
16		0.27	5.4		0.30	6.0		0.40	8.0	
	1.82			0.2			0.4			1.0
20		0.26	5.2		0.28	5.2		0.35	7.0	
	1.98			0.2			0.2			0.8
24		0.25	5.0		0.27	5.4		0.31	6.2	
	2.23			0.2			0.2			0.6
28		0.24	4.8		0.26	5.2		0.28	5.6	
	2.62			0.0			0.0			0.2
30		0.24	4.8		0.26	5.2		0.27	5.4	
	0.50			0.6			0.6			0.5
32		0.215	4.3		0.23	4.6		0.245	4.9	
	0.60			0.1			0.1			0.1
34		0.205	4.1		0.23	4.5		0.24	4.8	
	0.65			0.1			0.1			0.0
36		0.20	4.0		0.22	4.4		0.24	4.8	
	0.69			0.0			0.0			0.0
38		0.20	4.0		0.22	4.4		0.24	4.8	
	0.72			0.0			0.0			0.0
40		0.20	4.0		0.22	4.4		0.24	4.8	

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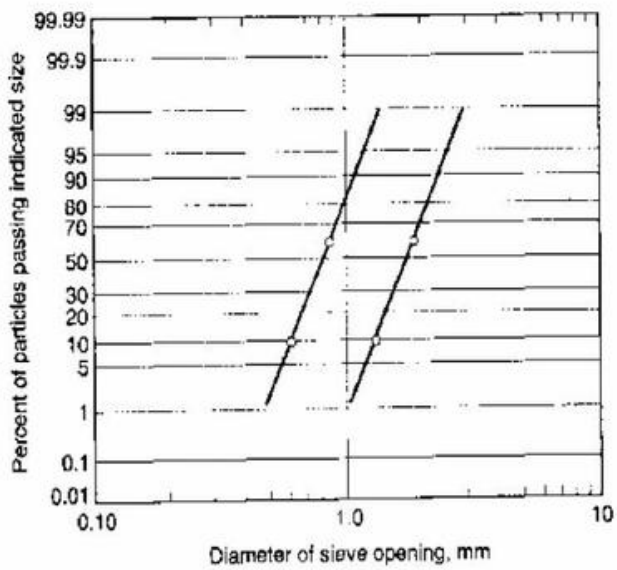
a. As shown in the computation table, the filter depth intervals are entered in col. 1.

b. The mean anthracite and sand particle size associated with each depth interval is entered in col. 2. The mean particle size for each interval can be estimated using the effective size and uniformity coefficient for each medium. For example, at a depth of 4 cm, 4/30ths of the total depth of 30 cm has been passed. Therefore, the particle size associated with a depth of 4 cm corresponds to the sieve size that 13.3 percent [(4/30) × 100] of the anthracite will pass. By plotting the particle size distributions on probability-log paper, and obtaining the equation of the line passing through the plotted points, the 13.3 percent passing anthracite particle size is estimated to be 1.29 mm (see figure below). Assuming the 1 percent size as the cutoff size (1.08 mm), the geometric mean particle size for the interval from zero and 4 cm is, as shown below, 1.18 mm. Following the same procedure, the remaining particle sizes for the anthracite and sand are entered in col. 2.

$$d_g = \sqrt{d_1 d_2} = \sqrt{1.08 \text{ mm} \times 1.29 \text{ mm}} = 1.18 \text{ mm}$$



c. The values of C/C_0 taken from the normalized removal ratio curves corresponding to the depths given in col. 1 are entered in cols. 3, 6, and 9 for each filtration rate.

d. The value of the concentration at each depth is entered in cols. 4, 7, and 10 for each filtration rate.

e. The concentration difference ΔC $(C_{z-1} - C_z)$ between the depths given in col. 1 is entered in cols. 5, 8, and 11 for each filtration rate.

3. Set up a computation table and determine the buildup of suspended solids and headloss within each layer of the filter for filter runs of various lengths. The necessary computations for a filtration rate of 160 L/m²·min (4 gal/ft²·min) are summarized in the following table.

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		Run length, h					
		10		15		20	
Depth, cm (1)	ΔC, mg/L (2)	Δq, mg/cm³ (3)	Δh, m (4)	Δq, mg/cm³ (5)	Δh, m (6)	Δq, mg/cm³ (7)	Δh, m (8)
0							
	7.4	17.76	0.47	26.64	1.15	35.52	2.17
4							
	3.2	7.68	0.06	11.52	0.14	15.36	0.26
8							
	2.2	5.28	0.02	7.92	0.05	10.56	0.10
12							
	1.2	2.88	—	4.32	0.01	5.76	0.02
16							
	0.4	0.96	—	1.44	—	1.92	—
20							
	0.2	0.48	—	0.72	—	0.96	—
24							
	0.2	0.48	—	0.72	—	0.96	—
28							
	0.0	0.00	—	0.00	—	0.00	—
30							
	0.6	2.88	0.43	4.32	1.04	5.76	1.95
32							
	0.1	0.48	0.01	0.72	0.01	0.96	0.03
36							
	0.1	0.48	0.01	0.72	0.01	0.96	0.02
38							
	0.0	0.0	—	0.0	—	0.0	—
40							
$\Sigma\,\Delta h$, m			1.00		2.41		4.55

Although the required computations for the other filtration rates are not shown, they are the same. The values of ΔC given in col. 2 are taken from col. 8 of the table prepared in Step 2. The values of q shown in cols. 3, 5, and 7 are determined by using the difference equation given in Step 1. To illustrate, for the anthracite layer between 4 and 8 cm from the top of the filter bed, the value of Δq after 20 h is as follows:

$$-v\frac{\Delta C}{\Delta z} = \frac{\Delta q}{\Delta t}$$

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where $v = 160\ \text{L/m}^2\cdot\text{min} = 0.016\ \text{L/cm}^2\cdot\text{min}$

$$\Delta C = 3.2\ \text{mg/L}$$
$$\Delta z = 4\ \text{cm} - 8\ \text{cm} = -4\ \text{cm}$$
$$\Delta t = (20\ \text{h} \times 60\ \text{min/h} - 0) = 1200\ \text{min}$$
$$\Delta q = \left(-\frac{0.016\ \text{L}}{\text{cm}^2\cdot\text{min}}\right)\left(\frac{3.2\ \text{mg/L}}{-4\ \text{cm}}\right)1200\ \text{min}$$
$$= 15.36\ \text{mg/cm}^3$$

The value of incremental headloss buildup Δh (col. 8) for the anthracite layer between 4 and 8 cm is obtained from Fig. 11–10 by entering with the value of Δq for this layer. The value of the headloss in the sand layers is determined in a similar manner. To simplify the computations, it is assumed that no intermixing occurs between the anthracite and sand. Once all of the Δh headloss values are entered, the entire column is summed to obtain the total headloss in the filter bed. The total headloss for other time periods and filtration rates is determined in exactly the same manner. Summary data for the other flowrates are as follows:

Run length, h	Headloss, m		
	80 L/m²·min	160 L/m²·min	240 L/m²·min
10	0.34	1.00	1.51
15	—	2.41	3.70
20	1.56	4.55	6.96
30	3.82		

4. The clean-water headloss at the three filtration rates, computed as illustrated previously in Example 11–1, is

Filtration rate, L/m²·min	Clean-water headloss, m
80	0.15
160	0.31
240	0.49

5. The total headloss including the clean-water headloss and the headloss due to the accumulation of material within the filter for each of the filtration rates is summarized below.

Run length, h	Headloss, m		
	80 L/m²·min	160 L/m²·min	240 L/m²·min
0	0.15	0.31	0.49
10	0.49	1.31	2.00
15	—	2.72	4.18
20	1.71		
30	3.97		

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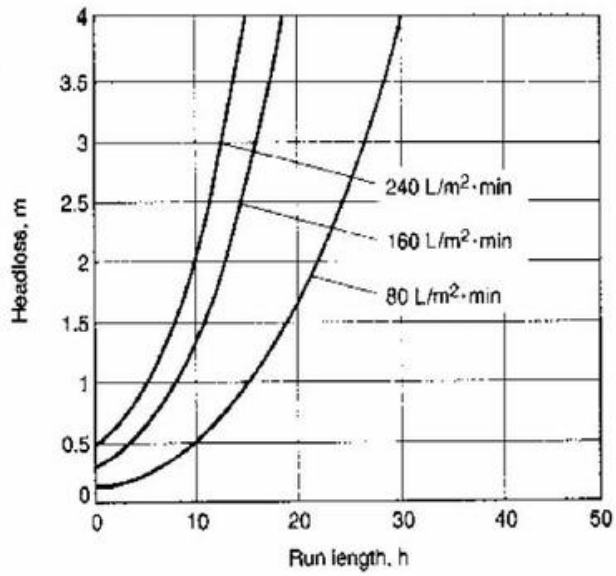
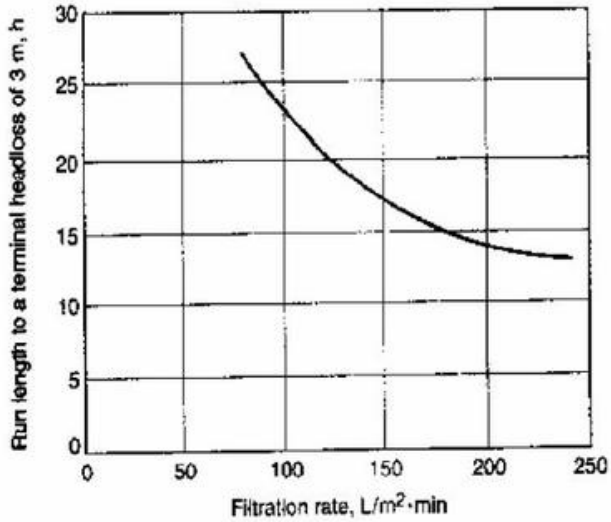
6. Plot curves of headloss versus run length for the three flowrates. The required curves, which are plotted using the data given in Step 5, are shown in the following figure.



7. Plot the curve of run length to reach a headloss of 3 m versus filtration rate. The required curve is shown below. The data needed to plot this curve are obtained from the headloss curve developed in Step 6 by finding the time required to reach a headloss of 3 m for each filtration rate.



Comment The use of unstratified filter beds for the filtration of treated effluents has been studied by Dahab and Young (1977). They found that unstratified filter beds having the same effective size as that used in the top layer of a dual-medium filter were essentially equivalent to dual-medium filters in terms of effluent quality and length of run.

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11-4 SELECTION AND DESIGN CONSIDERATIONS FOR DEPTH FILTERS

The ability to select and design filter technologies must be based on (1) knowledge of the types of filters that are available, (2) a general understanding of their performance characteristics, and (3) an appreciation of the process variables controlling depth filtration. Important design considerations for effluent filtration systems include (1) influent wastewater characteristics, (2) design and operation of the biological treatment process, (3) type of filtration technology to be used, (4) available flow-control options, (5) type of filter backwashing system to be employed, (6) necessary filter appurtenances, and (7) filter control systems and instrumentation. An understanding of issues related to effluent filtration with chemical addition, the type of filter problems encountered in the field, and the importance of pilot-plant studies is also necessary. These subjects are presented and discussed in this section.

Available Filtration Technologies

Historically, the granular-medium filter in which the complete filtration cycle (filtration and backwashing) occurs sequentially, as described in Sec. 11-3, has been used most commonly. However, during the past 20 years, a number of new types of filter technologies have been developed and are now available for the filtration of treated wastewater. The principal types of depth filters that have been used for the filtration of wastewater are identified in Table 11-6 and illustrated on Fig. 11-11. As shown in Table 11-6, the filters can be classified in terms of their operation as semicontinuous or continuous. Filters that must be taken off-line periodically to be backwashed are classified operationally as semicontinuous. Filters in which the filtration and backwash operations occur simultaneously are classified as continuous. Within each of these two classifications there are a number of different types of filters depending on bed depth (e.g., shallow, conventional, and deep bed), the type of filtering medium used (mono-, dual-, and multimedium), whether the filtering medium is stratified or unstratified, the type of operation (downflow or upflow), and the method used for the management of solids (i.e., surface or internal storage). For the mono- and dual-medium semicontinuous filters, a further classification can be made based on the driving force (e.g., gravity or pressure). Another important distinction that must be noted for the filters identified in Table 11-6 is whether they are proprietary or individually designed.

The five types of depth filters used most commonly for wastewater filtration are (1) conventional downflow filters, (2) deep-bed downflow filters, (3) deep-bed upflow continuous-backwash filters, (4) the pulsed-bed filter, and (5) traveling-bridge filters. These filter technologies are described briefly in this section. In addition a new filter employing a synthetic filter medium and a two-stage filtration system which incorporates phosphorus removal are also discussed. Intermittent and recirculating sand and textile filters, usually limited to small systems, are not considered in this text. Information on these filters may be found in Crites and Tchobanoglous (1998). It should be noted that many of the filters to be considered in the following discussion are proprietary and are supplied by the manufacturer as a complete unit. Thus, many of the design details presented in the following section apply only to filters that are designed individually.

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Table 11–6
Comparison of principal types of granular-medium filters

Type of filter (common name)	Type of filter operation	Filter-bed details			Typical direction of fluid flow
		Type of filter bed	Filtering medium	Typical bed depth, mm	
Conventional	Semicontinuous	Mono-medium (stratified or unstratified)	Sand or anthracite	760	Downward
Conventional	Semicontinuous	Dual-medium (stratified)	Sand and anthracite	920	Downward
Conventional	Semicontinuous	Multimedium (stratified)	Sand, anthracite, and garnet	920	Downward
Deep-bed	Semicontinuous	Mono-medium (stratified or unstratified)	Sand or anthracite	1830	Downward
Deep-bed	Semicontinuous	Mono-medium (stratified)	Sand or anthracite	1830	Upward
Deep-bed	Continuous	Mono-medium (unstratified)	Sand	1830	Upward
Pulsed-bed	Semicontinuous	Mono-medium (stratified)	Sand	280	Downward
Fuzzy filter	Semicontinuous	Mono-medium (unstratified)	Synthetic fiber	610[a]	Upward
Traveling-bridge	Continuous	Mono-medium (stratified)	Sand	610	Downward
Traveling-bridge	Continuous	Dual-medium (stratified)	Sand and anthracite	410	Downward

[a]Compressed depth.

Conventional Downflow Filters. Single-, dual-, or multimedium filter materials are utilized in conventional downflow depth filters. Typically sand or anthracite is used as the filtering material in single-medium filters (see Fig. 11–11*a*). Dual-medium filters usually consist of a layer of anthracite over a layer of sand (see Fig. 11–11*b*). Other combinations include (1) activated carbon and sand, (2) resin beads and sand, and (3) resin beads and anthracite. Multimedium filters typically consist of a layer of anthracite over a layer of sand over a layer of garnet or ilmenite. Other combinations include (1) activated carbon, anthracite, and sand, (2) weighted, spherical resin beads, anthracite, and sand, and (3) activated carbon, sand, and garnet.

Dual- and multimedium and deep-bed mono-medium depth filters were developed to allow the suspended solids in the liquid to be filtered to penetrate farther into the filter bed, and thus use more of the solids-storage capacity available within the filter bed (see Fig. 11–12). The deeper penetration of the solids into the filter bed also permits longer filter runs because the buildup of headloss is reduced. By comparison, in shal-

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Table 11-6 (*Continued*)

Backwash operation	Flowrate through filter	Solids storage location	Remarks	Type of design
Batch	Constant/variable	Surface and upper bed	Rapid headloss buildup	Individual
Batch	Constant/variable	Internal	Dual-medium design used to extend filter run length	Individual
Batch	Constant/variable	Internal	Multimedium design used to extend filter run length	Individual
Batch	Constant/variable	Internal	Deep bed used to store solids and extend filter run length	Individual
Batch	Constant	Internal	Deep bed used to store solids and extend filter run length	Proprietary
Continuous	Constant	Internal	Sand bed moves in countercurrent direction to fluid flow	Proprietary
Batch	Constant	Surface and upper bed	Air pulses used to break up surface mat and increase run length	Proprietary
Batch	Constant	Internal		Proprietary
Semicontinuous	Constant	Surface and upper bed	Individual filter cells backwashed sequentially	Proprietary
Semicontinuous	Constant	Surface and upper bed	Individual filter cells backwashed sequentially	Proprietary

low mono-medium beds, most of the removal occurs in the upper few millimeters of the bed (Tchobanoglous and Eliassen, 1970). The operation of conventional downflow filters has been described previously (see Fig. 11–4). Typically, water wash plus surface water and water wash plus air scour are the principal methods used in backwashing mono-, dual-, and multimedium filter beds. With either of these backwashing methods, the filter mediums are fluidized.

Deep-Bed Downflow Filters. The deep-bed downflow filter (see Fig. 11–11*c*) is similar to the conventional downflow filter with the exception that the depth of the filter bed and the size of the filtering medium (usually anthracite) are greater than the corresponding values in a conventional filter. Because of the greater depth and larger medium size (i.e., sand or anthracite), more solids can be stored within the filter bed and the run length can be extended. The maximum size of the filter medium used in these filters will depend on the ability to backwash the filter. In general, deep-bed filters are not fluidized completely during backwashing. To achieve effective cleaning, air scour plus water is used in the backwash operation.

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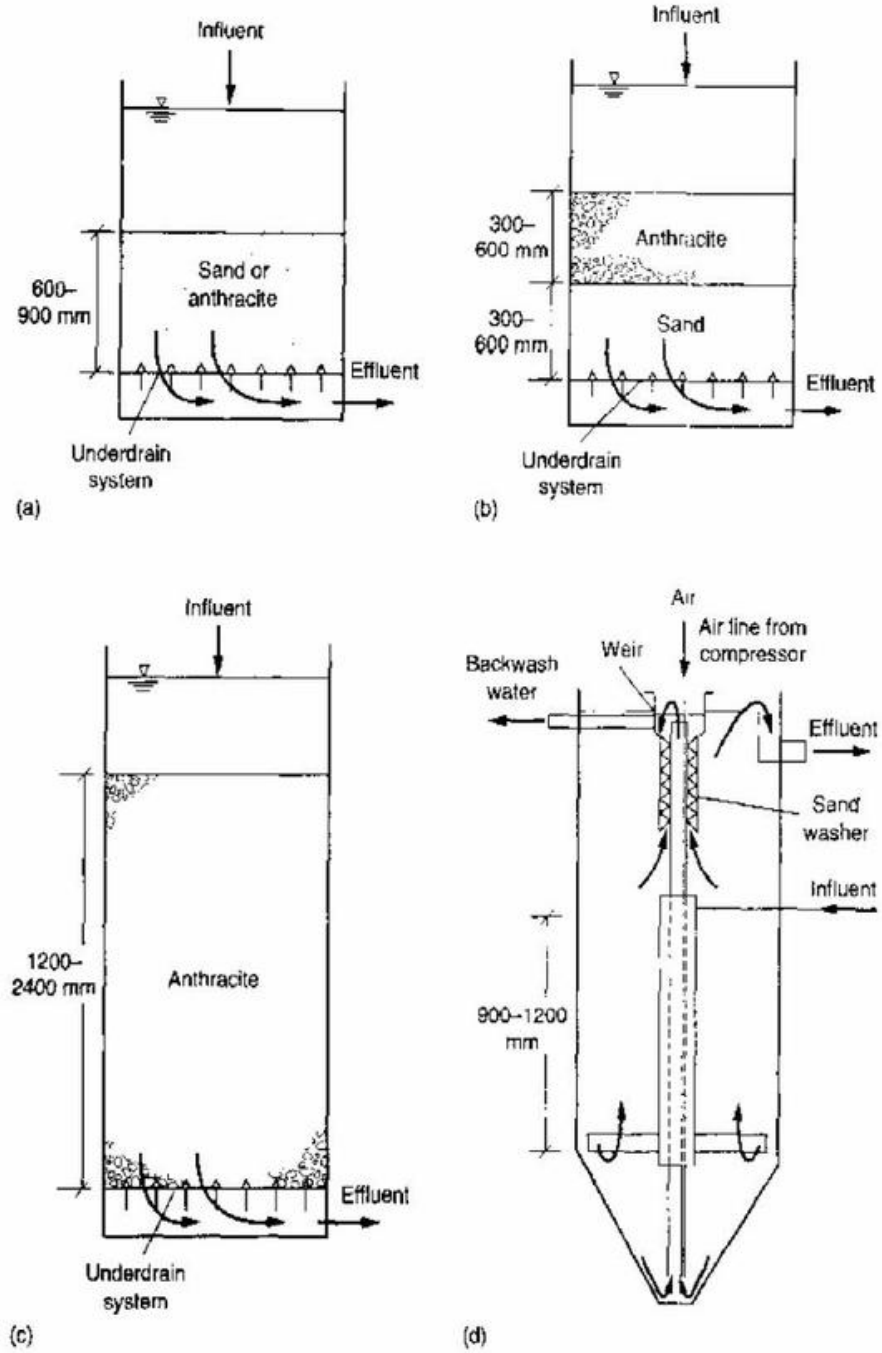

Figure 11–11

Definition sketches for the principal types of granular medium filters: (*a*) conventional mono-medium downflow filter, (*b*) conventional dual-medium downflow filter, (*c*) conventional mono-medium deep-bed downflow filter, (*d*) continuous backwash deep-bed upflow filter, (*e*) pulsed-bed filter, (*f*) traveling-bridge filter, (*g*) synthetic-medium filter, (*h*) pressure filter, and (*i*) slow sand filter.

Deep-Bed Upflow Continuous Backwash Filters. In this filter, shown schematically (see Fig. 11–11*d*) and pictorially on Fig. 11–13, the wastewater to be filtered is introduced into the bottom of the filter where it flows upward through a series of riser tubes and is distributed evenly into the sand bed through the open bottom of an inlet distribution hood. The water then flows upward through the downward-moving

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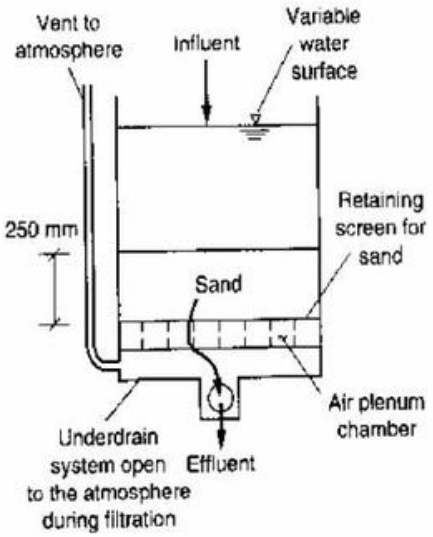
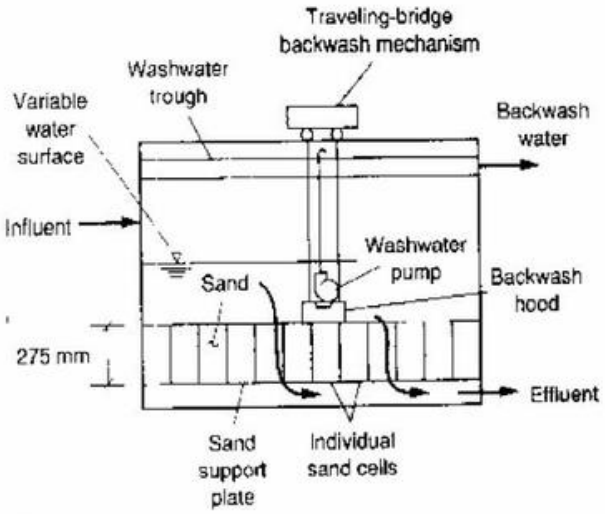


(e)



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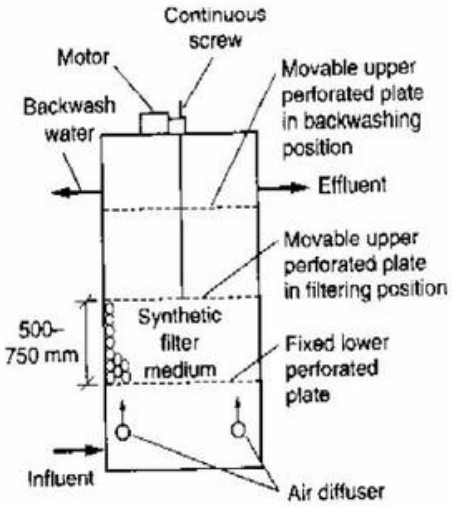
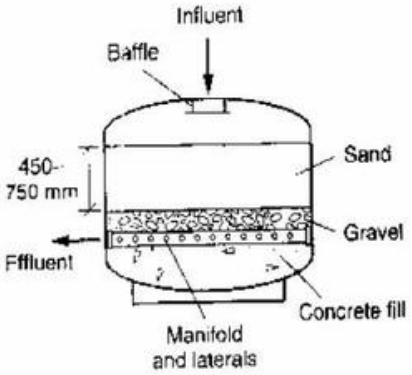
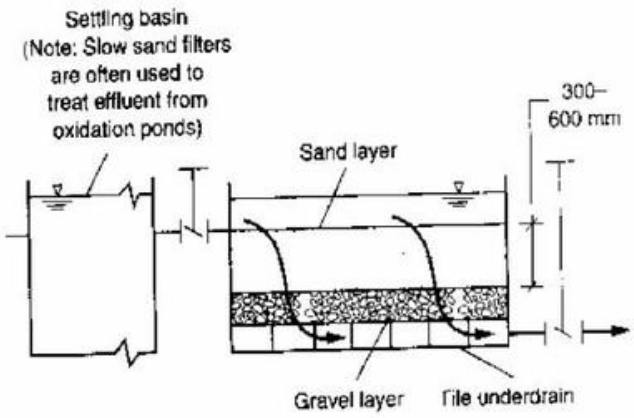


(g)



(h)



(i)

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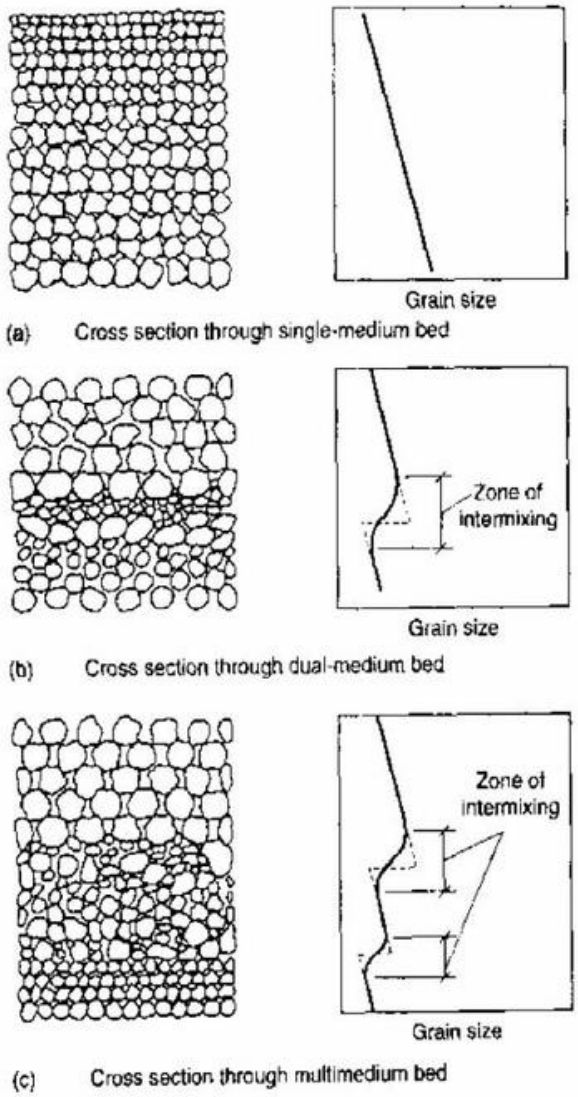
Figure 11–12

Schematic diagram of filter beds illustrating potential increase in storage capacity: (a) single medium, (b) dual medium, and (c) multimedium.



sand. The clean filtrate exits from the sand bed, overflows a weir, and is discharged from the filter. At the same time sand particles, along with trapped solids, are drawn downward into the suction of an airlift pipe that is positioned in the center of the filter. A small volume of compressed air, introduced into the bottom of the airlift, draws sand, solids, and water upward through the pipe by creating a fluid with a density less than 1.

Impurities are scoured (abraded) from the sand particles during the turbulent upward flow. Upon reaching the top of the airlift, the dirty slurry spills over into the central reject compartment. By setting the filtrate weir above the reject weir, a steady stream of clean filtrate flows upward, countercurrent to the movement of sand, through the washer section. The upflow liquid carries away the solids and reject water. Because the sand has a higher settling velocity than the removed solids, the sand is not carried out of the filter. The sand is cleaned further as it moves down through the washer. The cleaned sand is redistributed onto the top of the sand bed, allowing for a continuous uninterrupted flow of filtrate and reject water.

Pulsed-Bed Filters. The pulsed-bed filter (see Figs. 11–11*e* and 11–14) is a proprietary downflow gravity filter with an unstratified shallow layer of fine sand as the fil-

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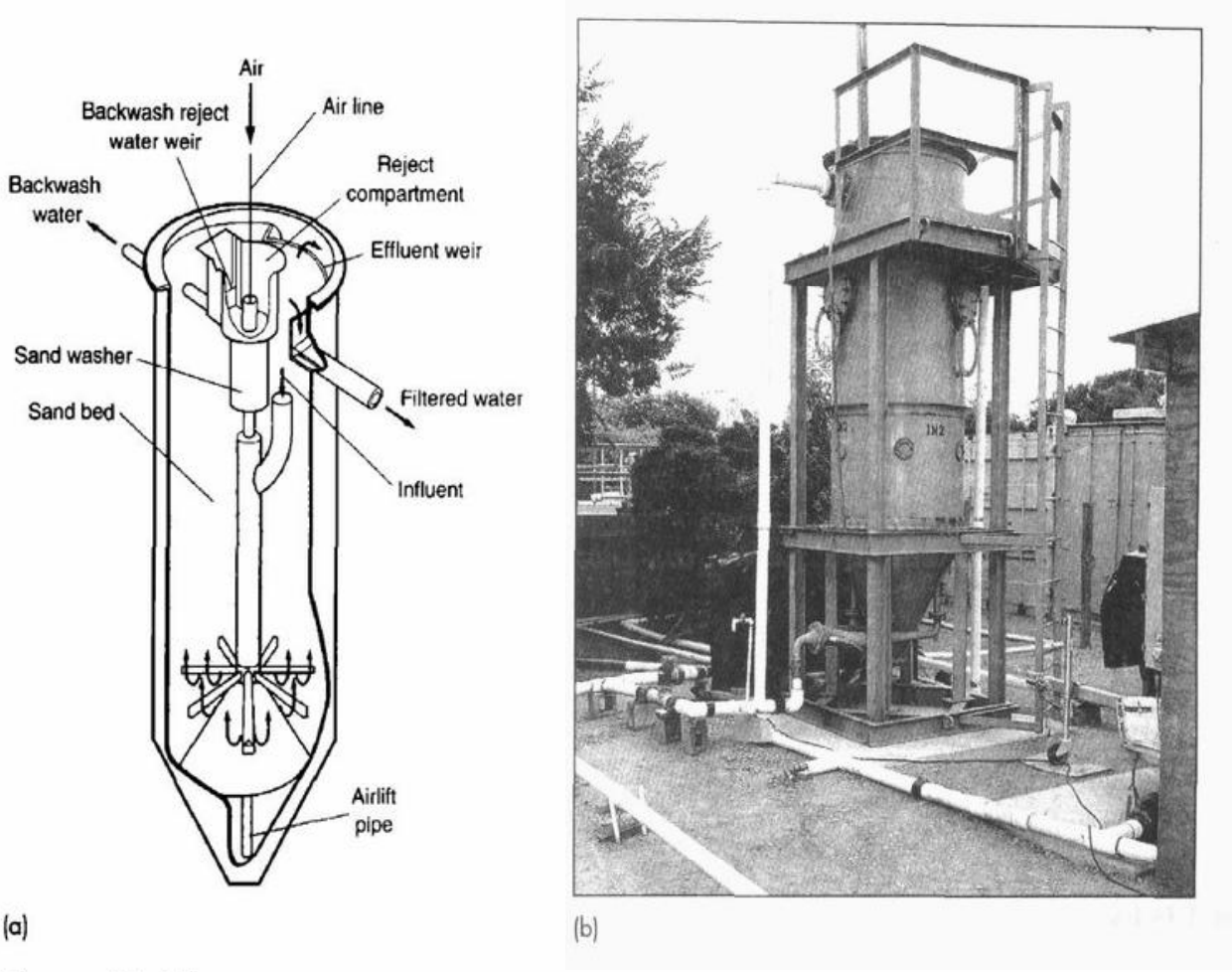

Figure 11-13

Deep-bed upflow continuous backwash filter: (*a*) schematic view and (*b*) view of test unit with older style influent feed system. *Note:* The influent feed assembly will vary depending on the filter configuration (see Fig. 11-22). *(Courtesy: Parkson Corporation.)*

tering medium. The shallow bed is used for solids storage, as opposed to other shallow-bed filters where solids are principally stored on the sand surface. An unusual feature of this filter is the use of an air pulse to disrupt the sand surface and thus allow penetration of suspended solids into the bed. The air pulse process involves forcing a volume of air, trapped in the underdrain system, up through the shallow filter bed to break up the surface mat of solids and renew the sand surface. When the solids mat is disturbed, some of the trapped material is suspended into the admixture over the sand, but the majority of solids are entrapped within the filter bed. The intermittent air pulse causes a folding over the sand surface, burying solids within the medium and regenerating the filter bed surface. The filter continues to operate with intermittent pulsing until a terminal headloss limit is reached. The filter then operates in a conventional backwash cycle to remove solids from the sand. It should be noted that during normal operation the filter underdrain is not flooded as it is in a conventional filter.

Traveling-Bridge Filters. The traveling-bridge filter (see Figs. *11-11f* and 11-15) is a proprietary continuous downflow, automatic backwash, low-head, granular medium-depth filter. The bed of the filter is divided horizontally into long independent

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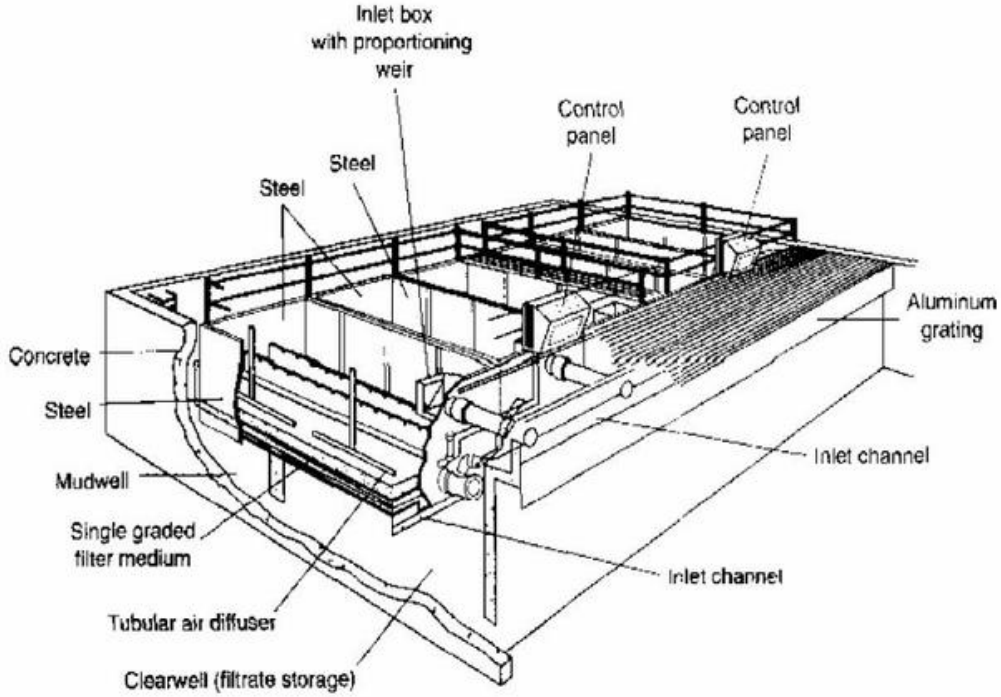
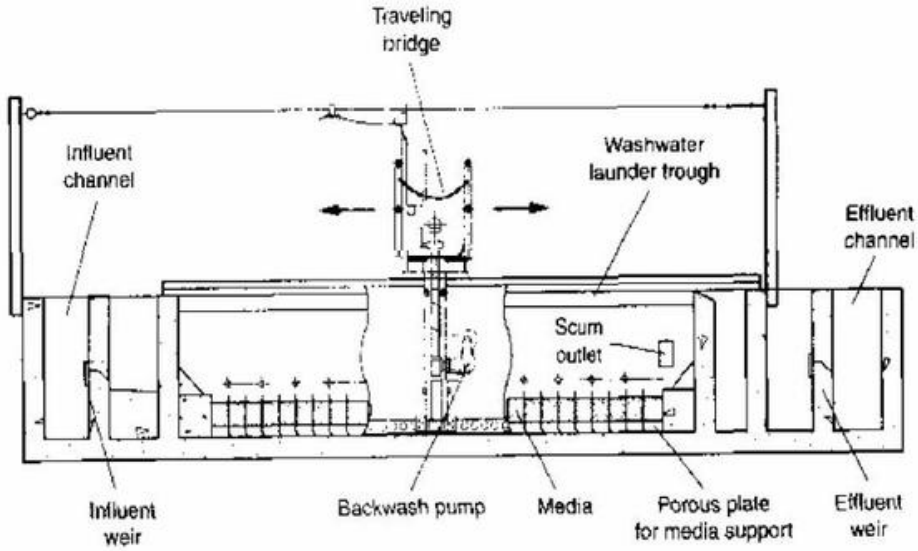


Figure 11-14

Schematic view of pulsed-bed filter. *(Courtesy of U.S. Filter.)*

Figure 11-15

Traveling-bridge filter. *(Courtesy: Aqua Aerobic, Inc.)*



filter cells. Each filter cell contains approximately 280 mm (11 in) of medium. Treated wastewater flows through the medium by gravity and exits to the clearwell plenum via a porous-plate, polyethylene underdrain. Each cell is backwashed individually by an overhead, traveling-bridge assembly, while all other cells remain in service. Water used for backwashing is pumped directly from the clearwell plenum up through the medium and deposited in a backwash trough. During the backwash cycle, wastewater is filtered continuously through the cells that are not being backwashed. The backwash mechanism includes a surface wash pump to assist in breaking up of the surface matting and

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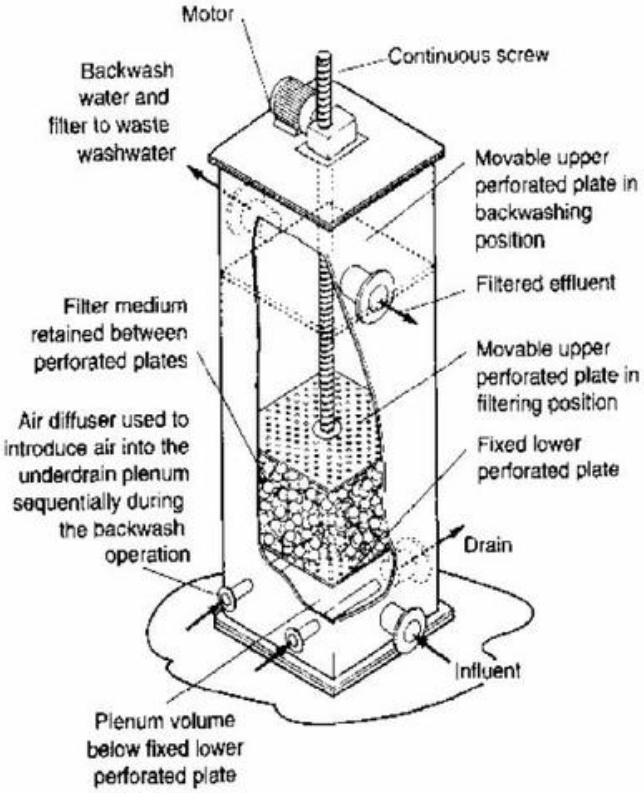
"mudballing" in the medium. Because the backwashing operation is performed on an "as needed" basis, the backwash cycle is termed semicontinuous (see Table 11–6).

Synthetic-Medium Filters. A synthetic-medium filter, developed originally in Japan, is now being used for the filtration wastewater. The highly porous synthetic filter medium, made of polyvaniladene, has a diameter of approximately 30 mm (1.25 in). Based on displacement tests, the porosity of the uncompacted quasi-spherical filter medium itself is estimated to be about 88 to 90 percent, and the porosity of the filter bed made up of the filter medium is approximately 94 percent. Unusual features of the filter are: (1) the porosity of the filter bed can be modified by compressing the filter medium and (2) the size of the filter bed is increased mechanically to backwash the filter (see Figs. 11–11*g* and 11–16). The filter medium also represents a departure from conventional filter media in that the fluid to be filtered flows through the medium as opposed to flowing around the filtering medium, as in sand and anthracite filters. As a result of the high porosity, filtration rates of 400 to 1200 L/m^2·min (10 to 30 gal/ft^2·min) have been pilot-tested (Caliskaner and Tchobanoglous, 2000).

In the filtering mode, secondary effluent is introduced in the bottom of the filter. The influent wastewater flows upward through the filter medium, retained by two porous plates, and is discharged from the top of the filter. To backwash the filter, the upper porous plate is raised mechanically. While flow to the filter continues, air is introduced sequentially from the left and right sides of the filter below the lower porous plate, causing the filter medium to move in a rolling motion. The filter medium is cleaned by the shearing forces as the wastewater moves past the filter and by abrasion as the filter medium rubs against itself. Wastewater containing the solids removed from

Figure 11–16

Schematic diagram of the Fuzzy filter. *(Courtesy: Schreiber Corporation.)*



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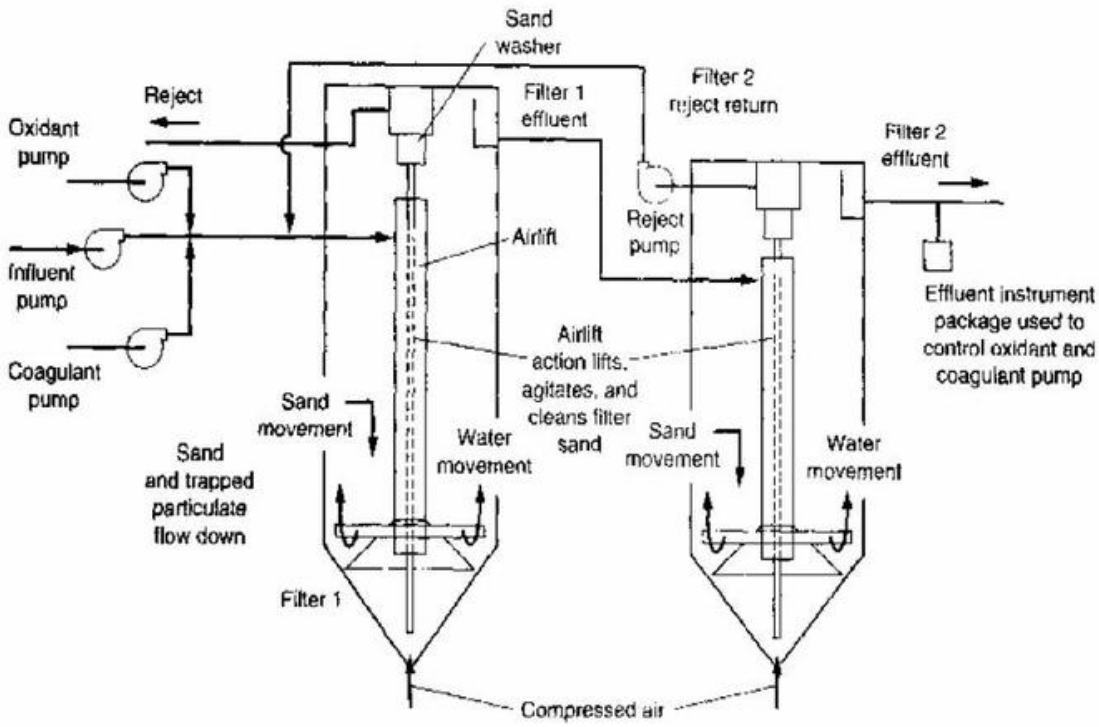

Figure 11–17

Flow diagram for a two-stage filtration process for the removal of turbidity, total suspended solids, and phosphorus. (Courtesy DSS Environmental, Inc.)

the filter is diverted for subsequent processing. To put the filter back into operation after the backwash cycle has been completed, the raised porous plate is returned to its original position. After a short flushing cycle, the filtered effluent valve is opened, and filtered effluent is discharged.

Two-Stage Filtration. A schematic flow diagram for a proprietary two-stage filtration process for the removal of turbidity, total suspended solids, and phosphorus is shown on Fig. 11–17. As shown, two deep-bed upflow continuous backwash filters are used in series to produce a high-quality effluent. A large-size sand diameter is used in the first filter to increase the contact time and to minimize clogging. A smaller sand size is used in the second filter to remove residual particles from the first-stage filter. The backwash water from the second filter which contains small particles and residual coagulant is recycled to the first filter to improve floc formation within the first-stage filter and the influent to waste ratio. Based on full-scale installations the reject rate has been found to be less than 5 percent. Phosphorus levels equal to or less than 0.02 mg/L have been achieved in the final filter effluent.

Performance of Different Types of Filter Technologies

The critical question associated with the selection of any granular-medium depth filter is whether it will perform as anticipated. Insight into the performance of granular-medium filters can be gained from a review of the turbidity, total suspended solids, and particle-removal data for depth filters.

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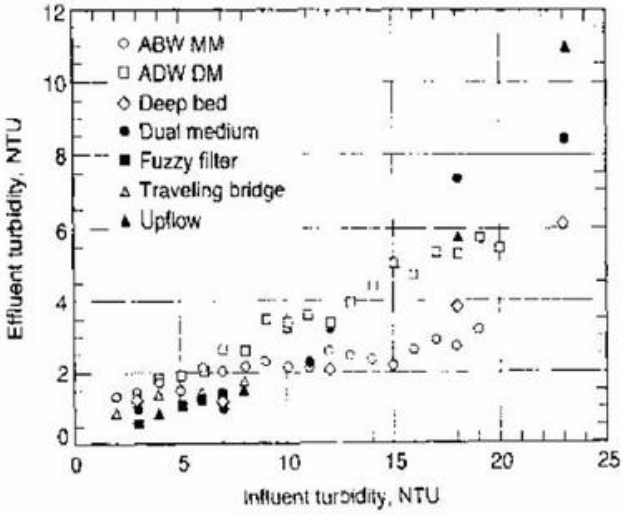

Figure 11-18

Performance data for seven different types of depth filters used for wastewater applications tested using the effluent from the same activated-sludge plant at filtration rate 160 L/m²·min (4 gal/ft²·min) with the exception of the Fuzzy filter, which was operated at 800 L/m²·min (20 gal/ft²·min).

Removal of Turbidity. The results of long-term testing of seven different types of pilot-scale filters on the effluent from the same activated-sludge process (SRT > 10 d), without chemical addition, are shown on Fig. 11-18. Long-term data from other large-scale wastewater reclamation plants are also shown. The principal conclusions to be reached from an analysis of the data presented on Fig. 11-18 are that (1) given a high-quality filter influent (turbidity less than 5 to 7 NTU) all of the filters tested including the large-scale plant are capable of producing an effluent with an average turbidity of 2 NTU or less, and (2) when the influent turbidity is greater than about 5 to 7 NTU, chemical addition will be required with all of the filters to achieve an effluent turbidity of 2 NTU or less.

Removal of Total Suspended Solids. Using the following two relationships between turbidity and suspended solids as given in Chap. 2 [see Eq. (2-16)], influent turbidity values of 5 to 7 NTU correspond to a total suspended solids concentration varying from about 10 to 17 mg/L, and an effluent turbidity of 2 NTU corresponds to a total suspended solids concentration varying from 2.8 to 3.2 mg/L.

Settled secondary effluent

$$\text{TSS, mg/L} = (2.0 \text{ to } 2.4) \times (\text{turbidity, NTU}) \tag{11-32}$$

Filter effluent

$$\text{TSS, mg/L} = (1.3 \text{ to } 1.5) \times (\text{turbidity, NTU}) \tag{11-33}$$

Long-term total suspended-solids performance data for the depth filters at the Donald C. Tillman Wastewater Reclamation Plant in Los Angeles, CA, are shown on Fig. 11-19. For the data shown on Fig. 11-19, the ratio of TSS to turbidity is about 1.33.

Removal of Particle Size. Typical data on the removal of particle sizes found in settled wastewater are shown on Fig. 11-20. As shown, the particle-removal rate is essentially independent of the filtration rate up to about 240 L/m²·min (6 gal/ft²·min).

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Figure 11-19

Probability distribution of product water turbidity and TSS for 1998 for the Donald C. Tillman Wastewater Reclamation Plant, Los Angeles, CA.

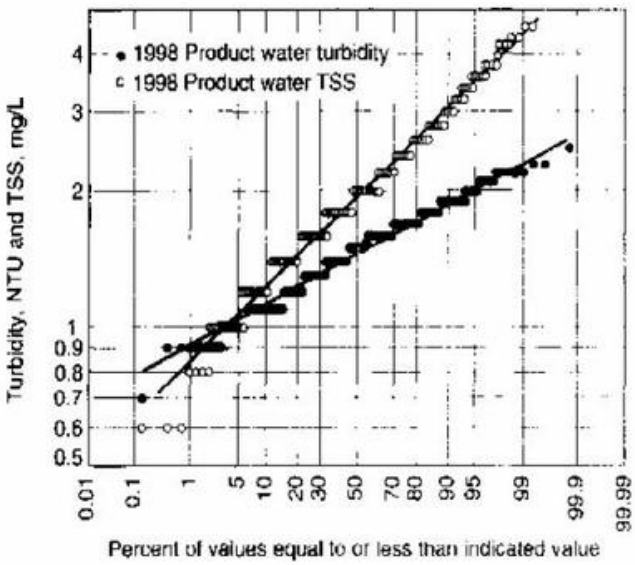
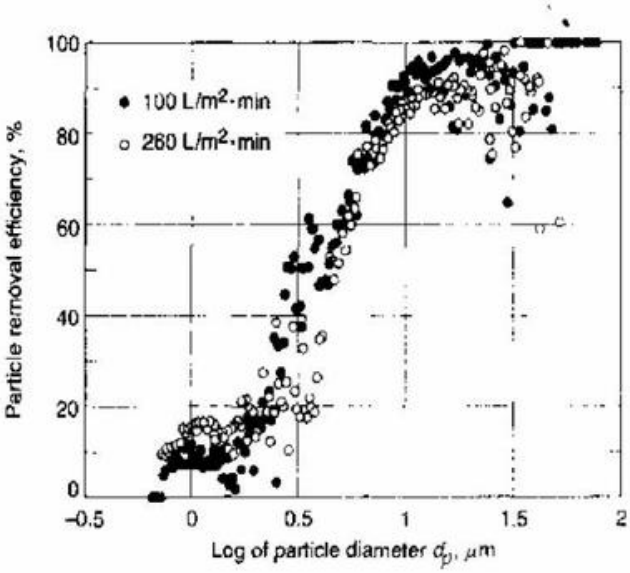


Figure 11-20

Particle-size removal efficiency for a depth filter for effluent from an activated-sludge plant.



It is significant that most depth filters will pass some particles with diameters up to 20 μm. Particles in the size range from 10 to 20 μm are of importance with respect to disinfection because they are of sufficient size to shield microorganisms (see Chap. 12).

Issues Related to Design and Operation of Treatment Facilities

For new wastewater-treatment plants, extra care should be devoted to the design of the secondary settling facilities. With properly designed settling facilities resulting in an effluent with low TSS (typically 5 mg/L), the decision on what type of filtration system is to be used is often based on plant-related variables, such as the space available, duration of filtration period (seasonal versus year-round), the time available for construction, and costs. For existing plants that have variable suspended solids concentrations in the treated effluent and must be retrofitted with effluent filtration, it may be appropriate to consider the type of a filter that can continue to function even when heavily loaded. The pulsed-bed filter and both downflow and upflow deep-bed coarse-medium filters have been used in such applications.

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Importance of Influent Wastewater Characteristics

The most important influent characteristics in the filtration of treated secondary effluents are the suspended-solids concentration, particle size and distribution, and floc strength.

Suspended Solids/Turbidity. Typically, the TSS concentration in the effluent from activated-sludge and trickling-filter plants varies between 6 and 30 mg/L. Because this concentration usually is the principal parameter of concern, turbidity is often used as a means of monitoring the filtration process. Corresponding turbidity values can vary from 3 to 15 NTU.

Particle Size and Distribution. Typical data on the particle size and distribution in the effluent from two activated-sludge plants were presented previously on Fig. 2-9 in Chap. 2. As illustrated, the particles fell into two distinct size ranges, small particles varying in areal size (equivalent circular diameter) from 0.8 to 1.2 μm and large particles varying in size from about 5 to 100 μm. In addition, a few particles larger than about 500 μm are almost always found in settled treated effluent. These particles are light and amorphous and do not settle readily (see discussion of hindered settling in Sec. 5-5). The weight fraction of the smaller particles was estimated to be approximately 40 to 60 percent of the total. This percentage will vary, however, depending on the operating conditions of the biological process and the degree of flocculation achieved in the secondary settling facilities.

The most significant observation related to particle size is that the distribution of sizes is bimodal. This observation is important because it will influence the removal mechanisms that may be operative during the filtration process. For example, it seems reasonable to assume that the removal mechanism for particles 1.0 μm in size would be different from that for particles in the size range from 10 to 100 μm. The bimodal particle-size distribution has also been observed in water-treatment plants.

Floc Strength. Floc strength, which will vary with the type of process and the mode of operation, is also important. For example, the residual floc from the chemical precipitation of biologically processed wastewater may be considerably weaker than the residual biological floc before precipitation. Further, the strength of the biological floc will vary with the mean cell-residence time, increasing with longer mean cell-residence time. The increased strength derives in part from the production of extracellular polymers as the mean cell-residence time is lengthened. At extremely long mean cell-residence times (15 days and longer), it has been observed that the floc strength will decrease due to floc breakup.

Selection of Filtration Technology

In selecting a filter technology, important issues that must be considered include (1) type of filter to be used: proprietary or individually designed, (2) the filtration rate, (3) filtration driving force, (4) number and size of filter units, and (5) the backwash-water requirements.

Type of Filter: Proprietary or Individually Designed. As noted in Table 11-6, the currently available filter technologies are either proprietary or individually designed. With proprietary filters, the manufacturer is responsible for providing

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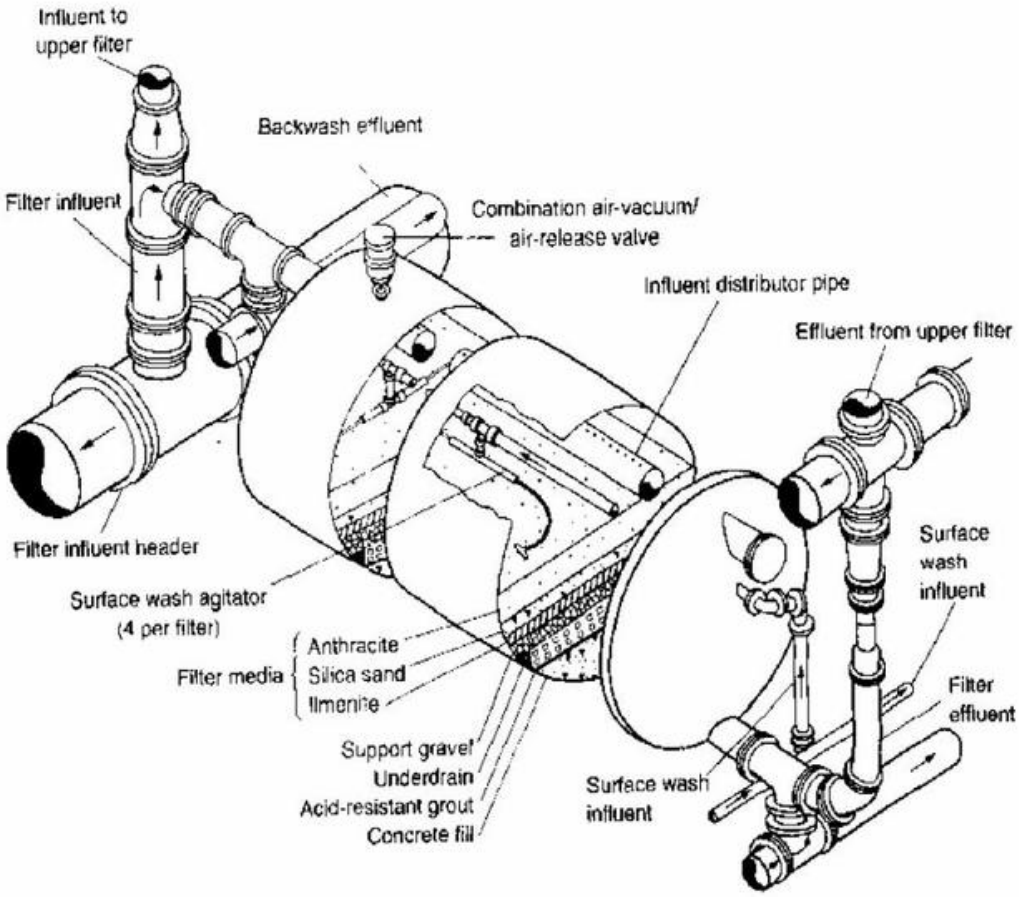
the complete filter unit and its controls, based on basic design criteria and performance specifications. In individually designed filters, the designer is responsible for working with several suppliers in developing the design of the system components. Contractors and suppliers then furnish the materials and equipment in accordance with the engineer's design.

Filtration Rate. The rate of filtration is important because it affects the real size of the filters that will be required. For a given filter application, the rate of filtration will depend primarily on the strength of the floc and the size of the filtering medium. For example, if the strength of the floc is weak, high filtration rates will tend to shear the floc particles and carry much of the material through the filter. It has been observed that filtration rates in the range of 80 to 320 $L/m^2 \cdot min$ (2 to 8 $gal/ft^2 \cdot min$) will not affect the effluent quality when filtering settled activated-sludge effluents.

Filtration Driving Force. Either the force of gravity or an applied pressure force can be used to overcome the frictional resistance to flow offered by the filter bed. Gravity filters of the type discussed above are used most commonly for the filtration of treated effluent at large plants. Pressure filters of the type shown on Fig. 11–21 operate in the same manner as gravity filters and are used at smaller plants. The only difference is that, in pressure filters, the filtration operation is carried out in a closed vessel under

Figure 11–21

Typical pressure filter with multimedium filter bed and surface wash used for wastewater filtration.



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pressurized conditions achieved by pumping. Pressure filters normally are operated at higher terminal headlosses, resulting in longer filter runs and reduced backwash requirements.

Number and Size of Filter Units. One of the early decisions to be made in the design of a depth filtration system is determining the number and size of filter units that will be required. The surface area required is based on the peak filtration and peak plant flowrates. The allowable peak filtration rate is usually established on the basis of regulatory requirements. Operating ranges for a given filter type are based on past experience, the results of pilot-plant studies, manufacturers' recommendations, and regulatory constraints. The number of units generally should be kept to a minimum to reduce the cost of piping and construction, but it should be sufficient to assure (1) that backwash flowrates do not become excessively large and (2) that when one filter unit is taken out of service for backwashing, the transient loading on the remaining units will not be so high that material contained in the filters will be dislodged. Transient loadings due to backwashing are not an issue with filters that backwash continuously. To meet redundancy requirements, a minimum of two filters will generally be required.

The sizes of the individual units should be consistent with the sizes of equipment available for use as underdrains, washwater troughs, and surface washers. Typically, width-to-length ratios for individually designed gravity filters vary from 1:1 to 1:4. A practical limit for the surface area on an individual depth filter (or filter cell) is about 100 m^2 (1100 ft^2), although larger filters units have been built. The layout of a bank of deep-bed upflow continuous-backwash filters is shown on Fig. 11–22. For proprietary

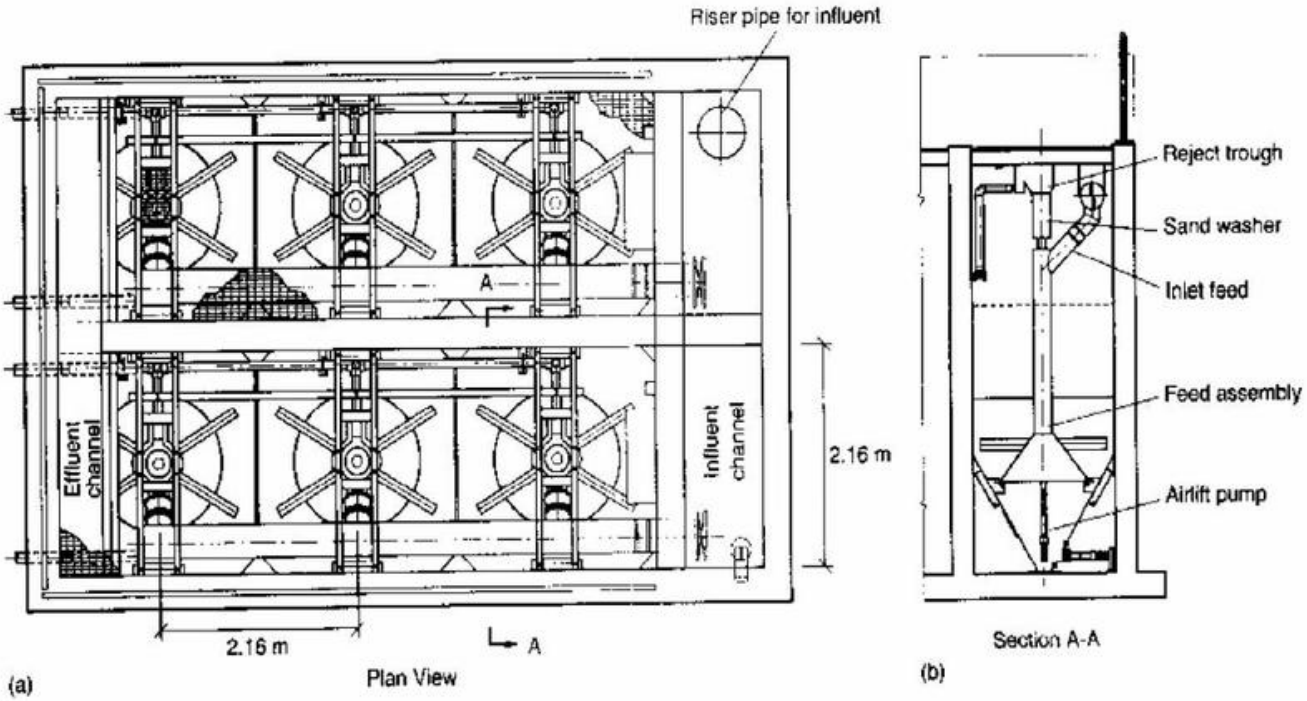

Figure 11–22

Layout of a bank of six deep-bed upflow continuous backwash filters. *(Courtesy Parkson Corp.)*

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and pressure filters, it is common practice to use standard sizes that are available from manufacturers. Depending on the manufacturing method, shipping constraints will also limit the size of pressure filters. The maximum diameter of vertical pressure filters is about 3.7 m (12 ft). The maximum diameter and length of horizontal pressure filters is about 3.7 m (12 ft) and 12 m (40 ft), respectively.

Backwash-Water Requirements. As noted in Table 11–6, depth filters operate in either a semicontinuous or continuous mode. In semicontinuous operation, the filter is operated until the effluent quality starts to deteriorate or the headloss becomes excessive, at which point the filter is taken out of service and backwashed to remove the accumulated solids. With filters operated in the semicontinuous mode, provision must be made for the backwash water needed to clean the filters. Typically, the backwash water is pumped from a filtered water clearwell or obtained by gravity from an elevated storage tank. For filters that operate continuously, such as the upflow filter (see Figs. 11–11*d* and 11–13) and the traveling-bridge filter (see Figs 11–11*f* and 11–15), the filtering and cleaning (backwashing) phases take place simultaneously. In the traveling-bridge filter, the backwash operation can be either continuous or semicontinuous as required. It should be noted that with filters that operate continuously, there is no turbidity breakthrough or terminal headloss.

Filter-Bed Characteristics

The principal variables that must be considered in the design of filters are identified in Table 11–7. In the application of filtration to the removal of residual suspended solids, it has been found that the nature of the particulate matter in the influent to be filtered, the filter-bed configuration, the size of the filter material or materials, and the filtration flowrate are perhaps the most important of the process variables.

Filter-Bed Configuration. The principal types of nonproprietary filter-bed configurations now used for wastewater filtration may be classified according to the number of filtering mediums that are used as mono-medium, dual-medium, or multimedium beds (see Fig. 11–12). In conventional downflow filters, the distribution of grain sizes for each medium after backwashing is from small to large. Typical design data for monomedium filters and dual- and multimedium filters are presented in Tables 11–8 and 11–9, respectively.

Selection of Filter Medium. Once the type of filter to be used has been selected, the next step is to specify the characteristics of the filtering medium, or media if more than one is used. Typically, this process involves the selection of the grain size as specified by the effective size d_{10}, uniformity coefficient UC, the 90 percent size, the specific gravity, solubility, hardness, and depth of the various materials used in the filter bed. Typical particle-size distribution ranges for sand and anthracite filtering material are shown on Fig. 11–23. The 90 percent size designated d_{90}, as read from a grain-size analysis, is commonly used to determine the required backwash rate for depth filters. Typical physical properties of filter materials used in depth filters are summarized in Table 11–10. The sizes of filter materials were given previously in Tables 11–7 and 11–8.

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Table 11-7
Principal variables in the design of granular-medium filters[a]

Variable	Significance
1. Required effluent quality	Usually fixed regulatory requirement
2. Influent wastewater characteristics	Affect the removal characteristics of a given filter-bed configuration. To a limited extent the listed influent characteristics can be controlled by the designer
a. Suspended-solids concentration	
b. Floc or particle size and distribution	
c. Floc strength	
d. Floc or particle charge	
e. Fluid properties	
3. Filter-medium characteristics	Affect particle-removal efficiency and headloss buildup
a. Effective size, d_{10}	
b. Uniformity coefficient, UC	
c. Type, grain shape, density, and composition	
4. Filter-bed characteristics	Porosity affects the amount of solids that can be stored within the filter. Bed depth affects initial headloss, length of run. Degree of intermixing will affect performance of filter bed
a. Bed depth	
b. Stratification	
c. Degree of medium intermixing	
d. Porosity	
5. Filtration rate	Used in conjunction with variables 2, 3, and 4 to compute clear-water headloss
6. Chemical usage	
7. Allowable headloss	Design variable
8. Backwash requirements	Affects size of filter piping and pipe gallery

[a] Adapted in part from Tchobanoglous and Eliassen (1970) and Tchobanoglous and Schroeder (1985).

The degree of intermixing in the dual-medium and multimedium beds depends on the density and size differences of the various media. To avoid extensive intermixing the settling rate of the filter mediums comprising the dual- and multimedium filters must have essentially the same settling velocity. The following relationship can be used to establish the appropriate sizes (Kawamura, 2000).

$$\frac{d_1}{d_2} = \left(\frac{\rho_2 - \rho_w}{\rho_1 - \rho_w}\right)^{0.667} \tag{11-34}$$

where d_1, d_2 = effective size of filter mediums
ρ_1, ρ_2 = density of filter mediums
ρ_w = density of water

The application of Eq. (11-34) is illustrated in Example 11-4.

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Table 11–8
Typical design data for depth filters with mono-medium[a]

Characteristic	Unit	Value: Range	Value: Typical
Shallow-bed (stratified)			
Anthracite			
Depth	mm	300–500	400
Effective size	mm	0.8–1.5	1.3
Uniformity coefficient	unitless	1.3–1.8	≤1.5
Filtration rate	$L/m^2 \cdot min$	80–240	120
Sand			
Depth	mm	300–360	330
Effective size	mm	0.45–0.65	0.45
Uniformity coefficient	unitless	1.2–1.6	≤1.5
Filtration rate	$L/m^2 \cdot min$	80–240	120
Conventional (stratified)			
Anthracite			
Depth	mm	600–900	750
Effective size	mm	0.8–2.0	1.3
Uniformity coefficient	unitless	1.3–1.8	≤1.5
Filtration rate	$L/m^2 \cdot min$	80–400	160
Sand			
Depth	mm	500–750	600
Effective size	mm	0.4–0.8	0.65
Uniformity coefficient	unitless	1.2–1.6	≤1.5
Filtration rate	$L/m^2 \cdot min$	80–240	120
Deep-bed (unstratified)			
Anthracite			
Depth	mm	900–2100	1500
Effective size	mm	2–4	2.7
Uniformity coefficient	unitless	1.3–1.8	≤1.5
Filtration rate	$L/m^2 \cdot min$	80–400	200
Sand			
Depth	mm	900–1800	1200
Effective size	mm	2–3	2.5
Uniformity coefficient	unitless	1.2–1.6	≤1.5
Filtration rate	$L/m^2 \cdot min$	80–400	200
Fuzzy filter			
Depth	mm	600–1080	800
Effective size	mm	25–30	28
Uniformity coefficient	unitless	1.1–1.2	1.1
Filtration rate	$L/m^2 \cdot min$	600–1000	800

[a] Adapted from Tchobanoglous (1988).

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Table 11-9 Typical design data for dual- and multimedium depth filters[a]

Characteristic	Unit	Value[b] Range	Typical
Dual-medium			
Anthracite (ρ = 1.60)			
Depth	mm	360–900	720
Effective size	mm	0.8–2.0	1.3
Uniformity coefficient	unitless	1.3–1.6	≤1.5
Sand (ρ = 2.65)			
Depth	mm	180–360	360
Effective size	mm	0.4–0.8	0.65
Uniformity coefficient	unitless	1.2–1.6	≤1.5
Filtration rate	L/m²·min	80–400	200
Multimedium			
Anthracite (top layer of quad-media filter, ρ = 1.60)			
Depth	mm	240–600	480
Effective size	mm	1.3–2.0	1.6
Uniformity coefficient	unitless	1.3–1.6	≤1.5
Anthracite (second layer of quad-media filter, ρ = 1.60)			
Depth	mm	120–480	240
Effective size	mm	1.0–1.6	1.1
Uniformity coefficient	unitless	1.5–1.8	1.5
Anthracite (top layer of tri-media filter, ρ = 1.60)			
Depth	mm	240–600	480
Effective size	mm	1.0–2.0	1.4
Uniformity coefficient	unitless	1.4–1.8	≤1.5
Sand (ρ = 2.65)			
Depth	mm	240–480	300
Effective size	mm	0.4–0.8	0.5
Uniformity coefficient	unitless	1.3–1.8	≤1.5
Garnet (ρ = 4.2)			
Depth	m	50–150	100
Effective size	mm	0.2–0.6	0.35
Uniformity coefficient	unitless	1.5–1.8	≤1.5
Filtration rate	L/m²·min	80–400	200

[a] Adapted from Tchobanoglous (1988).

[b] Anthracite, sand, and garnet sizes selected to limit the degree of intermixing. Use Eq. (11-34) for other values of density ρ.

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Figure 11–23

Typical particle-size distribution ranges for sand and anthracite used in dual-medium depth filters. Note that for sand the 10 percent size by weight corresponds to the 50 percent size by count.

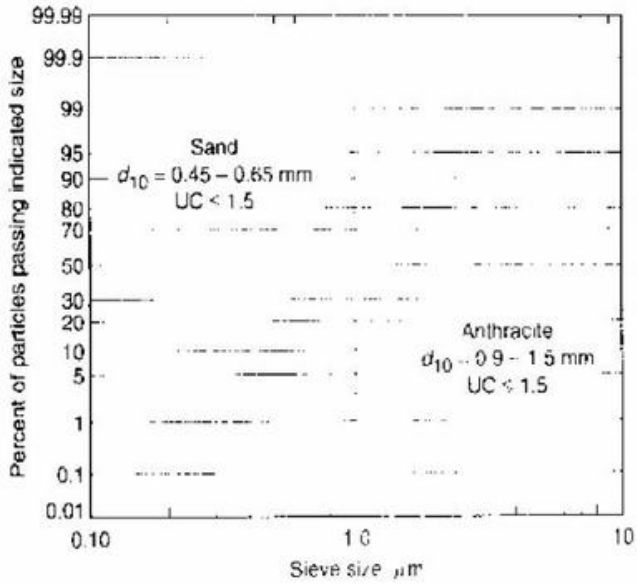

Table 11–10
Typical properties of filter materials used in depth filtration[a]

Filter material	Specific gravity	Porosity, α	Sphericity[b]
Anthracite	1.4–1.75	0.56–0.60	0.40–0.60
Sand	2.55–2.65	0.40–0.46	0.75–0.85
Garnet	3.8–4.3	0.42–0.55	0.60–0.80
Ilmenite	4.5	0.40–0.55	
Fuzzy filter medium		0.87–0.89	

[a] Adapted in part from Cleasby and Logsdon (1999).
[b] Sphericity is defined as ratio of the surface area of an equal volume sphere to the surface area of the filter medium particle.

EXAMPLE 11–4 **Determination of Filter-Medium Sizes** A dual-medium filter bed composed of sand and anthracite is to be used for the filtration of settled secondary effluent. If the effective size of the sand in the dual-medium filter is to be 0.55 mm, determine the effective size of the anthracite to avoid significant intermixing.

Solution

1. Summarize the properties of the filter mediums
   a. For sand
      i. Effective size = 0.55 mm
      ii. Specific gravity = 2.65 (see Table 11–10)
   b. For anthracite
      i. Effective size = to be determined, mm
      ii. Specific gravity = 1.7 (see Table 11–10)

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2. Compute the effective size of the anthracite using Eq. (11–34)

$$d_1 = d_2\left(\frac{\rho_2 - \rho_w}{\rho_1 - \rho_w}\right)^{0.667}$$

$$d_1 = 0.55\ \text{mm}\left(\frac{2.65 - 1}{1.7 - 1}\right)^{0.667}$$

$$d_1 = 0.97\ \text{mm}$$

Comment Another approach that can be used to assess whether intermixing will occur is to compare the fluidized bulk densities of the two adjacent layers (e.g., upper 450 mm of sand and lower 100 mm of anthracite).

Filter Flowrate Control

The principal methods now used to control the rate of flow through downflow gravity filters may be classified as (1) constant-rate filtration with fixed head, (2) constant-rate filtration with variable head, and (3) variable declining-rate filtration. A variety of other control methods are also in use (Cleasby and Logsdon, 1999; Kawumura, 2000).

Constant-Rate Filtration with Fixed Head. In constant-rate filtration with fixed head (see Fig. 11-24*a*), the flow through the filter is maintained at a constant rate. Constant-rate filtration systems are either influent controlled or effluent controlled. Pumps or weirs are used for influent control whereas an effluent modulating valve that can be operated manually or mechanically is used for effluent control. In effluent control systems, at the beginning of the run, a large portion of the available driving force is dissipated at the valve, which is almost closed. The valve is opened as the headloss builds up within the filter during the run. Because the required control valves are expensive and because they have malfunctioned on a number of occasions, alternative methods of flowrate control involving pumps and weirs have been developed and are coming into wider use.

Constant-Rate Filtration with Variable Head. In constant-rate variable filtration head (see Fig. 11-24*b*), the flow through the filter is maintained at a constant rate. Pumps or weirs are used for influent control. When the head or effluent turbidity reaches a preset value, the filter is backwashed.

Variable-Rate Filtration with Fixed or Variable Head. In variable-declining-rate filtration (see Fig. 11-24*c*), the rate of flow through the filter is allowed to decline as the rate of headloss builds up with time. Declining-rate filtration systems are either influent controlled or effluent controlled. When the rate of flow is reduced to the minimum design rate, the filter is removed from service and backwashed.

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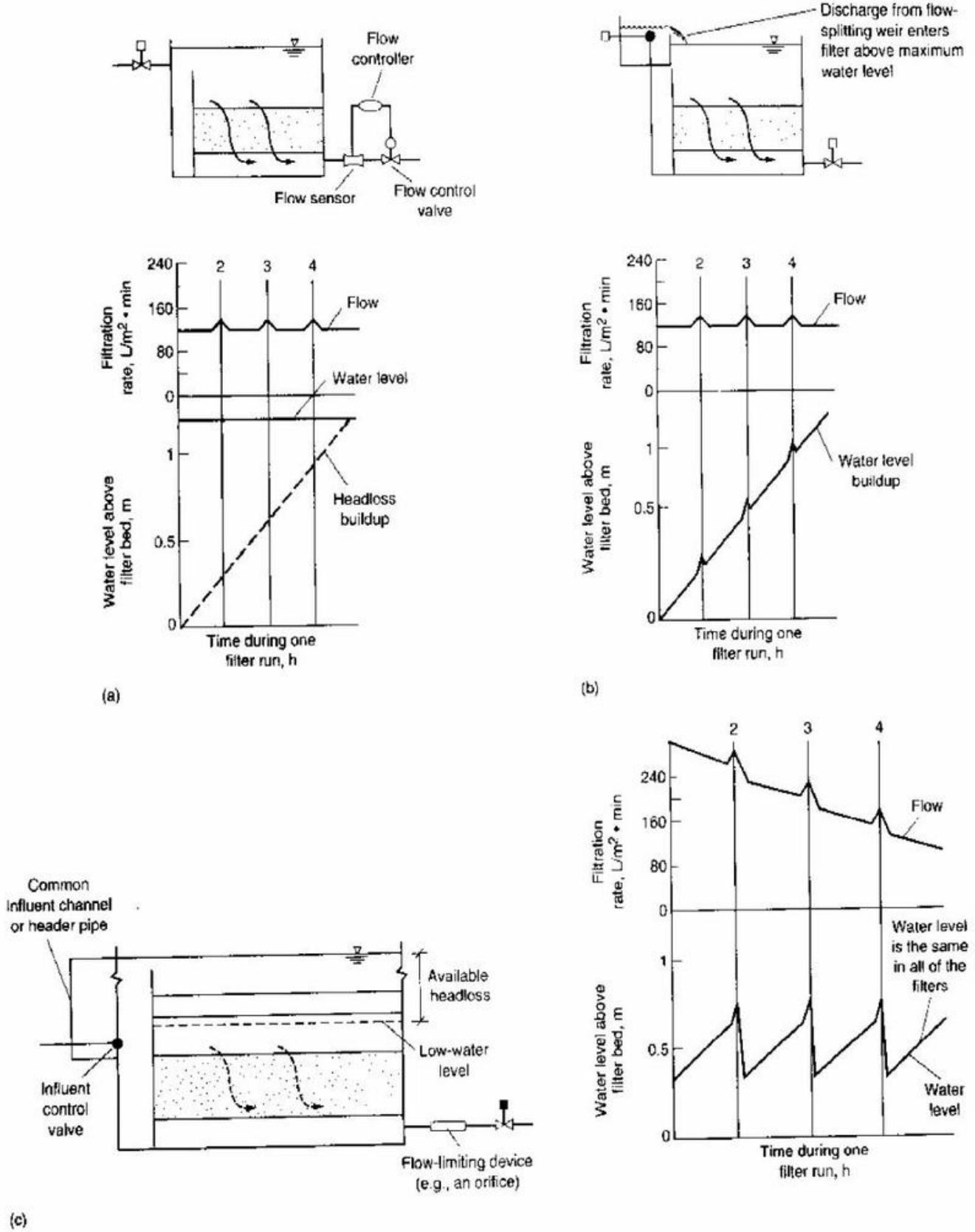

Figure 11-24

Definition sketch for filter operation: (a) fixed head, (b) variable head, and62 (c) variable-flow variable-head. Curves for filters in (a), (b), and (c) are for the operation of one filter in a bank of four filters. The numbers represent the filter that is backwashing during the filter run. In practice, the time before backwashing will not be the same for all of the filters. *(Adapted from Tchobanoglous and Schroeder, 1985.)*

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Filter Backwashing Systems

A filter bed can function properly only if the backwashing system cleans the material removed within the filter effectively. The methods commonly used for backwashing granular-medium filter beds operated in the semicontinuous mode include (1) water backwash with auxiliary surface water-wash agitation, (2) water backwash with auxiliary air scour, and (3) combined air-water backwashing. With the first two methods, fluidization of the granular medium is necessary to achieve effective cleaning of the filter bed at the end of the run. With the third method, fluidization is not necessary. Typical backwash flowrates required to fluidize various filter beds are reported in Table 11–11.

Water Backwash with Auxiliary Surface Wash. Surface washers (see Fig. 11–25) are often used to provide the shearing force required to clean the grains of the filtering medium used for wastewater filtration. Operationally, the surface washing cycle is started about 1 or 2 min before the water backwashing cycle is started. Both cycles are continued for about 2 min, at which time the surface wash is terminated. Water usage is as follows: for a single-sweep surface backwashing system, from 20 to 40 L/m^2·min (0.5 to 1.0 gal/ft^2·min); for a dual-sweep surface backwashing system, from 60 to 80 L/m^2·min (1.5 to 2.0 gal/ft^2·min).

Water Backwash with Auxiliary Air Scour. The use of air to scour the filter provides a more vigorous washing action than water alone. Operationally, air is usually applied for 3 to 4 min before the water backwashing cycle begins. In some systems, air is also injected during the first part of the water-washing cycle. Typical air flowrates range from 0.9 to 1.5 m^3/ m^2·min (3 to 5 ft^3/ft^2·min).

Combined Air-Water Backwash. The combined air-water backwash system is used in conjunction with the single-medium unstratified filter bed. Operationally, air and water are applied simultaneously for several minutes. The specific duration of the combined backwash varies with the design of the filter bed. Ideally, during the backwash operation, the filter bed should be agitated sufficiently so that the grains of the filter

Table 11–11 Typical backwash flowrates required to fluidize various filter beds

Type of filter	Size of critical granular medium	Minimum backwash velocity needed to fluidize bed[a]	
		m^3/m^2·min	m/h
Single medium (sand)	2 mm	1.8–2.0	110–120
Dual-media (anthracite and sand)	See Table 11–9	0.8–1.2	48–72
Tri-media (anthracite, sand, and garnet or ilmenite)	See Table 11–9	0.8–1.2	48–72
Fuzzy filter	30 mm	0.4–0.6	24–36

[a] Varies with size, shape, and specific gravity of the filter medium and the temperature of the backwash water.

Note: m^3/m^2·min × 24.5424 = gal/ft^2·min.

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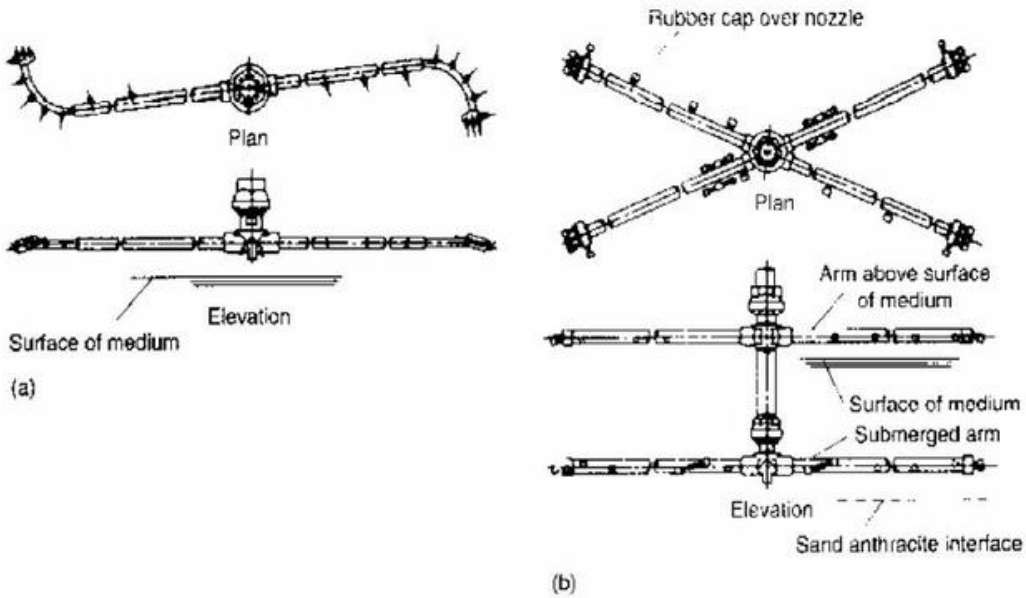

Figure 11-25

Surface washing facilities used to clean conventional granular-medium filters: (*a*) single arm agitator and (*b*) dual arm agitator.

Table 11-12
Typical air and water backwash rates used with single-medium sand and anthracite filters[a]

Medium	Medium characteristics: Effective size, mm	Medium characteristics: Uniformity coefficient	Backwash rate: Water, $m^3/m^2 \cdot min$[b]	Backwash rate: Air, $m^3/m^2 \cdot min$[c]
Sand	1.00	1.40	0.41	13.1
	1.49	1.40	0.61	19.7
	2.19	1.30	0.81	26.2
Anthracite	1.10	1.73	0.28	6.6
	1.34	1.49	0.41	13.1
	2.00	1.53	0.61	19.7

[a] Adapted from Dehab and Young (1977).
[b] In a combined air-water backwash process, the water backwash rates are typically less than 50 percent of the rates required to fluidize the bed.
[c] Air at 21°C (70°F) and 1.0 atm.
Note: $m^3/m^2 \cdot min \times 24.5424 = gal/ft^2 \cdot min$
$m^3/m^2 \cdot min \times 3.2808 = ft^3/ft^2 \cdot min$

medium move in a circular pattern from the top to the bottom of the filter as the air and water rise up through the bed. Some typical data on the quantity of water and air required are reported in Table 11–12. The reduced washwater requirements for the air-water backwash system can be appreciated by comparing the values given in Table 11–11 with those given in Table 11–12. At the end of the combined air-water backwash, a 2- to 3-min water backwash at subfluidization velocities is used to remove any air bubbles that

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may remain in the filter bed. This last step is required to eliminate the possibility of air binding within the filter.

Filter Appurtenances

The principal filter appurtenances are as follows: (1) the underdrain system used to support the filtering materials, collect the filtered effluent, and distribute the backwash water and air (where used); (2) the washwater troughs used to remove the spent backwash water from the filter; and (3) the surface washing systems used to help remove attached material from the filter medium.

Underdrain Systems. The type of underdrain system to be used depends on the type of backwash system. In conventional water-backwashed filters without air scour, it is common practice to place the filtering medium on a support consisting of several layers of graded gravel. The design of a gravel support for a granular medium is delineated in the AWWA Standard for Filtering Material B100-96. Typical underdrain systems are shown on Fig. 11–26.

Washwater Troughs. Washwater troughs are constructed of fiberglass, plastic, or sheet metal or of concrete with adjustable weir plates. The particular design of the trough will depend to some extent on the other equipment to be used in the design and construction of the filter. Loss of filter material during backwashing is a common operating problem. To reduce this problem, baffles can be placed on the underside of the washwater troughs as shown on Fig. 11–27.

Surface Washers. Surface washers for filters can be fixed or mounted on rotary sweeps, as shown on Fig. 11–25. According to data on a number of systems, rotary-sweep washers appear to be the most effective.

Filter Instrumentation and Control Systems

The supervisory control facilities for wastewater filtration include instrumentation systems for the control and monitoring of the filters. The control systems are similar to those used for water treatment; however, full automation of gravity wastewater filters is not required. Although full automation is not required, fully automatic control systems using programmed logic controllers (PLCs) are used routinely.

Flow through the filters may be controlled from a water level upstream of the filters or from the water level in each filter. These water levels are used in conjunction with rate-of-flow controllers or a control valve to limit or regulate the flowrate through a filter. Filter hydraulic operating parameters requiring monitoring include filtered water flowrate, total headloss across each filter, surface wash and backwash water flowrates, and air flowrate if an air/water backwash system is employed. Water-quality parameters in filtered water that are monitored usually include BOD, TSS, phosphorus, and nitrogen. Turbidity may also be monitored in systems where chemical addition is used. Signals from effluent turbidity monitors and effluent flowrate are often used to pace the chemical-feed system.

The sequencing of the backwash cycle for a conventional gravity filter should be semiautomatic preferably, incorporating manual start, followed by automatic operation to carry the backwash cycle through its various steps. The design of backwash systems

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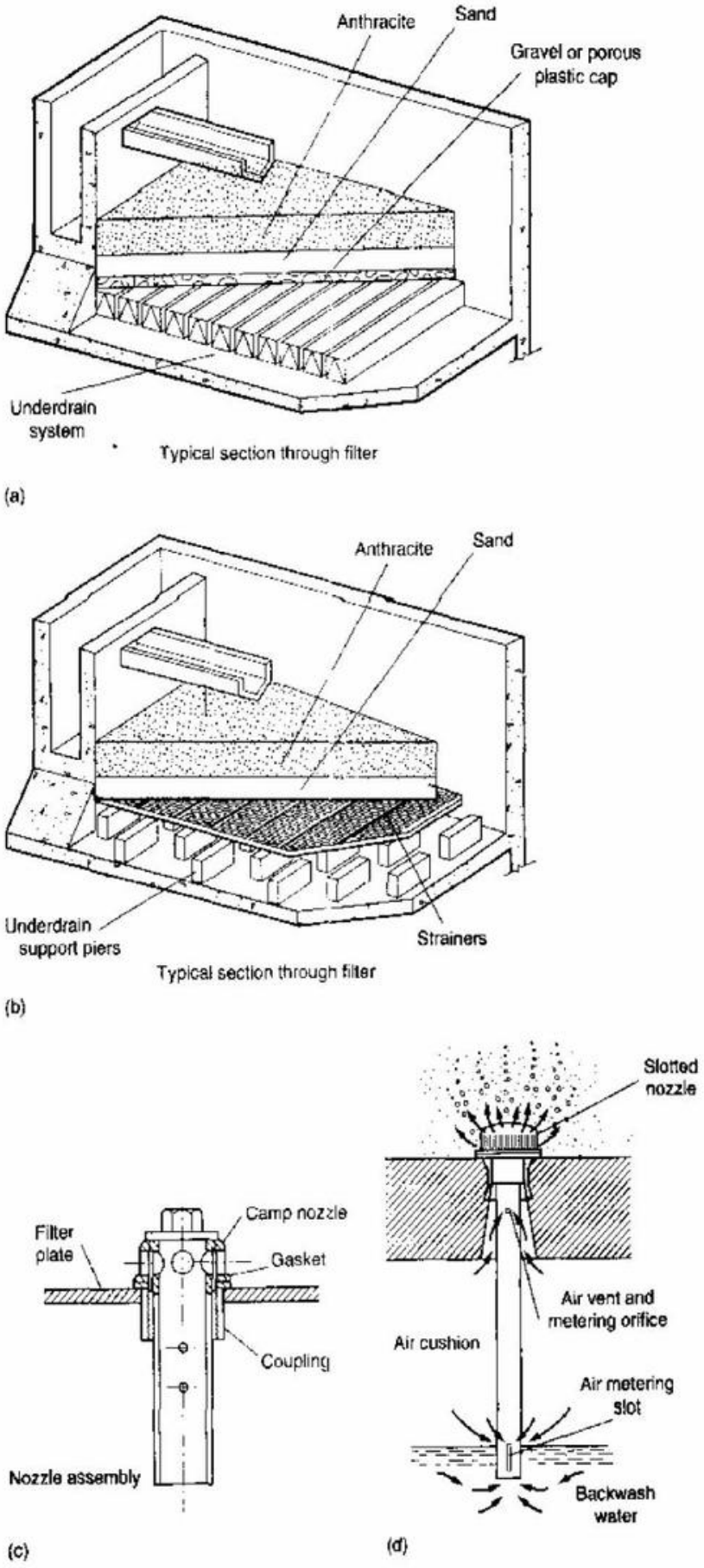
Figure 11-26

Typical underdrain systems used for wastewater filters: (a) underdrain system used with gravel or porous plastic cap support layer for filter media *(Courtesy F. B. Leopold Co., Inc.)*, (b) underdrain system used without gravel support layer, and (c, d) air-water nozzle used in filters without gravel support layer. *(Courtesy Walker Process and Infilco-Degremont, Inc., respectively.)*



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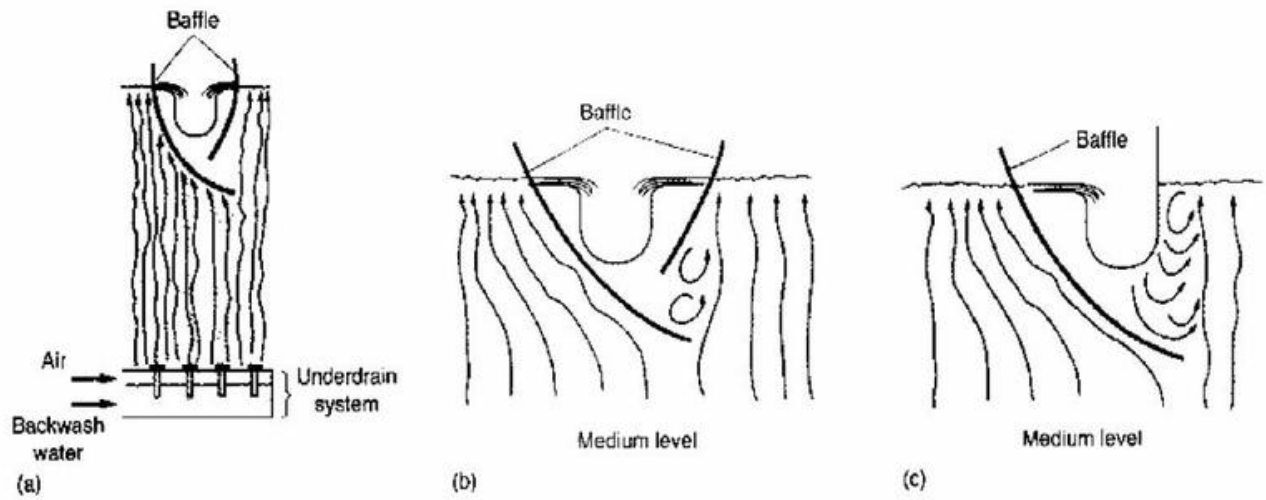

Figure 11-27

Details of baffle system developed for use with single-medium unstratified filter bed to minimize loss of filtering medium during backwash operation. (a) Section through filter showing flow patterns during air-water backwash operation. (b) Dual-baffle detail. (c) Single-baffle detail.

must recognize the impact of maximum wastewater temperatures experienced at treatment plants. Local control units should be provided at the filters to allow for local operation and backwashing by plant operators.

Effluent Filtration with Chemical Addition

Depending on the quality of the settled secondary effluent, chemical addition has been used to improve the performance of effluent filters. Chemical addition has also been used to achieve specific treatment objectives including the removal of specific contaminants such as phosphorus, metal ions, and humic substances. The removal of phosphorus by chemical addition is considered in Chap. 6. To control eutrophication, the contact filtration process is used in many parts of the country to remove phosphorus from wastewater-treatment plant effluents which are discharged to sensitive water bodies. The two-stage filtration process discussed previously (see Fig. 11-17) has proved to be very effective, achieving phosphorus levels of 0.2 mg/L in the filtered effluent. Chemicals commonly used in effluent filtration include a variety of organic polymers, alum, and ferric chloride. Use of organic polymers and the effects of the chemical characteristics of the wastewater on alum addition are considered in the following discussion.

Use of Organic Polymers. Organic polymers are typically classified as long-chain organic molecules with molecular weights varying from 10^4 to 10^6. With respect to charge, organic polymers can be cationic (positively charged), anionic (negatively charged), or nonionic (no charge). Polymers are added to settled effluent to bring about the formation of larger particles by bridging as described in Chap. 6. Because the chemistry of the wastewater has a significant effect on the performance of a polymer, the selection of a given type of polymer for use as a filter aid generally requires experimental testing. Common test procedures for polymers involve adding an initial dosage (usually 1.0 mg/L) of a given polymer and observing the effects. Depending upon the effects observed, the dosage should be increased by 0.5 mg/L increments or decreased

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by 0.25 mg/L increments (with accompanying observation of effects) to obtain an operating range. After the operating range is established, additional testing can be done to establish the optimum dosage.

A recent development is the use of lower-molecular-weight polymers that are intended to serve as alum substitutes. When these polymers are used, the dosage is considerably higher (≥10 mg/L) than with higher-molecular-weight polymers (0.25 to 1.25 mg/L). As with the mixing of alum, the initial mixing step is critical in achieving maximum effectiveness of a given polymer. In general, mixing times of less than 1 s with G values of >3500 s^{-1} are recommended (see Table 5–10 in Chap. 5). It should be noted that, as a practical matter, as treatment plants get larger it is difficult to achieve mixing times less than 1 s unless multiple mixing devices are used.

Effects of Chemical Characteristics of Wastewater on Alum Addition. As with polymers, the chemical characteristics of the treated wastewater effluent can have a significant impact on the effectiveness of aluminum sulfate (alum) when it is used as an aid to filtration. For example, the effectiveness of alum is dependent on pH (see Fig. 6–9 in Chap. 6). Although Fig. 6–9 was developed for water-treatment applications, it has been found to apply to most wastewater effluent filtration uses with minor variations. As shown on Fig. 6–9, the approximate regions in which the different phenomena associated with particle removal in conventional sedimentation and filtration processes are operative are plotted as a function of the alum dose and the pH of the treated effluent after alum has been added. For example, optimum particle removal by sweep floc occurs in the pH range of 7 to 8 with an alum dose of 20 to 60 mg/L. Generally, for many wastewater effluents that have high pH values (e.g., 7.3 to 8.5), low alum dosages in the range of 5 to 10 mg/L will not be effective. To operate with low alum dosages, pH control will generally be required.

Filter Problems

The principal problems encountered in wastewater filtration and the control measures that have proved to be effective are reported in Table 11–13. Because these problems can affect both the performance and operation of a filter system, care should be taken in the design phase to provide the necessary facilities that will minimize their impact. When filtering secondary effluent containing residual biological floc, semicontinuous filters should be backwashed at least once every 24 h to avoid the formation of mudballs and the buildup of grease. In most cases, the frequency of backwashing will be more often.

Need for Pilot-Plant Studies

Although the information presented earlier in this section and previously in Sec. 11–3 will help the reader understand the nature of the filtration operation as it is applied to the filtration of treated wastewater, it must be stressed that there is no generalized approach to the design of full-scale filters. The principal reason is the inherent variability in the characteristics of the influent suspended solids to be filtered. For example, changes in the degree of flocculation of the suspended solids in the secondary settling facilities will significantly affect the particle sizes and their distribution in the effluent. This in turn will affect the performance of the filter. Further, because the characteristics of the effluent suspended solids will also vary with the organic loading on the process as well as with the time of day, filters must be designed to function under a rather wide range of oper-

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Table 11-13 Summary of commonly encountered problems in depth filtration of wastewater and control measures for those problems

Problem	Description/control
Turbidity breakthrough[a]	Unacceptable levels of turbidity are recorded in the effluent from the filter, even though the terminal headloss has not been reached. To control the buildup of effluent turbidity levels, chemicals and polymers have been added to the filter. The point of chemical or polymer addition must be determined by testing
Mudball formation	Mudballs are an agglomeration of biological floc, dirt, and the filtering medium or media. If the mudballs are not removed, they will grow into large masses that often sink into the filter bed and ultimately reduce the effectiveness of the filtering and backwashing operations. The formation of mudballs can be controlled by auxiliary washing processes such as air scour or water surface wash concurrent with, or followed by, water wash.
Buildup of emulsified grease	The buildup of emulsified grease within the filter bed increases the headloss and thus reduces the length of filter run. Both air scour and water surface wash systems help control the buildup of grease. In extreme cases, it may be necessary to steam clean the bed or to install a special washing system
Development of cracks and contraction of filter bed	If the filter bed is not cleaned properly, the grains of the filter-bed filtering medium become coated. As the filter compresses, cracks develop, especially at the sidewalls of the filter. Ultimately, mudballs may develop. This problem can be controlled by adequately backwashing and scouring
Loss of filter medium or media (mechanical)	In time, some of the filter material may be lost during backwashing and through the underdrain system (where the gravel support has been upset or the underdrain system has been installed improperly). The loss of the filter material can be minimized through the proper placement of washwater troughs and underdrain systems. Special baffles have also proved effective
Loss of filter medium or media (operational)	Depending on the characteristics of the biological floc, grains of the filter material can become attached to it, forming aggregates light enough to be floated away during the backwashing operations. The problem can be minimized by the addition of an auxiliary air and/or water scouring system
Gravel mounding	Gravel mounding occurs when the various layers of the support gravel are disrupted by the application of excessive rates of flow during the backwashing operation. A gravel support with an additional 50- to 75-mm (2- to 3-in) layer of high-density material, such as ilmenite or garnet, can be used to overcome this problem

[a] Turbidity breakthrough does not occur with filters that operate continuously.

ating conditions. The best way to ensure that the filter configuration selected for a given application will function properly is to conduct pilot-plant studies (see Fig. 11-28).

Because of the many variables that can be analyzed, care must be taken not to change more than one variable at a time so as to confound the results in a statistical sense. Testing should be carried out at several intervals, ideally throughout a full year, to assess seasonal variations in the characteristics of the effluent to be filtered. All test

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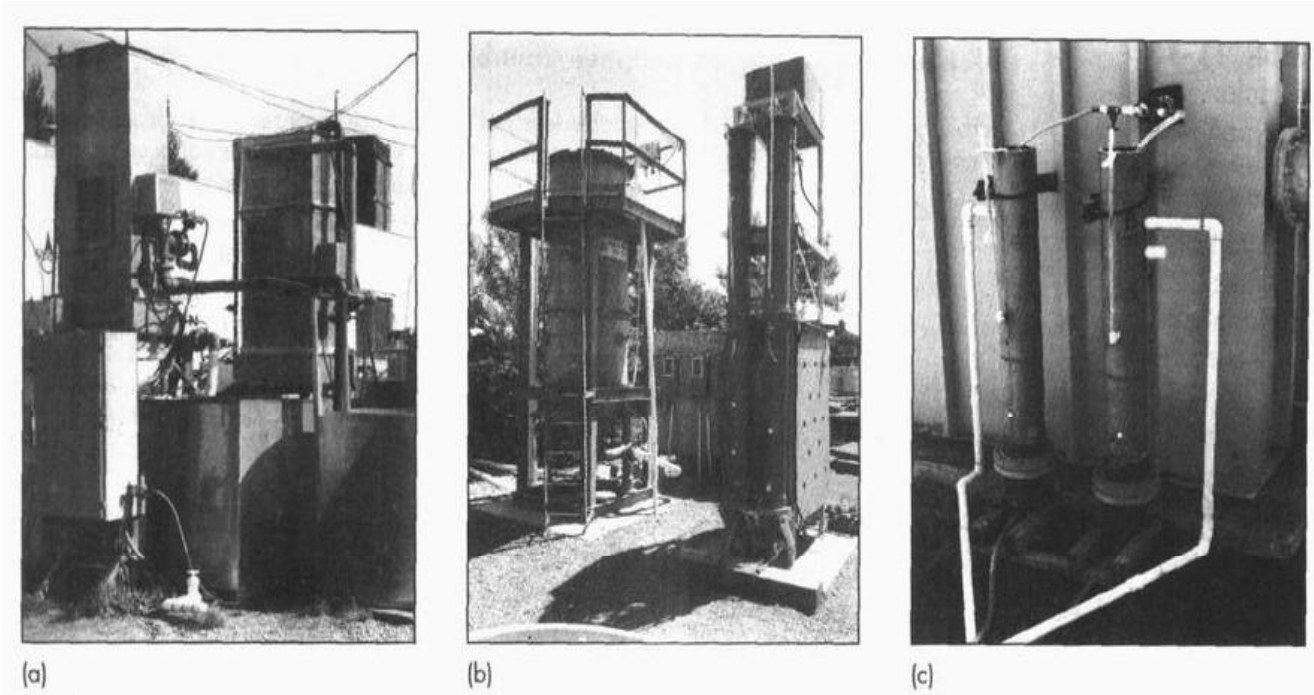

Figure 11-28

Typical pilot-plant facilities used to conduct filtration studies: (a) pulsed-bed filter, (b) continuous backwash upflow filter and dual-medium and deep bed monomedium filter columns, and (c) shallow bed monomedium.

results should be summarized and evaluated in several different ways to ensure their proper analysis. Because the specific details of each test program will be different, no generalization on the best method of analysis can be given.

11-5 SURFACE FILTRATION

Surface filtration, as shown on Fig. 11-3*b*, involves the removal of particulate material suspended in a liquid by mechanical sieving by passing the liquid through a thin septum (i.e., filter material). The mechanical sieving action is similar to a kitchen colander. Materials that have been used as filter septums include woven metal fabrics, cloth fabrics of different weaves, and a variety of synthetic materials. Membrane filters, discussed in the following section, are also surface filtration devices but are differentiated on the basis of the size of the pores in the filter medium. Cloth-medium surface filters have openings in the size range from 10 to 30 μm or larger. In membrane filters the pore size can vary from 0.0001 to 1.0 μm. Surface filtration has been used to remove the residual suspended solids from secondary effluents and from stabilization pond effluents. The focus of this section is on the removal of residual suspended solids from secondary and pond effluents discussed in the previous chapters. In this application, surface filtration is used as a replacement for depth filtration. The two principal types of cloth-medium surface filtration devices are described below.

Discfilter®

The Discfilter® (DF) consists of a series of disks composed of two vertically mounted parallel disks that are used to support the filter cloth. Each disk is connected to a cen-

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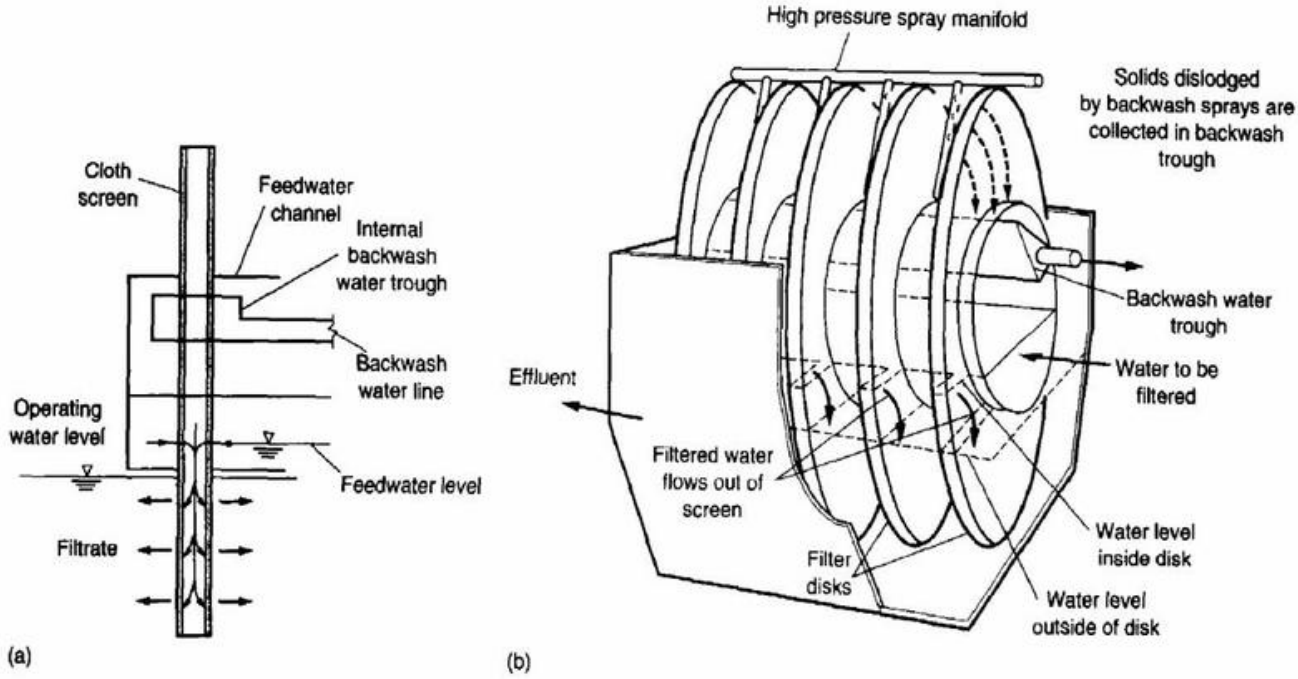

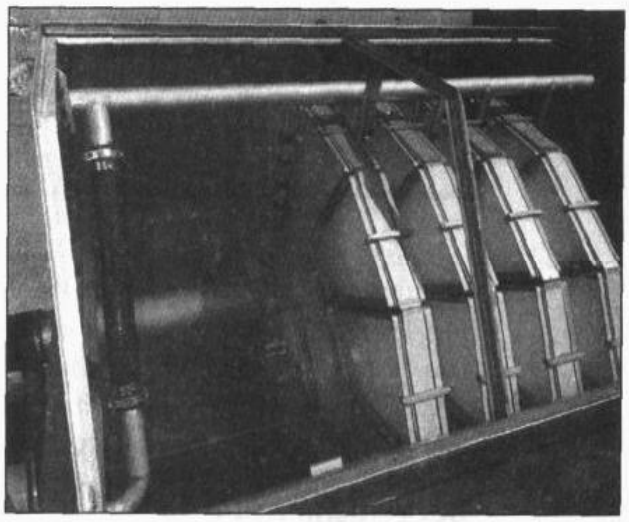

Figure 11-29

Discfilter® surface filter: (*a*) definition sketch for operation, (*b*) pictorial schematic, and (*c*) photo of pilot test unit (*Note:* disk is full size). *(Part b adapted from US Filter.)*

tral feed tube (see Fig. 11–29). The two-dimensional cloth screen material used with the DF can be of either polyester or Type 316 stainless steel. Typical design information for a DF can be found in Table 11–14.

Operation of DF. Operationally, as shown on Fig. 11–29*a*, water enters through a central channel and flows outward through the filter cloth. During normal operation 60 to 70 percent of the surface area of the DF is submerged and the disk rotates, depending on the headloss, from 1 to 8.5 r/min. The DF has the ability to operate in an intermittent or a continuous-backwash mode. When operating in a continuous-backwash mode, the disks of the DF will both produce filtered water and be backwashed simultaneously. The

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Table 11–14 Typical design information for surface filtration of secondary settled effluent using a Discfilter®

Item	Unit	Typical value	Remarks
Size of opening in screen material	μm	20–35	Two-dimensional stainless-steel or polyester screen cloths are available in sizes ranging from 10 to 60 μm
Hydraulic loading rate	$m^3/m^2 \cdot min$	0.25–0.83	Depends on characteristics of suspended solids that must be removed
Headloss through screen	mm	75–150	Based on submerged surface area of drum
Drum submergence	% height % area	70–75 60–70	Bypass should be provided when headloss exceeds 200 mm
Drum diameter	m	0.75–1.50	Varies depending on screen design; 3 m (10 ft) is most commonly used size; smaller sizes increase backwash requirements
Drum speed	m/min	4.5 at 50 mm headloss 30–40 at 150 mm headloss	Maximum rotational speed is limited
Backwash requirements	% throughput	2 at 350 kPa 5 at 100 kPa	

different stages the disk filter passes through during each rotation are identified on Fig. 11–29*b*. At the start of a rotation feedwater enters the central feed tube from where it is distributed into the disks. While the disk filter is submerged, water and particles smaller than openings in the cloth screen pass through the filter to the effluent collection channel. Those particles larger than the screen are retained.

Backwashing the DF. As the disk continues to rotate past the effluent water level, the remaining feedwater passes through the cloth screen until no water remains in the disk. The screen containing the captured wastewater solids continues to rotate until it passes the backwash spray jets. As the disk rotates past the backwash spray jets, the particles that have been retained on the cloth screen are flushed from the surface of the screen. The mixture of backwash water and solids falls from the disk and is collected in the backwash water trough. After passing by the backwash nozzles, the cleaned screens are ready to begin filtering water again. When the DF is operating in an intermittent backwash mode, the backwash spray jets are activated only when headloss through the filter reaches a preset level.

Cloth-Media Disk Filter®

The Cloth-Media Disk Filter (CMDF), as illustrated on Fig. 11–30, also consists of several disks mounted vertically in a tank. Two different types of cloths can be used in the CMDF: (1) a needle felt cloth, made of polyester or (2) a synthetic pile fabric cloth. The

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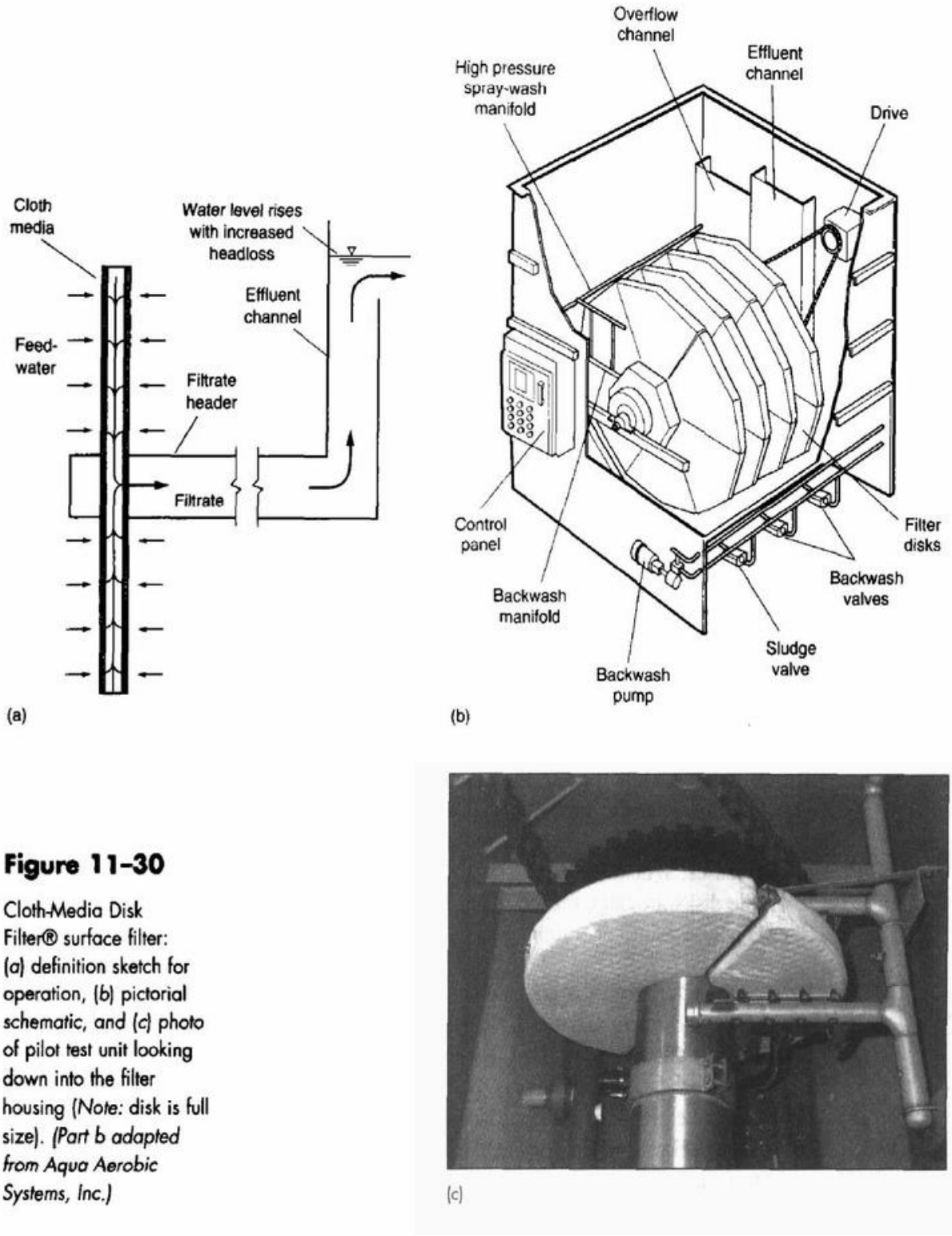

Figure 11-30

Cloth-Media Disk Filter® surface filter: (a) definition sketch for operation, (b) pictorial schematic, and (c) photo of pilot test unit looking down into the filter housing (*Note:* disk is full size). *(Part b adapted from Aqua Aerobic Systems, Inc.)*

needle felt cloth has a random three-dimensional weave to facilitate particle removal. In addition to the normal backwashing, the felt cloth must be cleaned periodically with a high-pressure spray. The pile fabric does not require the use of a high-pressure spray wash and can be cleaned completely by use of a backwash alone. Typical design information for the CMDF can be found in Table 11-15.

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Table 11–15
Typical design information for surface filtration of secondary settled effluent using a Cloth-Media Disk Filter®

Item	Unit	Typical value	Remarks
Nominal pore size	μm	10	Polyester three-dimensional needle felt cloths are employed as the filter material
Hydraulic loading rate	$m^3/m^2 \cdot min$	0.1–0.27	Depends on characteristics of suspended solids that must be removed
Headloss through screen	mm	50–300	Based on solids accumulation on or within the cloth
Disk submergence	% height % area	100 100	
Disk diameter	m	0.90 or 1.80	Two sizes are available
Disk speed	r/min	Stationary during normal operation 1 during backwash	
Backwash and sludge wasting requirements	% throughput	4.5 at 0.1 $m^3/m^2 \cdot min$ 7.2 at 0.27 $m^3/m^2 \cdot min$	Is a function of hydraulic loading rate and feedwater quality

Operation of the CMDF. Operationally, as shown on Fig. 11–30*a*, water enters the feed tank and flows through the filter cloth into a central collection tube or header. The resulting CMDF filtrate is collected in a central tube or a filtrate header (see Fig. 11–30*a*), where it flows to final discharge over an overflow weir in the effluent channel. As solids accumulate on and in the cloth medium, resistance to flow or headloss increases. When the headloss through the cloth medium reaches a predetermined level, the disks are backwashed. After filtering to waste after the backwash cycle, the filter is put back into operation.

Backwashing the CMDF. When a backwash cycle is initiated, the disks remain submerged and rotate at 1 r/min. Vacuum suction heads, located on either side of the CMDF, draw filtrate water from the filtrate header back through the cloth media into the vacuum heads while the disk is rotating. This reversal of flow removes particles that have become entrapped on the surface and within the cloth medium. The normal duration of the backwash cycle for the CMDF is 1 min.

Over time, particles will accumulate in the cloth medium that cannot be removed by a typical backwash. This accumulation of particles leads to increased headloss across the filter, an increase in the backwash suction pressure, and shorter run times between backwashes. When the backwash suction pressure reaches 124 kPa (18 lb/in^2), or the time exceeds the desired preset time interval, a high-pressure spray wash is initiated automatically. Prior to the high-pressure spray wash, the tank inlet valve is closed and the influent flow is stopped. A standard backwash is initiated to remove the outer layer of solids from the cloth. The filter-to-waste valve is opened, the water level is lowered to below the midline of the disks, and then a high-pressure spray wash is initiated.

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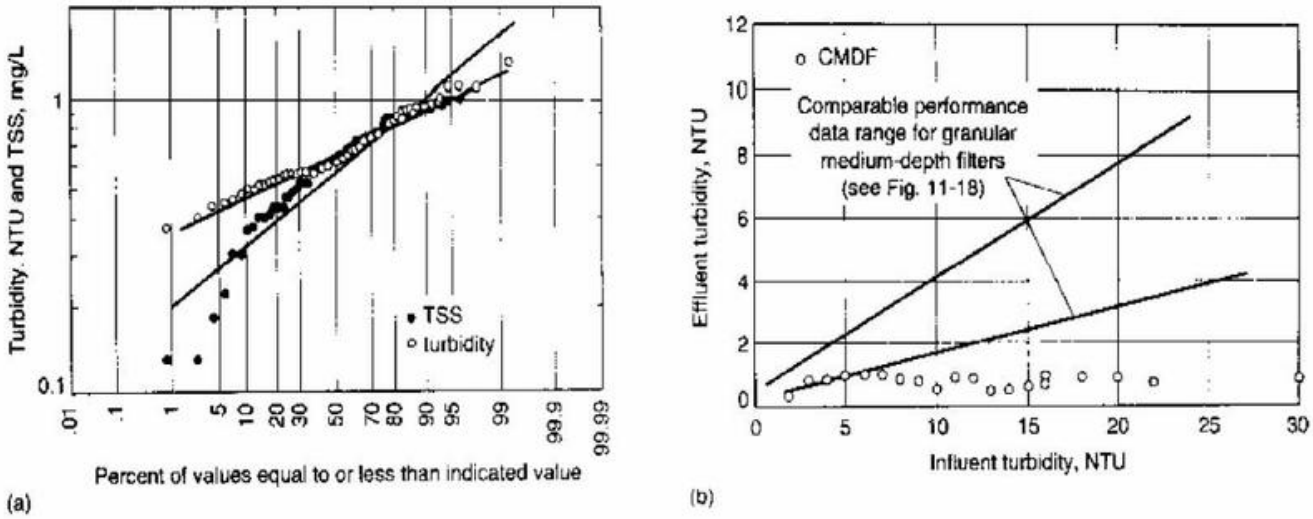

Figure 11–31

Performance data for the Cloth-Media Disk Filter® for secondary effluent: (a) effluent probability distributions for turbidity and TSS and (b) effluent turbidity as function of the influent turbidity at a filtration rate of 176 L/m²·min (4.4 gal/ft²·min).

During the high-pressure spray wash, the disks rotate slowly at 1 r/min, while filtrate water is sprayed at a high pressure from the outside of the filter cloth. The high-pressure spray wash flushes the particles that have become lodged inside the cloth filter medium effectively in two revolutions of the disk. After the end of the high-pressure spray wash, the inlet valve to the CMDF opens to allow the wastewater to flow to the filter. The disks continue to rotate and the filter-to-waste valve remains open until the solids that were flushed from the filter cloth to the filtrate side of the CMDF are removed from the filtrate header and the effluent line. The time interval between high-pressure spray washes is a function of the feedwater quality.

Performance Characteristics

To evaluate the performance capabilities of CMDF, a pilot unit employing a full-size disk (0.9 m, 3 ft, see Table 11–15) was used to treat secondary effluent from an activated-sludge process that had TSS and turbidity values ranging from 3.9 to 30 mg/L and 2 to 30 NTU, respectively. Based on a long-term study, it was found, as shown on Fig. 11–31*a*, that both the TSS and turbidity values of the filtered effluent were less than 1, 92 percent of the time (Riess et al., 2001). The performance of the CMDF as compared to depth filters all tested with effluent from a previous activated-sludge plant treating the same wastewater is shown on Fig. 11–31*b*. As shown, the effluent from the CMDF remained constant over all of the influent turbidity values tested up to 30 NTU. The DF has not been tested. Based on observations of the cloth surface, it appears that the material removed on the filter also acts as a filter (autofiltration). The percentage of backwash water used ranged between 4 and 10.

In reviewing the data presented on Fig. 11–31*b*, it is important to note the activated-sludge treatment process is operated as an extended aeration treatment process (SRT > 15 *d*). If either of the cloth filters is to be used to filter effluent from activated-sludge

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processes operated with extremely low SRT values (i.e., 1 to 2 *d*), pilot testing should be undertaken, as the characteristics of the residual solids to be filtered are significantly different.

11-6 MEMBRANE FILTRATION PROCESSES

Filtration, as defined in Secs. 11-3 and 11-5, involves the separation (removal) of particulate and colloidal matter from a liquid. In membrane filtration the range of particle sizes is extended to include dissolved constituents (typically 0.0001 to 1.0 μm). The role of the membrane, as shown on Fig. 11-32, is to serve as a selective barrier that will allow the passage of certain constituents and will retain other constituents found in the liquid (Cheryan, 1998). To introduce membrane technologies and their application, the following subjects are considered in this section: (1) membrane process terminology, (2) membrane classification, (3) membrane configurations, (4) application of membrane technologies, (5) electrodialysis, (6) the need for pilot-plant studies, and (7) the disposal of concentrated waste streams, which is considered at the end of this section.

Membrane Process Terminology

Terms commonly encountered when considering the application of membrane processes are summarized in Table 11-16. Referring to Fig. 11-32 and Table 11-16, the influent to the membrane module is known as the *feed stream* (also known as feedwater). The liquid that passes through the semipermeable membrane is known as *permeate* (also known as the product stream or permeating stream) and the liquid containing the retained constituents is known as the *concentrate* (also known as the retentate, reject, retained phase, or waste stream). The rate at which the permeate flows through the membrane is known as the rate of *flux*, typically expressed as kg/m^2·d (gal/ft^2·d).

Membrane Process Classification

Membrane processes include microfiltration (MF), ultrafiltration (UF), nanofiltration (NF), reverse osmosis (RO), dialysis, and electrodialysis (ED). Membrane processes can be classified in a number of different ways including (1) the type of material from which the membrane is made, (2) the nature of the driving force, (3) the separation mechanism, and (4) the nominal size of the separation achieved. Each of these methods

Figure 11-32

Definition sketch for a membrane process.

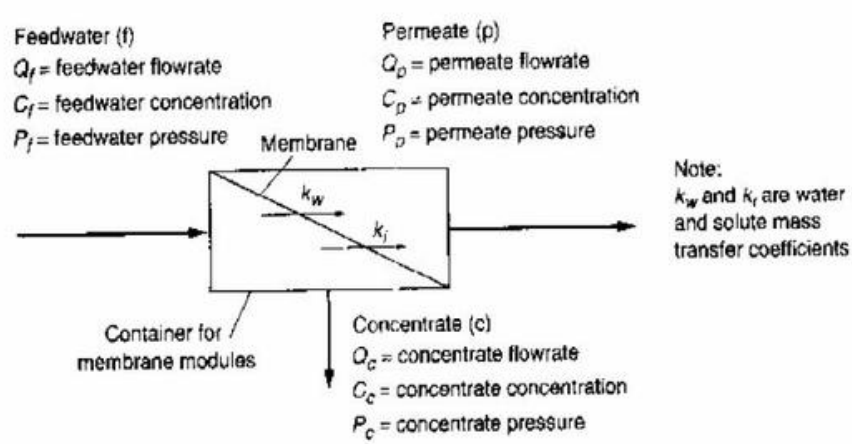

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Table 11-16 Terminology used to describe membrane processes[a]

Term	Description
Array or train	Multiple interconnected stages in series
Brine	Concentrate stream containing total dissolved solids greater than 36,000 mg/L.
Concentrate, retentate, retained phase, reject, residual stream	The portion of the feed stream that does not pass through the membrane that contains higher TDS than the feed stream
Feed stream, feedwater	Input stream to the membrane array
Flux	Mass or volume rate of transfer through the membrane surface
Fouling	Deposition of existing solid material in the element on the feed stream of the membrane. Fouling can be either reversible or irreversible
Lumen	The interior of a hollow fiber membrane
Mass transfer coefficient (MTC)	Mass or volume unit transfer through membrane based on driving force
Membrane element	A single membrane unit containing a bound group of spiral-wound or hollow fine-fiber membranes to provide a nominal surface area
Module	A complete unit comprised of the membranes, the pressure support structure for the membranes, the feed inlet and outlet permeate and retentate ports, and an overall support structure
Molecular weight cutoff (MWCO)	The molecular weight of the smallest material rejected by the membrane, usually expressed in Daltons (D)
Permeate, product, permeating stream	The portion of the feed stream that passes through the membrane that contains lower TDS than the feed stream
Reject ion	Percent solute concentration reduction of permeate stream relative to feed stream
Pressure vessel	A single tube that contains several membrane elements in series
Scaling	Precipitation of solids in the element due to solute concentration on the feed stream of the membrane
Size exclusion	Removal of particles by sieving
Solvent	Liquid containing dissolved constituents (TDS), usually water
Solute	Dissolved constituents (TDS) in raw, feed, permeate, and concentrate streams
Stage or bank	Pressure vessels arranged in parallel
System arrays	Number of arrays needed to produce the required plant flow
Train or array	Multiple interconnected stages in series

[a] Adapted in part from AWWA (1996), Cheryan (1998), and Taylor and Wiesner (1999).

of classifying membrane processes is considered in the following discussion. The general characteristics of membrane processes including typical operating ranges are reported in Table 11-17. The focus of the following discussion is on pressure-driven membrane processes. Electrodialysis is considered separately following the discussion of the application of pressure-driven membranes.

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Table 11-17
General characteristics of membrane processes

Membrane process	Membrane driving force	Typical separation mechanism	Operating structure (pore size)	Typical operating range, μm	Permeate description	Typical constituents removed
Microfiltration	Hydrostatic pressure difference	Sieve	Macropores (>50 nm)	0.08–2.0	Water + dissolved solutes	TSS, turbidity, protozoan oocysts and cysts, some bacteria and viruses
Ultrafiltration	Hydrostatic pressure difference	Sieve	Mesopores (2–50 nm)	0.005–0.2	Water + small molecules	Macromolecules, colloids, most bacteria, some viruses, proteins
Nanofiltration	Hydrostatic pressure difference	Sieve + solution/ diffusion + exclusion	Micropores (<2 nm)	0.001–0.01	Water + very small molecules, ionic solutes	Small molecules, some harness, viruses
Reverse osmosis	Hydrostatic pressure difference	Solution/ diffusion + exclusion	Dense (<2 nm)	0.0001–0.001	Water, very small molecules, ionic solutes	Very small molecules, color, hardness, sulfates, nitrate, sodium, other ions
Dialysis	Concentration difference	Diffusion	Mesopores (2–50 nm)	—	Water + small molecules	Macromolecules, colloids, most bacteria, some viruses, proteins
Electrodialysis	Electromotive force	Ion exchange with selective membranes	Micropores (<2 nm)	—	Water + ionic solutes	Ionized salt ions

Membrane Materials. Membranes used for the treatment of water and wastewater typically consist of a thin skin having a thickness of about 0.20 to 0.25 μm supported by a more porous structure of about 100 μm in thickness. Most commercial membranes are produced as flat sheets, fine hollow fibers, or in tubular form. The flat sheets are of two types, asymmetric and composite. Asymmetric membranes are cast in one process and consist of a very thin (less than 1 μm) layer and a thicker (up to 100 μm) porous layer that adds support and is capable of high water flux. Thin-film composite (TFC) membranes are made by bonding a thin cellulose acetate, polyamide, or other active layer (typically 0.15 to 0.25 μm thick) to a thicker porous substrate, which provides stability. Membranes can be made from a number of different organic and inorganic materials. The membranes used for wastewater treatment are typically organic. The principal types of membranes used include polypropylene, cellulose acetate, aromatic polyamides, and thin-film composite (TFC). The choice of membrane and system configuration is based on minimizing membrane clogging and deterioration, typically based on pilot-plant studies.

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Driving Force. The distinguishing characteristic of the first four membrane processes considered in Table 11–17 (MF, UF, NF, and RO) is the application of hydraulic pressure to bring about the desired separation. Dialysis involves the transport of constituents through a semipermeable membrane on the basis of concentration differences. Electrodialysis involves the use of an electromotive force and ion-selective membranes to accomplish the separation of charged ionic species.

Removal Mechanisms. The separation of particles in MF and UF is accomplished primarily by straining (sieving), as shown on Fig. 11–33*a*. In NF and RO, small particles are rejected by the water layer adsorbed on the surface of the membrane which is known as a *dense* membrane (see Fig. 11–33*b*). Ionic species are transported across the membrane by diffusion through the pores of the macromolecule comprising the membrane. Typically NF can be used to reject constituents as small as 0.001 μm whereas RO can reject particles as small as 0.0001 μm. Straining is also important in NF membranes, especially at the larger pore size openings.

Size of Separation. The pore sizes in membranes are identified as macropores (>50 nm), mesopores (2 to 50 nm), and micropores (< 2 nm). Because the pore sizes in RO membranes are so small, the membranes are defined as dense. The classification of membrane processes on the basis of the size of separation is shown on Fig. 11–34 and in Table 11–17. Referring to Fig.11–34, it can be seen that there is considerable overlap in the sizes of particles removed, especially between NF and RO. Nanofiltration is used most commonly in water-softening operations in place of chemical precipitation.

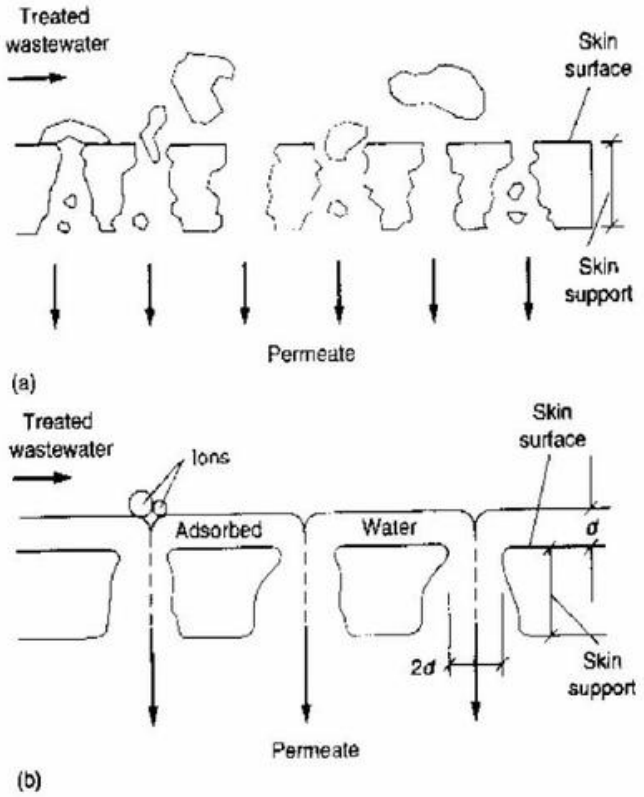

Figure 11–33

Definition sketch for the removal of wastewater constituents: (*a*) removal of large molecules and particles by sieving (size exclusion) mechanism and (*b*) rejection of ions by adsorbed water layer. *(Adapted from Tchobanoglous and Schroeder, 1985.)*

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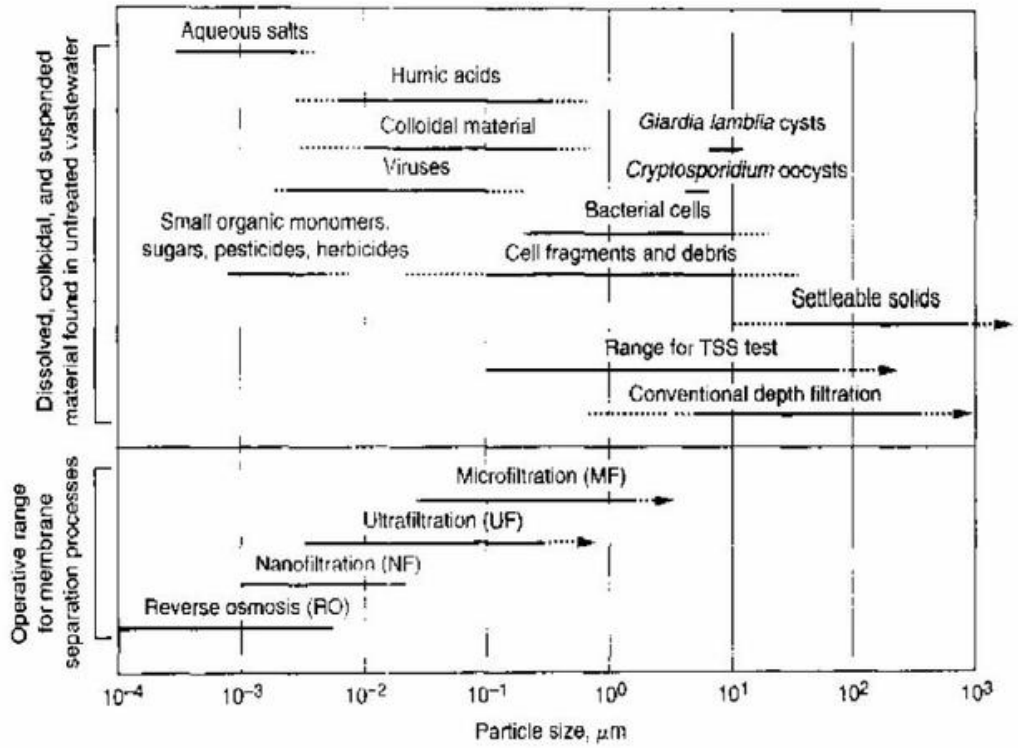
Figure 11-34

Comparison of the size of the constituents found in wastewater and the operating size ranges for membrane technologies. The operating size range for conventional depth filtration is also shown.



Membrane Configurations

In the membrane field, the term *module* is used to describe a complete unit comprised of the membranes, the pressure support structure for the membranes, the feed inlet and outlet permeate and retentate ports, and an overall support structure. The principal types of membrane modules used for wastewater treatment are (1) tubular, (2) hollow fiber, and (3) spiral wound (AWWA, 1996). Plate and frame and pleated cartridge filters are also available but are used more commonly in industrial applications. The various membranes, described below, are illustrated on Fig. 11–35.

Tubular Modules. In the tubular configuration the membrane is cast on the inside of a support tube. A number of tubes (either singly or in a bundle) are then placed in an appropriate pressure vessel. The feedwater is pumped through the feed tube and the product water is collected on the outside of the tubes (see Fig. 11–35*a*). The concentrate continues to flow through the feed tube. These units are generally used for water with high suspended solids or plugging potential. Tubular units are the easiest to clean, which is accomplished by circulating chemicals and pumping a "foamball" or "spongeball" through to mechanically wipe the membrane. Tubular units produce at a low product rate relative to their volume, and the membranes are generally expensive.

Hollow Fiber. The hollow-fiber membrane module, as shown on Fig. 11–35*b*, consists of a bundle of hundreds to thousands of hollow fibers. The entire assembly is inserted into a pressure vessel. The feed can be applied to the inside of the fiber (inside-out flow) or the outside of the fiber (outside-in flow).

Spiral Wound. In the spiral-wound membrane, a flexible permeate spacer is placed between two flat membrane sheets. The membranes are sealed on three sides.

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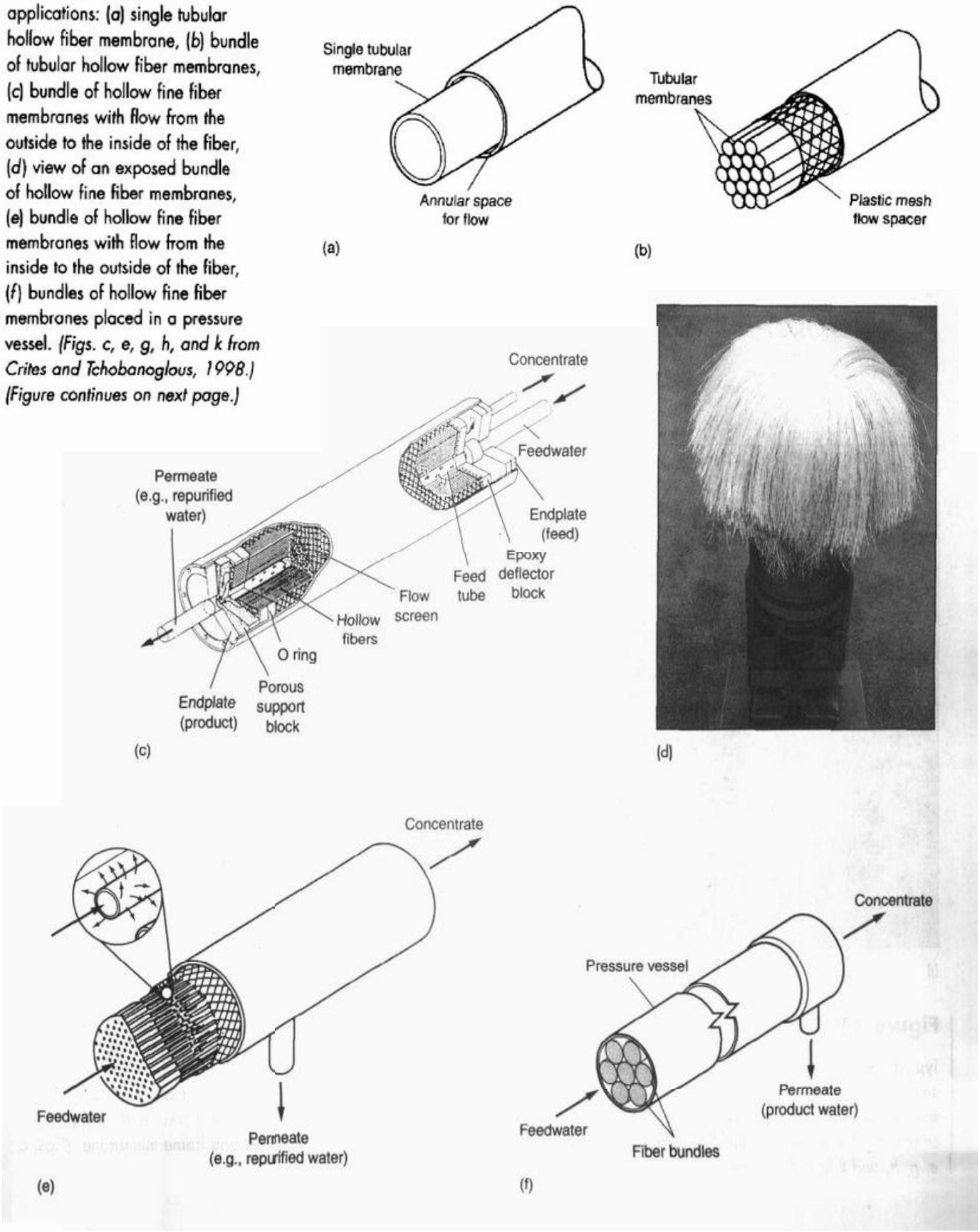
Figure 11-35

Typical membrane elements and modules used in membrane applications: (*a*) single tubular hollow fiber membrane, (*b*) bundle of tubular hollow fiber membranes, (*c*) bundle of hollow fine fiber membranes with flow from the outside to the inside of the fiber, (*d*) view of an exposed bundle of hollow fine fiber membranes, (*e*) bundle of hollow fine fiber membranes with flow from the inside to the outside of the fiber, (*f*) bundles of hollow fine fiber membranes placed in a pressure vessel. *(Figs. c, e, g, h, and k from Crites and Tchobanoglous, 1998.)* *(Figure continues on next page.)*



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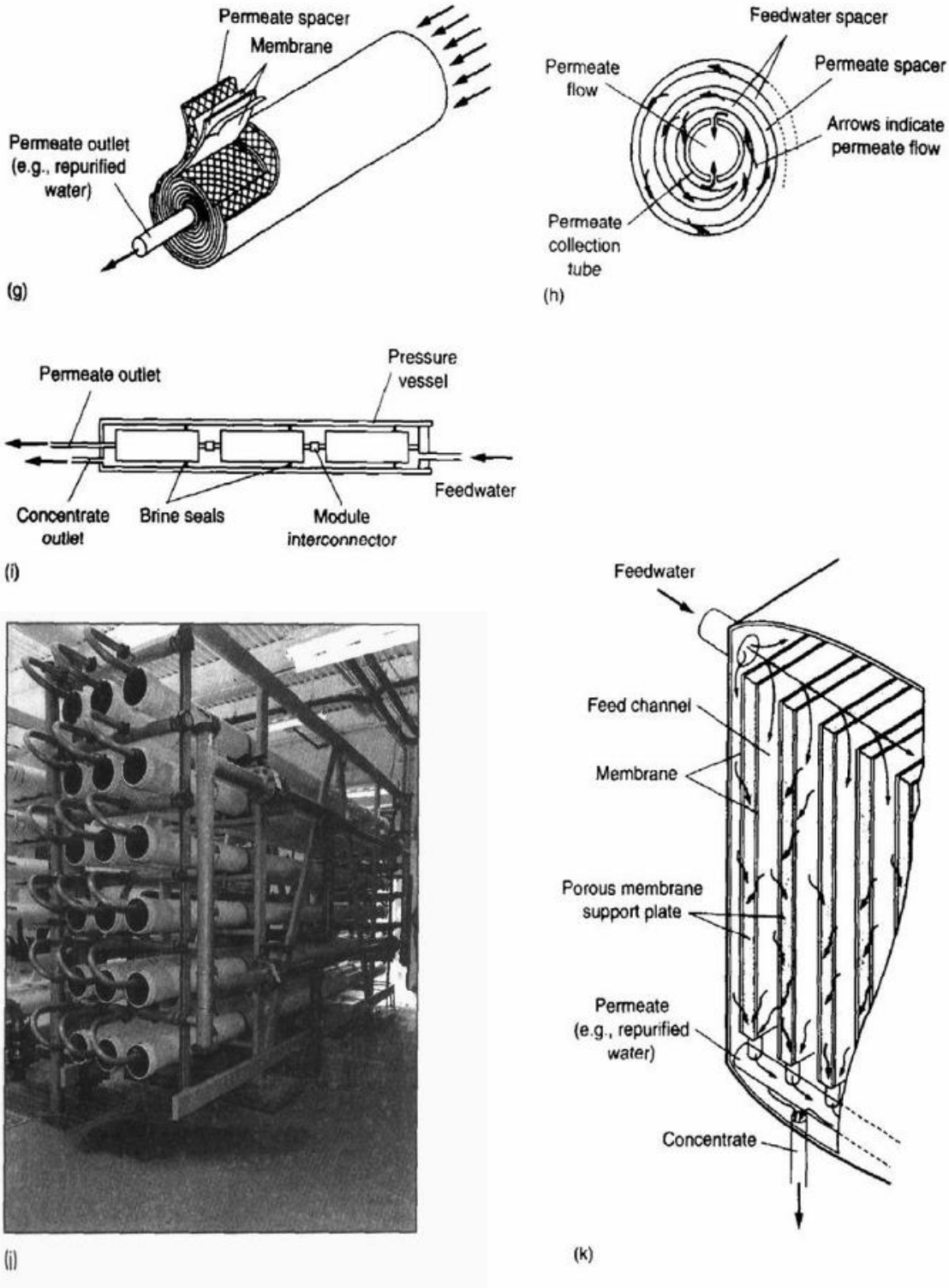

Figure 11–35 *(Continued)*

Typical membrane elements and modules used in membrane applications: (*g*) cutaway of spiral-wound thin-film composite membrane module, (*h*) section through spiral-wound thin-film composite membrane module, (*i*) pressure vessel containing 3 spiral-wound thin-film composite membrane modules in series, (*j*) typical installation of reverse osmosis pressure vessels containing 6 spiral-wound thin-film composite membrane modules in series, and (*k*) parallel plate and frame membrane. *(Figs. c, e, g, h, and k from Crites and Tchobanoglous, 1998.)*

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The open side is attached to a perforated pipe. A flexible feed spacer is added and the flat sheets are rolled into a tight circular configuration (see Figs. 11–35*g* and *h*). Thin-film composites are used most commonly in spiral-wound membrane modules. The term spiral derives from the fact that the flow in the rolled-up arrangement of membranes and support sheets follows a spiral flow pattern.

Plate and Frame. Plate and frame membrane modules (see Fig. 11–35*k*) are comprised of a series of flat membrane sheets and support plates. The water to be treated passes between the membranes of two adjacent membrane assemblies. The plate supports the membranes and provides a channel for the permeate to flow out of the unit. The plate and frame configuration is used most commonly for electrodialysis modules (sometimes identified as stacks).

Pleated-Cartridge Filters. Pleated-cartridge filters are used most commonly in microfiltration applications and usually are designed as disposable units. Pleated-cartridge filters are used almost exclusively to concentrate virus from treated wastewater for analysis (see Chap. 2).

Pressure Vessels. With the exception of the pressure vessels, which need to be manufactured specially, most of the equipment used for membrane filtration is standard to the chemical and process industry. The primary purpose of the pressure vessel (or tube) is to support the membrane and keep the feedwater and product streams isolated. The vessel must also be designed to prevent leaks and pressure losses to the outside, minimize the buildup of salt or fouling, and permit easy replacement of the membranes. Depending on the operating pressure and the characteristic of the feedwater, a variety of materials have been used, including plastic and fiberglass tubes and plumbing components. Steel pressure tubes are required for some reverse-osmosis applications, and stainless steel is required for seawater and brackish water having high TDS. Centrifugal pumps can be used for MF, UF, and NF; positive-displacement pumps or high-pressure turbine pumps are necessary for RO.

Membrane Operation

The operation of membrane processes is quite simple. A pump is used to pressurize the feed solution and to circulate it through the module. A valve is used to maintain the pressure of retentate. The permeate is withdrawn, typically at atmospheric pressure. As constituents in the feedwater accumulate on the membranes (often termed membrane fouling), the pressure builds up on the feed side, the membrane flux (i.e., flow through membrane) starts to decrease, and the percent rejection also starts to decrease (see Fig. 11–36). When the performance has deteriorated to a given level, the membrane modules are taken out of service and backwashed and/or cleaned chemically. The operational configurations and parameters for the various membrane processes are considered in the following discussion.

Microfiltration and Ultrafiltration. Three different process configurations are used with microfiltration and ultrafiltration units as illustrated on Fig. 11–37. In the first configuration known as *cross flow* (Fig. 11–37*a*) the feedwater is pumped with cross flow tangential to the membrane. Water that does not pass through the membrane

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Figure 11–36

Definition sketch for the performance of a membrane filtration system.

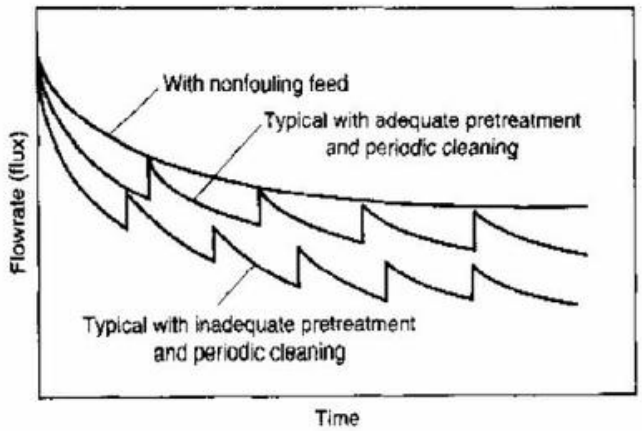
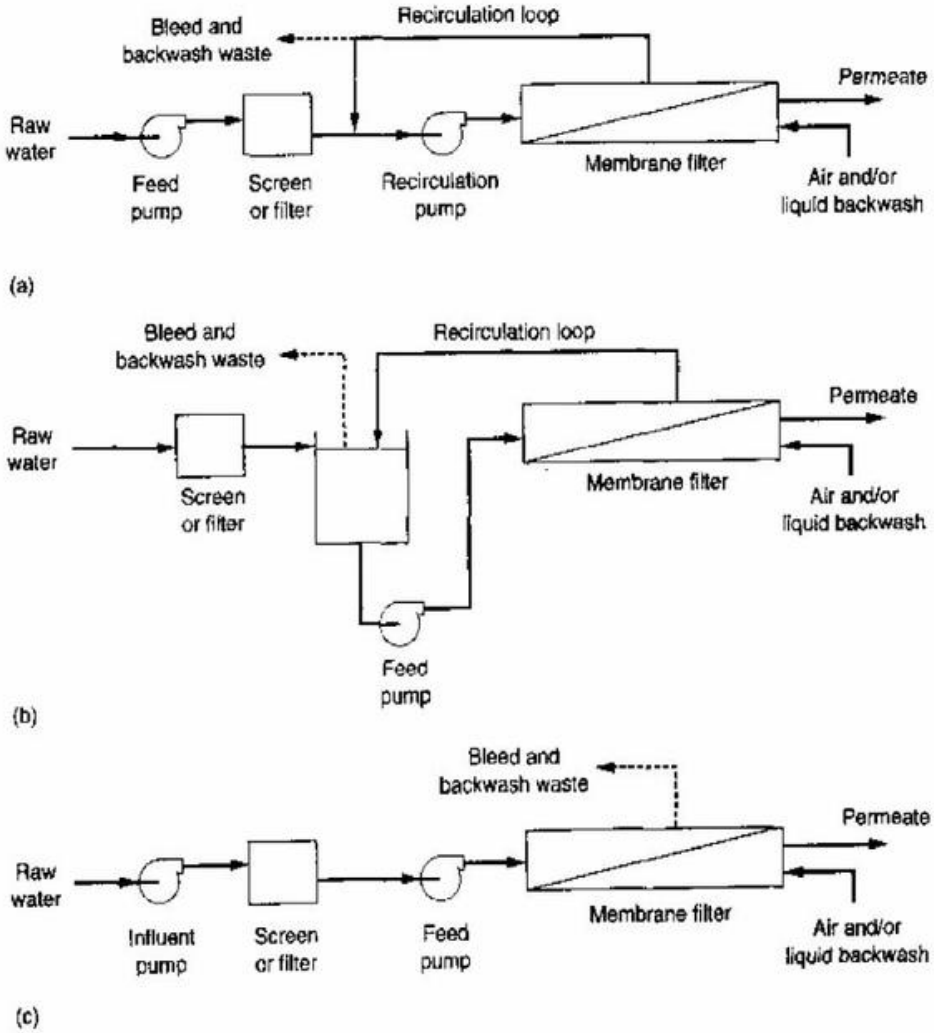


Figure 11–37

Typical operational modes for MF and UF membrane processes: (a) cross flow, (b) cross flow with reservoir, and (c) direct feed. *(Adapted from Jacangelo and Buckley, 1996.)*



is recirculated through the membrane after blending with additional feedwater. The second configuration, also known as cross flow (Fig. 11–37*b*), is similar to the first with the exception that the water that does not pass through the membrane is recirculated to a storage reservoir. The third configuration is known as direct feed (also *dead-end*) (Fig. 11–37*c*) in that there is no cross flow. All of the water applied to the membrane

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passes through the membrane. Raw feedwater is used periodically to flush the accumulated material from the membrane surface.

For the cross-flow mode of operation (see Fig. 11–37*a* and *b*), the transmembrane pressure is given by the following expression:

$$P_{tm} = \left[\frac{P_f + P_c}{2}\right] - P_p \tag{11–35}$$

where P_{tm} = transmembrane pressure gradient, kPa
P_f = inlet pressure of feed stream, kPa
P_c = pressure of concentrate stream, kPa
P_p = pressure of permeate stream, kPa

The overall pressure drop across the filter module for the cross-flow mode of operation is given by

$$P = P_f - P_p \tag{11–36}$$

where P = pressure drop across module, kPa
P_f and P_p as defined above

For the direct-feed mode of operation (see Fig. 11–37*c*) the transmembrane pressure is given by the following expression:

$$P_{tm} = P_f - P_p \tag{11–37}$$

where P_{tm} = transmembrane pressure gradient, kPa
P_f and P_p as defined above

The total permeate flow from a membrane system is given by

$$Q_p = F_w A \tag{11–38}$$

where Q_p = permeate stream flowrate, kg/s
F_w = transmembrane water flux rate, kg/m^2·s
A = membrane area, m^2

As would be expected, the transmembrane water flux rate is a function of the quality of the feed stream, the degree of pretreatment, the characteristics of the membrane, and the system operating parameters.

The recovery rate r is defined as

$$r, \% = \frac{Q_p}{Q_f} \times 100 \tag{11–39}$$

where Q_p = permeate stream flow, kg/s
Q_f = feed stream flow, kg/s

It should be noted that there is a difference in the recovery rate (which refers to the water) and the rate of rejection (which refers to the solute) as given below:

$$R, \% = \frac{C_f - C_p}{C_f} \times 100 = 1 - \frac{C_p}{C_f} \times 100 \tag{11–40}$$

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The corresponding mass balance equations are

$$Q_f = Q_p + Q_c \tag{11-41}$$

$$Q_f C_f = Q_p C_p + Q_c C_c \tag{11-42}$$

Three different operating modes can be used to control the operation of a membrane process with respect to flux and the transmembrane pressure (TMP). The three modes, illustrated on Fig. 11–38, are (1) constant flux in which the flux rate is fixed and the TMP is allowed to vary (increase) with time; (2) constant TMP in which the TMP is fixed and the flux rate is allowed to vary (decrease) with time, and (3) both the flux rate and the TMP are allowed to vary with time. Traditionally, the constant-flux mode of operation has been used. However, based on the results of a recent study with various wastewater effluents (Bourgeous et al., 1999), it appears the mode in which both the flux rate and the TMP are allowed to vary with time may be the most effective mode of operation.

Reverse Osmosis. When two solutions having different solute concentrations are separated by a semipermeable membrane, a difference in chemical potential will exist across the membrane (Fig. 11–39). Water will tend to diffuse through the membrane from the lower-concentration (higher-potential) side to the higher-concentration (lower-potential) side. In a system having a finite volume, flow continues until the pressure dif-

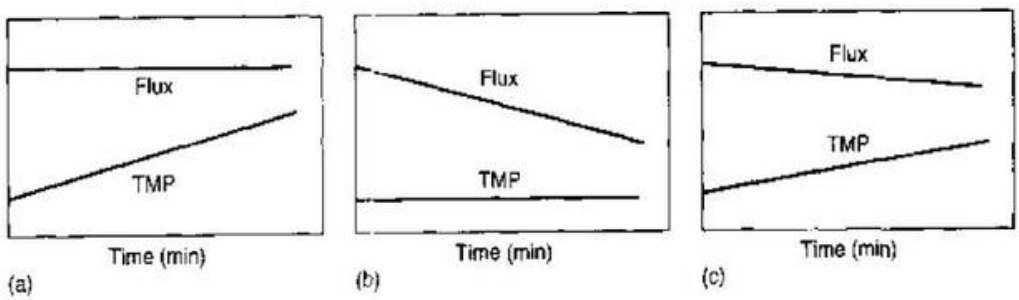

Figure 11-38

Three modes of membrane operation: (a) constant flux, (b) constant pressure, and (c) nonrestricted flux and pressure. (Adapted from Bourgeous et al., 1999.)

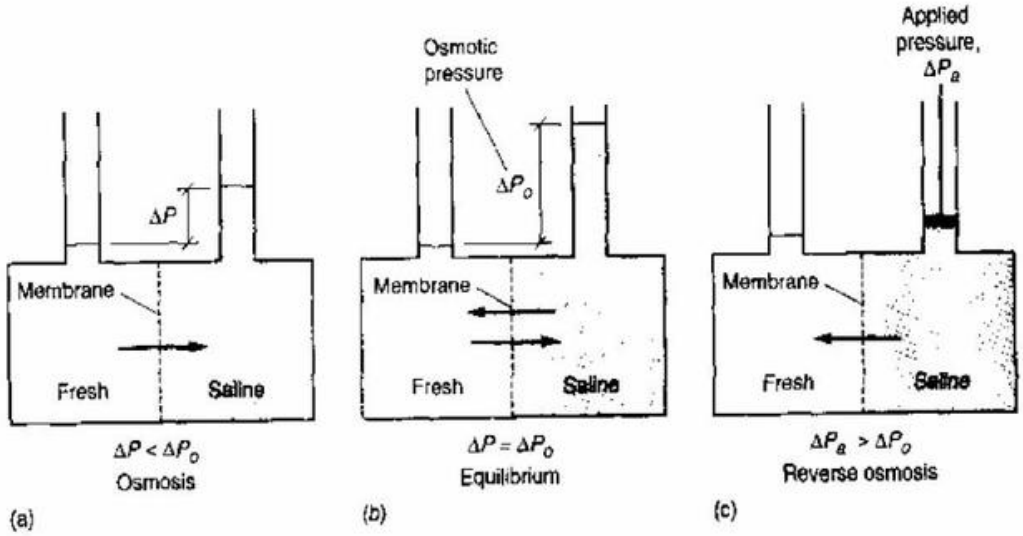

Figure 11-39

Definition sketch of osmotic flow: (a) osmotic flow, (b) osmotic equilibrium, and (c) reverse osmosis.

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ference balances the chemical potential difference. This balancing pressure difference is termed the *osmotic pressure* and is a function of the solute characteristics and concentration and temperature. If a pressure gradient opposite in direction and greater than the osmotic pressure is imposed across the membrane, flow from the more concentrated to the less concentrated region will occur and is termed *reverse osmosis* (see Fig. 11–39*c*).

A number of different models have been developed to determine the surface area of membrane and the number of arrays required (see Fig. 11–40). The basic equations used to develop the various models are as follows (Taylor and Wiesner, 1999). Referring to Fig. 11–32, the flux of water through the membrane is a function of the pressure gradient:

$$F_w = k_w(\Delta P_a - \Delta\Pi) = \frac{Q_p}{A} \tag{11-43}$$

where F_w = water flux rate, kg/m^2·s

k_w = water mass transfer coefficient involving temperature, membrane characteristics, and solute characteristics, s/m

ΔP_a = average imposed pressure gradient, kPa

$$= \left[\frac{P_f + P_c}{2}\right] - P_p$$

$\Delta\Pi$ = osmotic pressure gradient, kPa

$$= \left[\frac{\Pi_f + \Pi_c}{2}\right] - \Pi_p$$

Q_p = permeate stream flow, kg/s

A = membrane area, m^2

Some solute passes through the membrane in all cases. Solute flux can be described adequately by an expression of the form

$$F_i = k_i\,\Delta C_i = \frac{Q_p C_p}{A} \tag{11-44}$$

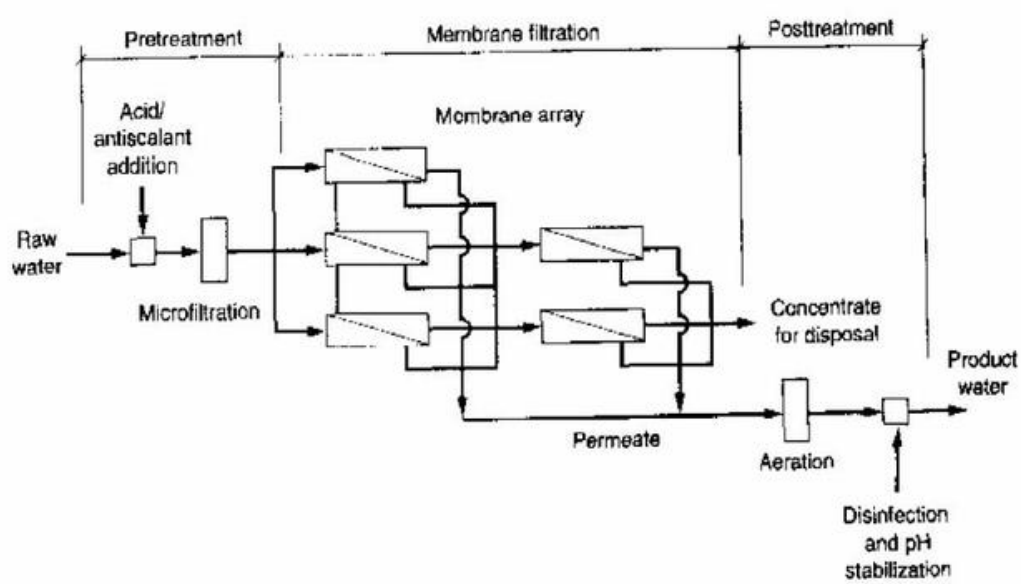

Figure 11–40

Typical flow diagram for reverse osmosis membrane process with pre- and posttreatment.

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where F_i = flux of solute species i, kg/m²·s
k_i = solute mass transfer coefficient, m/s
ΔC_i = solute concentration gradient, kg/m³

$$= \left[\frac{C_f + C_c}{2}\right] - C_p$$

C_f = solute concentration in feed stream, kg/m³
C_c = solute concentration in concentrate stream, kg/m³
C_p = solute concentration in permeate stream, kg/m³

The recovery rate and the rate of rejection are as described above for the micro- and ultrafiltration membranes. Use of the above equations to estimate the required surface area for desalination is illustrated in Example 11–5.

EXAMPLE 11–5 Determination of Membrane Area Required for Demineralization A brackish water having a TDS concentration of 3000 g/m³ is to be desalinized using a thin-film composite membrane having a flux rate coefficient k_w of 1.5×10^{-6} s/m and a mass transfer rate coefficient k_i of 1.8×10^{-6} m/s. The product water is to have a TDS of no more than 200 g/m³. The flowrate is to be 0.010 m³/s. The net operating pressure ($\Delta P_a - \Delta\Pi$) will be 2500 kPa. Assume the recovery rate will be 90 percent. Estimate the rejection rate and the concentration of the concentrate stream.

Solution

1. The problem involves determination of the membrane area required to produce 0.010 m³/s of water and the TDS concentration of the permeate. If the permeate TDS concentration is well below 200 g/m³, blending of feed and permeate will reduce the required membrane area.
2. Estimate membrane area using Eq. (11–43).

$$F_w = k_w(\Delta P_a - \Delta\Pi)$$
$$= (1.5 \times 10^{-6}\ \text{s/m})(2500\ \text{kg/m}^2) = 3.75 \times 10^{-3}\ \text{kg/m}^2\cdot\text{s}$$

$$Q_p = F_w \times A$$

$$A = \frac{(0.010\ \text{m}^3/\text{s})(10^3\ \text{kg/m}^3)}{(3.75 \times 10^{-3}\ \text{kg/m}^2\cdot\text{s})} = 2667\ \text{m}^2$$

3. Estimate permeate TDS concentration using Eq. (11–44).

$$F_i = k_i\,\Delta C_i = \frac{Q_p C_p}{A}$$

$$Q_p C_p = k_i\left(\left[\frac{C_f + C_c}{2}\right] - C_p\right)A$$

Assume $C_c \approx C_f$ and solve for C_p

$$C_p = \frac{k_i A C_f}{Q_p + k_i A}$$

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Assume $Q_p = r\,Q_f$

$$C_p = \frac{(1.8 \times 10^{-6}\ \text{m/s})(2667\ \text{m}^2)(3.0\ \text{kg/m}^3)}{(0.01)(0.9) + (1.8 \times 10^{-6}\ \text{m/s})(2667\ \text{m}^3)} = 0.152\ \text{kg/m}^3$$

The permeate solute concentration is lower than necessary. It may be possible to reduce the area by blending.

4. Estimate the rejection rate using Eq. (11–40)

$$R, \% = \frac{C_f - C_p}{C_f} \times 100$$

$$R = \frac{(3.0\ \text{kg/m}^3 - 0.152\ \text{kg/m}^3)}{(3.0\ \text{kg/m}^3)} \times 100 = 95\%$$

5. Estimate the concentrate stream TDS using Eq. (11–42)

$$C_c = \frac{Q_f C_f - Q_p C_p}{Q_c}$$

$$C_c = \frac{(0.1\ \text{L})(3.0\ \text{kg/m}^3) - (0.9\ \text{L})(0.152\ \text{kg/m}^3)}{0.1\ \text{L}} = 31.4\ \text{kg/m}^3$$

Membrane Fouling

The term *fouling* is used to describe the potential deposition and accumulation of constituents in the feed stream on the membrane. Membrane fouling is an important consideration in the design and operation of membrane systems as it affects pretreatment needs, cleaning requirements, operating conditions, cost, and performance. Constituents in wastewater that can bring about membrane fouling are identified in Table 11–18. Fouling of the membrane, as reported in Table 11–18, can occur in three general forms: (1) a buildup of the constituents in the feedwater on the membrane surface, (2) the formation of chemical precipitates due to the chemistry of the feedwater, and (3) damage to the membrane due to the presence of chemical substances that can react with the membrane or biological agents that can colonize the membrane.

Membrane Fouling Caused by Buildup of Solids. Three accepted mechanisms resulting in resistance to flow due to the accumulation of material within a lumen (see Fig. 11–41) are (1) pore narrowing, (2) pore plugging, and (3) gel/cake formation caused by concentration polarization (Ahn et al., 1998). Gel/cake formation, caused by concentration polarization, occurs when the majority of the solid matter in the feed is larger than the pore sizes or molecular weight cutoff of the membrane. Concentration polarization can be described as the buildup of matter close to or on the membrane surface that causes an increase in resistance to solvent transport across the membrane. Some degree of concentration polarization will always occur in the operation of a membrane system. The formation of a gel or cake layer, however, is an extreme

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Table 11–18 Constituents in wastewater that can affect the performance of membranes through the mechanism of fouling

Type of membrane fouling	Responsible constituents	Remarks
Fouling (cake formation sometimes identified as biofilm formation)	Metal oxides Organic and inorganic colloids Bacteria Microorganisms Concentration polarization	Damage to membranes can be limited by controlling these substances (for example, by the use of microfiltration before reverse osmosis)
Scaling (precipitation)	Calcium sulfate Calcium carbonate Calcium fluoride Barium sulfate Metal oxide formation Silica	Scaling can be reduced by limiting salt content, by adding acid to limit the formation of calcium carbonate, and by other chemical treatments (e.g., the addition of antiscalants)
Damage to membrane	Acids Bases pH extremes Free chlorine Bacteria Free oxygen	Damage to membranes can be limited by controlling these substances. Extent of damage depends on the nature of the membrane

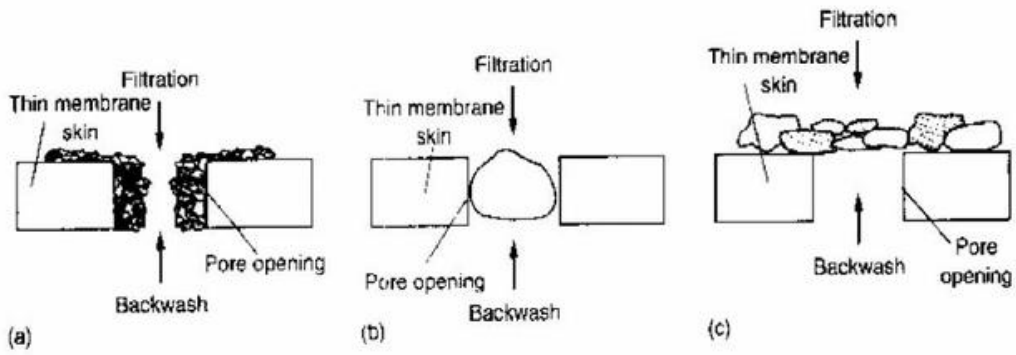

Figure 11–41 Modes of membrane fouling: (a) pore narrowing, (b) pore plugging, and (c) gel/cake formation caused by concentration polarization. *(Adapted from Bourgeous et al., 1999.)*

case of concentration polarization where a large amount of matter has actually accumulated on the membrane surface, forming a gel or cake layer. The mechanisms of pore plugging and pore narrowing will occur only when the solid matter in the feedwater is smaller than the pore size or the molecular weight cutoff. As the name describes, pore plugging occurs when particles the size of the pores become stuck in the pores of the membrane. Pore narrowing consists of solid material attaching to the interior surface of the pores, which results in a narrowing of the pores. It has been hypothesized that once the pore size is reduced, concentration polarization is amplified further, causing an increase in fouling (Crozes et al.,1997).

Control of Membrane Fouling. Typically, three approaches are used to control membrane fouling: (1) pretreatment of the feedwater, (2) membrane backflushing, and (3) chemical cleaning of the membranes. Pretreatment is used to reduce the TSS and bacterial content of the feedwater. Often the feedwater will be conditioned chemi-

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cally to limit chemical precipitation within the units. The most commonly used method of eliminating the accumulated material from the membrane surface is backflushing with water and/or air. Chemical treatment is used to remove constituents that are not removed during conventional backwashing. Chemical precipitates can be removed by altering the chemistry of the feedwater and by chemical treatment. Damage of the membrane due to deleterious constituents typically cannot be reversed.

Pretreatment for Nanofiltration and Reverse Osmosis. A very high-quality feed is required for efficient operation of a nanofiltration or reverse osmosis unit. Membrane elements in the reverse osmosis unit can be fouled by colloidal matter and constituents in the feed stream. The following pretreatment options have been used singly and/or in combination.

1. Pretreatment of a secondary effluent by chemical clarification and multimedia filtration or by multimedia filtration and ultrafiltration is usually necessary (see Fig. 11–40) to remove colloidal material.
2. Cartridge filters with a pore size of 5- to 10-μm have also been used to reduce residual suspended solids.
3. To limit bacterial activity it may be necessary to disinfect the feedwater using either chlorine, ozone, or UV radiation.
4. The exclusion of oxygen may be necessary to prevent oxidation of iron, manganese, and hydrogen sulfide.
5. Depending on the type of membrane, removal of chlorine (with sodium bisulfite) and ozone may be necessary.
6. The removal of iron and manganese may also be necessary to decrease scaling potential.
7. To inhibit scale formation, the pH of the feed should be adjusted (usually with sulfuric acid) within the range from 4.0 to 7.5.

Regular chemical cleaning of the membrane elements (about once a month) is necessary to restore the membrane flux.

Assessing Need for Pretreatment for NF and RO. To assess the treatability of a given wastewater with NF and RO membranes, a variety of fouling indexes have been developed over the years. The three principal indexes are the silt density index (SDI), the modified fouling index (MFI), and the mini plugging factor index (MPFI). Fouling indexes are determined from simple membrane tests. The sample must be passed through a 0.45 μm Millipore filter with a 47 mm internal diameter at 210 kPa (30 lb/in^2) gage to determine any of the indexes. The time to complete data collection for these tests varies from 15 min to 2 h, depending on the fouling nature of the water.

The most widely used index is the SDI. The SDI is defined as follows:

$$\mathrm{SDI} = \frac{100[1 - (t_i/t_f)]}{t} \tag{11–45}$$

where t_i = time to collect initial sample of 500 mL
t_f = time to collect final sample of 500 mL
t = total time for running the test

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The silt density index is a static measurement of resistance that is determined by samples taken at the beginning and end of the test. The SDI does not measure the rate of change of resistance during the test. Recommended SDI values are reported in Table 11–19. The calculation of the SDI is demonstrated in Example 11–6.

The modified fouling index (MFI) is determined using the same equipment and procedure used for the SDI, but the volume is recorded every 30 s over a 15-min filtration period. Derived from a consideration of cake filtration, the MFI is defined as follows:

$$\frac{1}{Q} = a + \text{MFI} \times V \tag{11-46}$$

where Q = average flow, L/s
a = constant
MFI = modified fouling index, s/L^2
V = volume, L

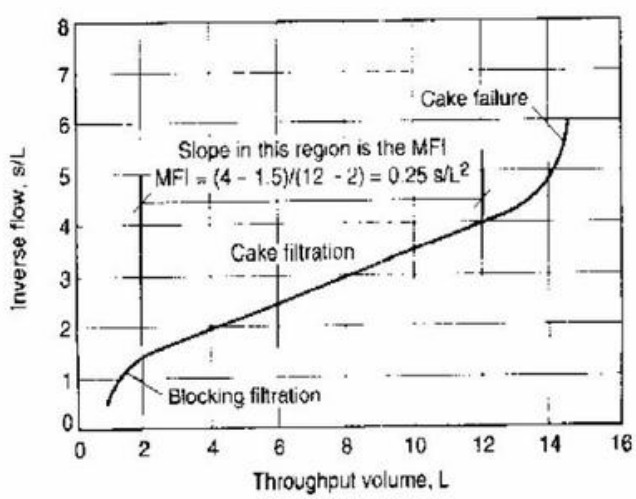
The value of the MFI is obtained as the slope of the straight-line portion of the curve obtained by plotting the inverse flow versus the cumulative volume (see Fig. 11–42). Recommended MFI values are reported in Table 11–19.

Table 11–19
Approximate values for fouling indexes[a]

Membrane process	Fouling index SDI	Fouling index MFI, s/L^2
Nanofiltration	0–2	0–10
Reverse osmosis hollow fiber	0–2	0–2
Reverse osmosis spiral wound	0–3	0–2

[a] Adapted in part from Taylor and Wiesner (1999) and AWWA (1996).

Figure 11–42
Typical plot to determine the modified fouling factor.



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EXAMPLE 11–6 Silt Density Index for Reverse Osmosis Determine the silt density index for a proposed feedwater from the following test data. If a spiral-wound RO membrane is to be used, will pretreatment be required?

Test run time = 30 min

Initial 500 mL = 2 min

Final 500 mL = 10 min

Solution

1. Calculate the SDI using Eq. (11–45).

$$\text{SDI} = \frac{100[1 - (t_i/t_f)]}{t}$$

$$\text{SDI} = \frac{100[1 - (2/10)]}{30} = 2.67$$

2. Compare the SDI to the acceptable criteria.

Comment Calculated SDI value of 2.67 is less than 3 (see Table 11–19); therefore, no further pretreatment would be needed normally. As a practical matter, because the SDI value is close to 3.0 it may be prudent to consider some form of pretreatment to prolong the filtration cycle.

Application of Membranes

With evolving health concerns (see Chap. 2) and the development of new and lower-cost membranes, the application of membrane technologies in the field of environmental engineering has increased dramatically within the past 5 years. The increased use of membranes is expected to continue well into the future. In fact, the use of conventional filtration technology, such as described in Secs. 11–3 and 11–4, may be a thing of the past within 10 to 15 years, especially in light of the need to remove resistant organic constituents of concern (see Chap. 13). Typical applications of membrane technologies for wastewater treatment are reported in Table 11–20. The principal applications of the various membrane technologies for the removal of specific constituents found in wastewater are summarized in Table 11–21. A comparison of the constituents found in wastewater and the operating ranges for the various membrane technologies was presented previously on Fig. 11–34. Typical operating ranges in terms of operating pressures and flux rates, along with the types of membranes used, are reported in Table 11–22.

Typical energy consumption and product recovery values for various membrane systems are presented in Table 11–23. In reviewing the information presented in Table 11–23, it is important to note that the reported operating pressure values for all of the membrane processes, excluding electrodialysis, are considerably lower than comparable values of 5 years ago. It is anticipated that operating pressures will continue to go down as new membranes are developed. At the present time where the use of membranes is being considered, special attention must be devoted to the characteristics of the wastewater to be

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Table 11-20
Typical applications for membrane technologies in wastewater treatment[a]

Applications	Description
Microfiltration and ultrafiltration	
Aerobic biological treatment	Membrane is used to separate the treated wastewater from the active biomass in an activated-sludge process. The membrane separation unit can be internal immersed in the bioreactor or external to the bioreactor (see Fig. 11-45). Such processes are known as membrane bioreactor (MBR) processes
Anaerobic biological treatment	Membrane is used to separate the treated wastewater from the active biomass in an anaerobic complete-mix reactor
Membrane aeration biological treatment	Plate and frame, tubular, and hollow membranes are used to transfer pure oxygen to the biomass attached to the outside of the membrane (see Fig. 11-47*a*). Such processes are known as membrane aeration bioreactor (MABR) processes
Membrane extraction biological treatment	Membranes are used to extract degradable organic molecules from inorganic constituents such as acids, bases, and salts from the waste stream for subsequent biological treatment (see Fig. 11-47*b*). Such processes are known as extractive membrane bioreactor (EMBR) processes
Pretreatment for effective disinfection	Used to remove residual suspended solids from settled secondary effluent or from the effluent from depth or surface filters to achieve effective disinfection with either chlorine or UV radiation for reuse applications
Pretreatment for nanofiltration and reverse osmosis	Microfilters are used to remove residual colloidal and suspended solids as a pretreatment step for additional processing
Nanofiltration	
Effluent reuse	Used to treat prefiltered effluent (typically with microfiltration) for indirect potable reuse applications such as groundwater injection. Credit is also given for disinfection when using nanofiltration
Wastewater softening	Used to reduce the concentration of multivalent ion contributing to hardness for specific reuse applications
Reverse osmosis	
Effluent reuse	Used to treat prefiltered effluent (typically with microfiltration) for indirect potable reuse applications such as groundwater injection. Credit is also given for disinfection when using reverse osmosis
Effluent dispersal	Reverse osmosis processes have proved capable of removing sizable amounts of selected compounds such as NDMA
Two-stage treatment for boiler use	Two stages of reverse osmosis are used to produce water suitable for high-pressure boilers

[a] Adapted in part from Stephenson et al. (2000).

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Table 11-21 Application of membrane technologies for the removal of specific constituents found in wastewater[a]

Constituent	Membrane technology MF	UF	NF	RO	Comments
Biodegradable organics		✔	✔	✔	
Hardness			✔	✔	
Heavy metals			✔	✔	
Nitrate			✔	✔	
Priority organic pollutants		✔	✔	✔	
Synthetic organic compounds			✔	✔	
TDS			✔	✔	
TSS	✔	✔			TSS removed during pretreatment for NF and RO
Bacteria	✔[b]	✔	✔	✔	Used for membrane disinfection. Removed as pretreatment for NF and RO with MF and UF
Protozoan cysts and oocysts and helminth ova	✔	✔	✔	✔	
Viruses			✔	✔	Used for membrane disinfection

[a] Specific removal rates will depend on the composition and constituent concentrations in the treated wastewater.

[b] Variable performance.

processed. Advantages and disadvantages of MF and UF and RO membrane technology are compared in Table 11-24.

Microfiltration. Microfiltration membranes are the most numerous on the market, are the least expensive, and as reported in Table 11-22, are commonly made of polypropylene, acrylonitrile, nylon, and polytetrafluoroethylene. Typical operating information for microfiltration and ultrafiltration membrane technologies used for wastewater including operating pressures and flux rates is also presented in Table 11-22. Microfiltration technologies, as discussed below, can be used in a variety of ways in wastewater-treatment and water reuse systems.

In advanced treatment applications, microfiltration has been used, most commonly as a replacement for depth filtration to reduce turbidity, remove residual suspended solids, and reduce bacteria to condition the water for effective disinfection (see Fig. 11-43*a*) and as a pretreatment step for reverse osmosis (see Figs. 11-40 and 11-43*a*). The MF units at Dublin San Ramon Sanitary District, used to produce treated water for specific reuse applications or as a pretreatment step for RO, are shown on Fig. 11-44. Corresponding performance data are presented in Table 11-25. Typical

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Table 11-22
Typical characteristics of membrane technologies used in wastewater-treatment applications[a]

Membrane technology	Typical operating range, μm	Operating pressure		Rate of flux		Membrane details	
		lb/in²	kPa	gal/ft²·d	L/m²·d	Type	Configuration
Microfiltration	0.08–2.0	1–15	7–100	10–40	405–1600	Polypropylene, acrylonitrile, nylon, and polytetrafluoroethylene	Spiral wound, hollow fiber, plate and frame
Ultrafiltration	0.005–0.2	10–100	70–700	10–20	405–815	Cellulose acetate, aromatic polyamides	Spiral wound, hollow fiber, plate and frame
Nanofiltration	0.001–0.01	75–150	500–1000	5–20	200–815	Cellulose acetate, aromatic polyamides	Spiral wound, hollow fiber
Reverse osmosis	0.0001–0.001	125–1000	850–7000	8–12	320–490	Cellulose acetate, aromatic polyamides	Spiral wound, hollow fiber, thin-film composite

[a] Adapted from Crites and Tchobanoglous (1998).

Note: kPa × 0.1450 = lb/in²

L/m²·d × 0.024542 = gal/ft²·d

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Table 11–23 Typical energy consumption and product recovery values for various membrane systems

Membrane process	Operating pressure		Energy consumption, kWh per		Product recovery, %
	lb/in²	kPa	1000 gal	m³	
Microfiltration	15	100	0.1	0.4	94–98
Ultrafiltration	75	525	0.8	3.0	70–80
Nanofiltration	125	875	1.4	5.3	80–85
Reverse osmosis	225	1575	2.7	10.2	70–85
Reverse osmosis	400	2800	4.8	18.2	70–85
Electrodialysis			2.5	9.5	75–85

Table 11–24 Advantages and disadvantages of membrane treatment technologies

Advantages	Disadvantages
Microfiltration and ultrafiltration	
• Can reduce the amount of treatment chemicals • Smaller space requirements (footprint); membrane equipment requires 50 to 80 percent less space than conventional plants • Reduced labor requirements; can be automated easily • New membrane design allows use of lower pressures; system cost may be competitive with conventional wastewater-treatment processes • Removes protozoan cysts, oocysts, and helminth ova; may also remove limited amounts of bacteria and viruses	• Uses more electricity; high-pressure systems can be energy-intensive • May need pretreatment to prevent fouling; pretreatment facilities increase space needs and overall costs • May require residuals handling and disposal of concentrate • Requires replacement of membranes about every 3 to 5 years • Scale formation can be a serious problem. Scale-forming potential difficult to predict without field testing • Flux rate (the rate of feedwater flow through the membrane) gradually declines over time. Recovery rates may be considerably less than 100 percent • Lack of a reliable low-cost method of monitoring performance
Reverse osmosis	
• Can remove dissolved constituents • Can disinfect treated water • Can remove NDMA and other related organic compounds • Can remove natural organic matter (a disinfection by-product precursor) and inorganic matter	• Works best on groundwater or low solids surface water or pretreated wastewater effluent • Lack of a reliable low-cost method of monitoring performance • May require residuals handling and disposal of concentrate • Expensive compared to conventional treatment

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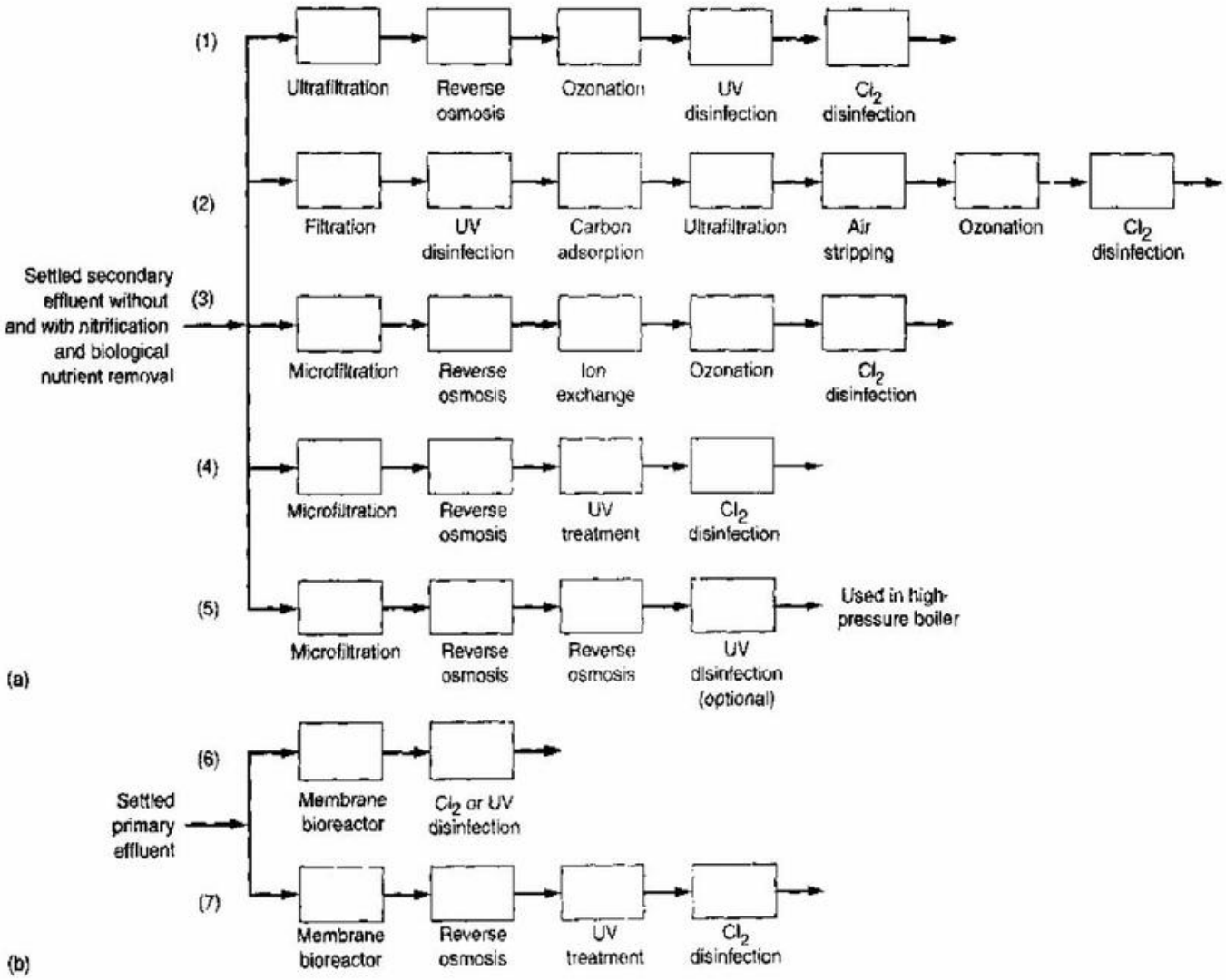

Figure 11–43

Typical flow diagrams employing membranes for advanced wastewater treatment with (a) settled secondary effluent and (b) settled primary effluent. Within the past 5 years, the use of membranes has essentially replaced the use of chemical precipitation in advanced wastewater-treatment applications. In flow diagram 5, two stages of reverse osmosis are used to produce an extremely low TDS water suitable for use in high-pressure boilers. In flow diagram 7, UV can be used for both disinfection and the destruction of NDMA. Chlorine is added in flow diagrams 1 through 4 and 7 to limit the growth of microorganisms in the distribution or conveyance system.

performance data reported in the literature are also included in Table 11–25 for the purpose of comparison. Care should be used in applying the performance data reported in Table 11–25 as it has been found that the performance of MF is to a large extent site-specific, especially with respect to fouling.

A relatively recent development, the use of membranes for biological treatment, promises to be one of the most important uses of membranes in wastewater treatment. As reported in Table 11–20, membranes have been used for both aerobic and anaerobic treatment of wastewaters. Typical membrane bioreactors (MBRs) are illustrated on Fig. 11–45. On Fig. 11–45*a*, the membrane separation unit is internal, immersed in the bioreactor. Treated effluent is withdrawn from the bioreactor with the application of a

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Figure 11-44

Microfiltration (MF) units at Dublin San Ramon Sanitary District used as a pretreatment step for RO.

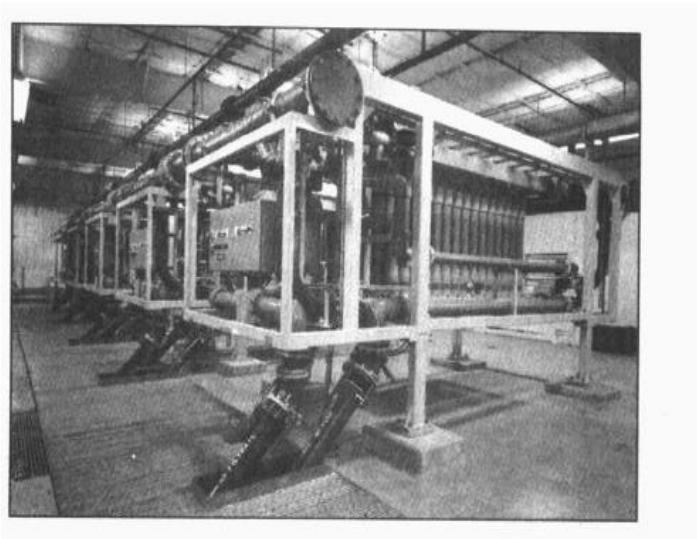


Table 11-25 Performance summary for the Dublin San Ramon Sanitary District MF for the period from 4/00 through 12/00[a,b]

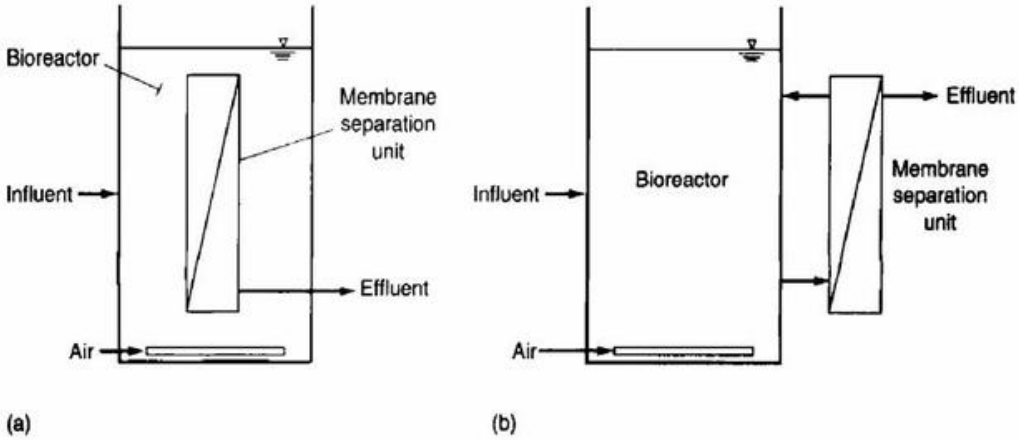
Constituent	MF influent, mg/L	MF effluent, mg/L	Average reduction, %	Reduction reported in literature, %
TOC	10–31	9–16	57	45–65
BOD	11–32	<2–9.9	86	75–90
COD	24–150	16–53	76	70–85
TSS	8–46	<0.5	97	95–98
TDS	498–622	498–622	0	0–2
NH_3–N	21–42	20–35	7	5–15
NO_3–N	<1–5	<1–5	0	0–2
PO_4^-	6–8	6–8	0	0–2
SO_4^{2-}	90–120	90–120	0	0–1
Cl^-	93–115	93–115	0	0–1
Turbidity	2–50 NTU	0.03–0.08 NTU	>99	

[a] From Whitley Burchett & Associates (1999).

[b] Typical flux rate during test period was 1600 L/m²·d

Figure 11-45

Schematic flow diagrams for the membrane bioreactor activated-sludge process: (a) with internal membrane biosolids separation unit and (b) with external biosolids separation unit.



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vacuum (on the order of 50 kPa). On Fig. 11–45*b*, the membrane separation unit is external to the bioreactor. A photograph of a typical membrane separation unit installed in a test reactor is shown on Fig. 11–46. The aeration system installed below the membrane separation unit is used to clean the hollow fibers continuously by the shearing action of the air bubbles as they rise through the liquid (Stephenson et al., 2000).

Typical performance data for the MBR activated-sludge process are presented in Table 11–26. As reported in Table 11–26, the quality of the effluent is ideal for a number of reuse application or for further processing by nanofiltration or reverse osmosis. In addition to offering excellent performance, MBRs require a considerably smaller aerial footprint. It is also interesting to note that MBRs have also been used for the anaerobic treatment of wastewater. The current literature abounds with articles delineating the use of membranes in a variety of wastewater-treatment applications. The biological characteristics of membrane bioreactors are discussed in greater detail in Chap. 8.

In addition to the MBR process, plate and frame, tubular, and hollow membranes have been used to transfer pure oxygen to the biomass attached to the outside of the membrane (see Fig. 11–47*a*). Such processes are known as membrane aeration bioreactor (MABR) processes. In another application, membranes are used to extract degradable soluble organic molecules from inorganic constituents in a waste stream (see Fig. 11–47*b*). The extracted organic constituents are treated in an external bioreactor. Such processes are known as extractive membrane bioreactor (EMBR) processes.

Figure 11–46

Photograph of horizontal type membrane bundle used in membrane bioreactors installed in pilot-scale reactor for testing with clear water.

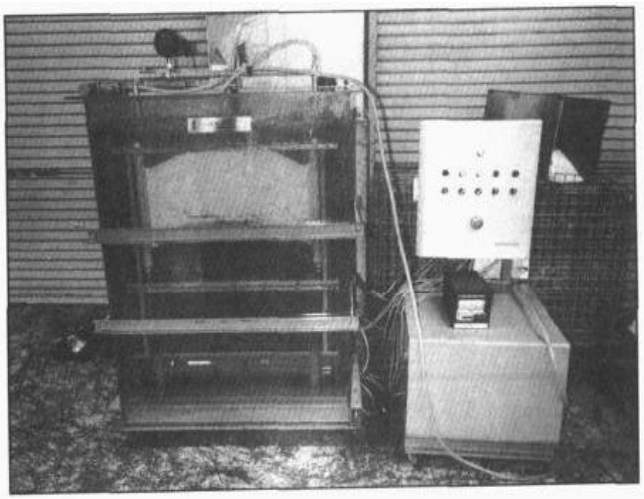

Table 11–26
Typical performance data for membrane bioreactor used to treat domestic wastewater[a]

Parameter	Unit	Typical
Effluent BOD	mg/L	<5
Effluent COD	mg/L	<30
Effluent NH_3	mg/L	<1
Effluent TN	mg/L	<10
Effluent turbidity	NTU	<1

[a] For additional details on membrane bioreactors, see Sec. 8–9 in Chap. 8.

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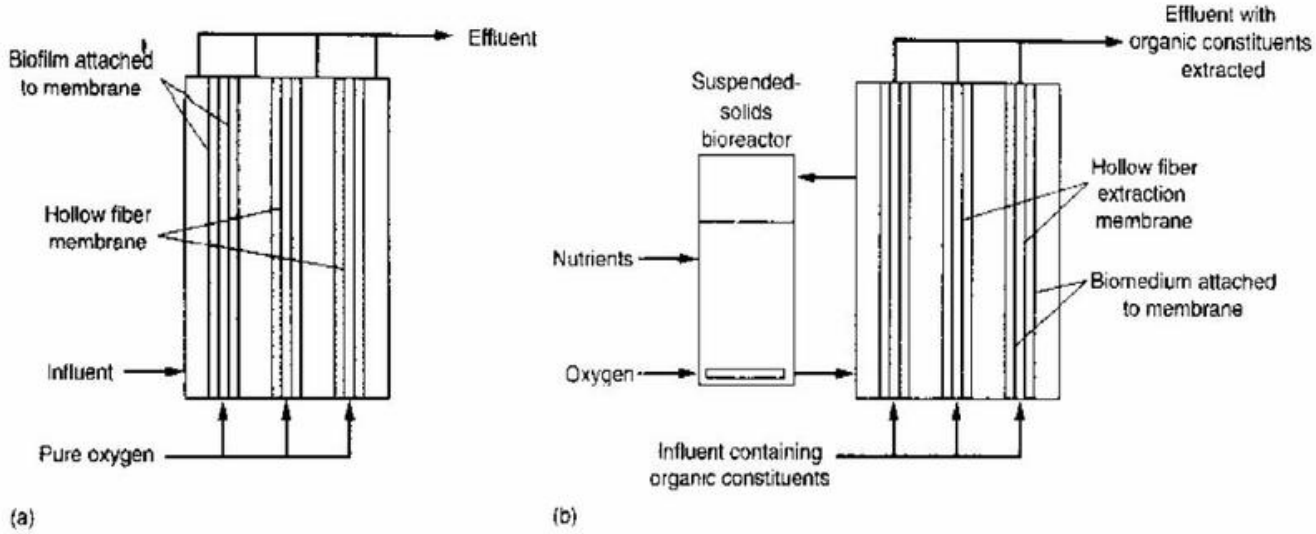

Figure 11-47

Schematic flow diagrams for alternative biological treatment processes employing membranes: (*a*) membrane aeration bioreactor, (*b*) extractive membrane bioreactor with external treatment unit.

Additional details on aeration and extractive membrane bioreactors applications may be found in Stephenson et al. (2000).

Ultrafiltration. Ultrafiltration (UF) membranes are used for many of the same applications as described above for microfiltration. Some UF membranes with small pore sizes have also been used to remove dissolved compounds with high molecular weight, such as colloids, proteins, and carbohydrates. The membranes do not remove sugar or salt. Ultrafiltration is used typically in industrial applications for the production of high-purity process rinse water. Typical operating data are presented in Table 11–22.

Nanofiltration. Nanofiltration, also known as "loose" RO, can reject particles as small as 0.001 μm. Nanofiltration is used for the removal of selected dissolved constituents from wastewater such as the multivalent metallic ions responsible for hardness. The advantages of nanofiltration over lime softening include the production of a product water that meets the most stringent reuse water quality requirements. Because both inorganic and organic constituents and bacteria and viruses are removed, disinfection requirements are minimized. Although most NF facilities use polyamide TFC membranes in a spiral-wound configuration, more than ten different types of membranes are available. Other membranes include polyamide hollow fiber, polyvinyl acetate spiral wound, and asymmetric cellulose acetate in a tubular configuration.

Reverse Osmosis. Worldwide, reverse osmosis (RO) is used primarily for desalination. In wastewater treatment, RO is used for the removal of dissolved constituents from wastewater remaining after advanced treatment with depth filtration or microfiltration. The membranes exclude ions, but require high pressures to produce the deionized water. Typical flow diagrams involving the use of reverse osmosis are shown on Figs. 11–40 and 11–43. Typical operating information for reverse osmosis used for wastewater including

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operating pressures and flux rate rates is reported in Table 11–22. Typical RO units used for the repurification of pretreated water (with MF) are shown on Fig. 11–48. Corresponding performance data are presented in Table 11–27. Typical performance data reported in the literature are also included in Table 11–27. As noted above, care should be used in applying the performance data reported in Table 11–27 as it has been found that the performance of RO is also site-specific, especially with respect to fouling. Disinfection of the RO feedwater is usually practiced to minimize or limit the bacterial growth on the membrane. Care must be taken with polyamide and TFC membranes because they are sensitive to chemical oxidants. The need for cleaning depends on how long the flux rate is maintained.

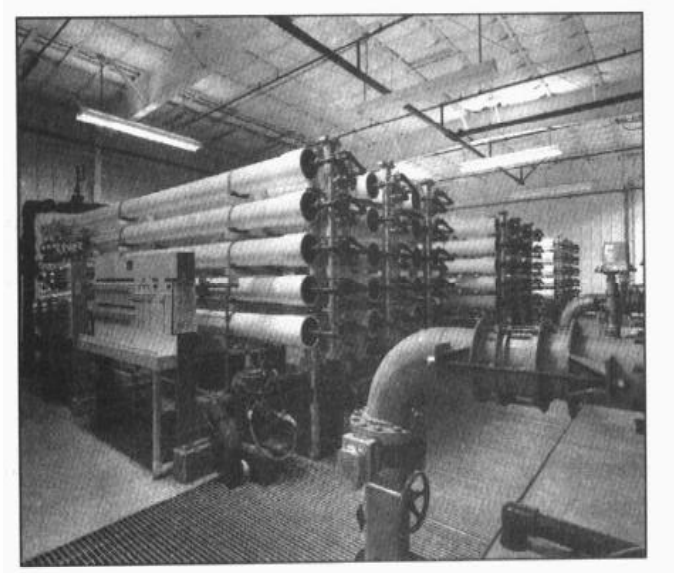

Figure 11–48
Reverse osmosis (RO) units at Dublin San Ramon Sanitary District used for the repurification of water pretreated with MF (see Fig. 11–44).

Table 11–27
Performance summary for the Dublin San Ramon Sanitary District RO for the period from 4/1999 through 12/1999[a,b]

Constituent	RO influent, mg/L	RO effluent, mg/L	Average reduction, %	Reduction reported in literature, %
TOC	9–16	<0.5	>94	85–95
BOD	<2–9.9	<2	>40	30–60
COD	16–53	<2	>91	85–95
TSS	<0.5	~0	>99	95–100
TDS	498–622	9–19	0	90–98
NH_3–N	20–35	1–3	96	90–98
NO_3–N	<1–5	0.08–3.2	96	65–85
PO_4^-	6–8	0.1–1	~99	95–99
SO_4^{2-}	90–120	<0.5–0.7	99	95–99
Cl^-	93–115	0.9–5.0	97	90–98
Turbidity	0.03–0.08 NTU	0.03 NTU	50	40–80

[a] From Whitley Burchett & Associates (1999).
[b] Typical flux rate during test period was 348 L/m²·d.

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Electrodialysis

In the electrodialysis process, ionic components of a solution are separated through the use of semipermeable ion-selective membranes. Because the electrodialysis process is being evaluated for a variety of reuse applications, the purpose of the following discussion is (1) to introduce the theory of electrodialysis and (2) to consider some applications.

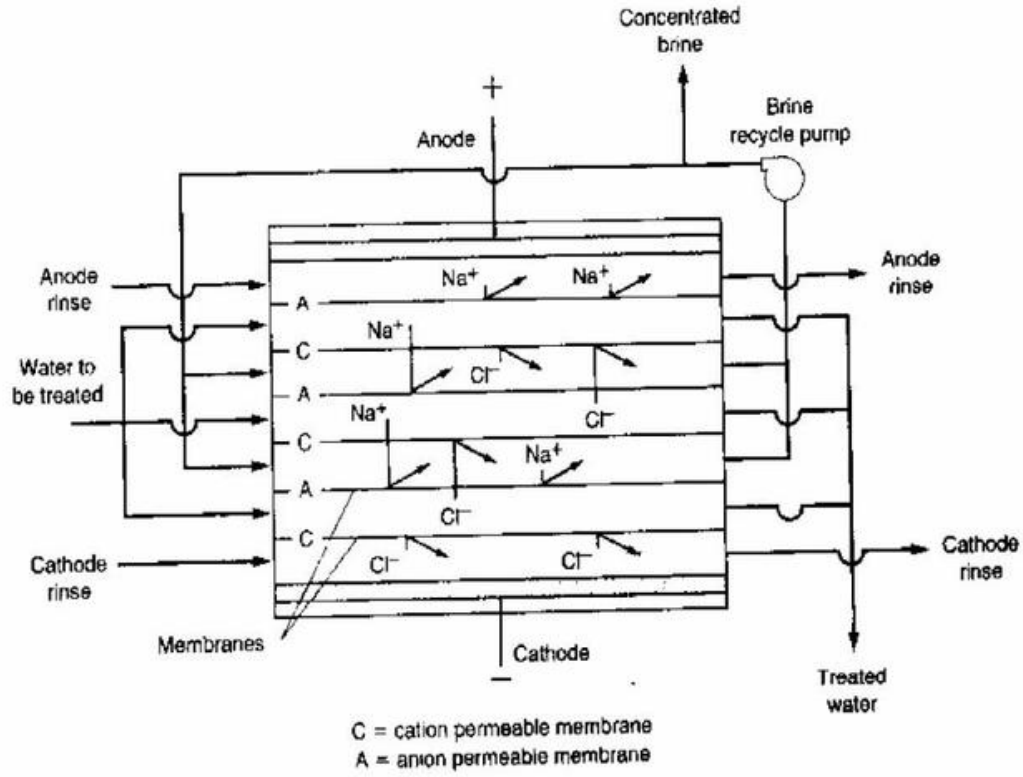
Theory of Electrodialysis. Application of an electrical potential between the two electrodes causes an electric current to pass through the solution, which in turn causes a migration of cations toward the negative electrode and a migration of anions toward the positive electrode (see Fig. 11–49). Because of the alternate spacing of cation- and anion-permeable membranes, cells of concentrated and dilute salts are formed. Wastewater is pumped through the membranes, which are separated by spacers and assembled into stacks. The wastewater is usually retained for about 10 to 20 days in a single stack or stage. Dissolved solids removals vary with the (1) wastewater temperature, (2) amounts of electrical current passed, (3) type and amount of ions, (4) permselectivity of the membrane, (5) fouling and scaling potential of the wastewater, (6) wastewater flowrates, and (7) number and configuration of stages.

The current required for electrodialysis can be estimated using Faraday's laws of electrolysis. Because one Faraday of electricity will cause one gram equivalent of a substance to migrate from one electrode to another, the number of gram equivalents removed per unit time is given by

$$\text{Gram eq/unit time} = QN\eta \tag{11-47}$$

Figure 11–49

Schematic of electrodialysis unit.



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where Q = flowrate, L/s
N = normality of solution, eq/L
η = electrolyte removal as a fraction

The corresponding expression for the current for a stack of membranes is given by

$$I = \frac{FQN\eta}{nE_c} \tag{11-48}$$

where I = current, amp
F = Faraday's constant
= 96,485 amp·s/gram equivalent = 96,485 A·s/eq
n = number of cells in the stack
E_c = current efficiency expressed as a fraction

In the analysis of the electrodialysis process, it has been found that the capacity of the membrane to pass an electrical current is related to the current density (*CD*) and the normality (*N*) of the feed solution. Current density is defined as the current in milliamperes that flows through a square centimeter of membrane perpendicular to the current direction. The relationship between current density and the solution normality is known as the *current density to normality* (*CD/N*) ratio. High values of the ratio are indicative that there is insufficient charge to carry the current. When high ratios exist, a localized deficiency of ions may occur on the surface of the membrane, causing a condition called *polarization*. Polarization should be avoided, as it results in high electrical resistance leading to excessive power consumption. In practice, *CD/N* ratios will vary from 500 to 800 when the current density is expressed as mA/cm^2. The resistance of an electrodialysis unit used to treat a particular water must be determined experimentally. Once the resistance R and the current flow I are known, the power required can be computed using Ohm's law as follows:

$$P = E \times I = R(I)^2 \tag{11-49}$$

where P = power, W
E = voltage, V
= $R \times I$
R = resistance, Ω
I = current, A

The application of these relationships is considered in Example 11-7.

EXAMPLE 11-7 Area and Power Requirements for Electrodialysis Determine the area and power required to demineralize 4000 m^3/d of treated wastewater to be used for industrial cooling water using an electrodialysis unit comprised of 240 cells. Assume the following data apply:

1. TDS concentration = 2500 mg/L
2. Cation and anion concentration = 0.010 g-eq/L

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3. Efficiency of salt removal = 50 percent
4. The current efficiency = 90 percent
5. The *CD/N* ratio = 500 mA/cm^2
6. Resistance = 5.0 Ω

Solution

1. Calculate the current using Eq. (11–48).

$$I = \frac{FQN\eta}{nE_c}$$

$$Q = (4000\ \text{m}^3/\text{d})(10^3\ \text{L/m}^3)/(86{,}400\ \text{s/d}) = 46.3\ \text{L/s}$$

$$I = \frac{(96{,}485\ \text{A·s/g eq})(46.3\ \text{L/s})(0.010\ \text{g eq/L})(0.50)}{240 \times 0.90}$$

$$I = 103.4\ \text{A}$$

2. Determine the power required using Eq. (11–49).

$$P = R(I)^2$$

$$P = (5.0\ \Omega)(103.4\ \text{A})^2 = 53{,}477\ \text{W} = 53.5\ \text{kW}$$

3. Determine the required surface area.
   a. Determine the current density:

$$CD = (500)\ (\text{normality}) = 500\ \text{mA/cm}^2 \times 0.010 = 50\ \text{mA/cm}^2$$

   b. The required area is

$$\text{Area} = \frac{(103.4\ \text{A})(10^3\ \text{mA/A})}{(50\ \text{mA/cm}^2)} = 2068\ \text{cm}^2$$

   c. Determine area of membrane assuming a square configuration will be used.

$$\text{Area per membrane} = \sqrt{2068\ \text{cm}^2} \approx 45\ \text{cm}^2$$

Comment The actual performance will have to be determined from pilot tests.

Application. The electrodialysis process may be operated in either a continuous or a batch mode. The units can be arranged either in parallel to provide the necessary hydraulic capacity or in series to effect the desired degree of demineralization. A typical flow diagram for electrodialysis membrane process with pretreatment is shown on Fig. 11–50. Makeup water, usually about 10 percent of the feed volume, is required to wash the membranes continuously. A portion of the concentrate stream is recycled to maintain nearly

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Figure 11–50
Typical flow diagram for electrodialysis membrane process with pretreatment.

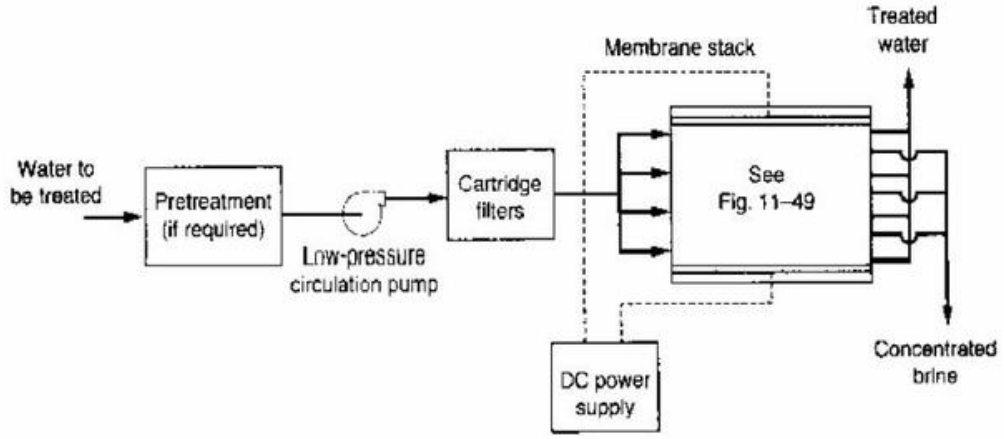

Table 11–28
Typical operating parameters for electrodialysis units

Parameter	Unit	Range
Detention time in stack	d	10–20
CD/N ratio	mA/cm^2	500–800
Membrane resistance, Ω	ohms	4–8
Salt-removal efficiency	%	40–60
Current efficiency	%	85–95
Concentrate stream flow	% of feed	10–20

equal flowrates and pressures on both sides of each membrane. Sulfuric acid is fed to the concentrate stream to maintain a low pH and thus minimize scaling. Typical operating parameters for the electrodialysis process are reported in Table 11–28.

Problems associated with the electrodialysis process for wastewater renovation include chemical precipitation of salts with low solubility on the membrane surface and clogging of the membrane by the residual colloidal organic matter in wastewater-treatment plant effluents. To reduce membrane fouling, activated carbon pretreatment, possibly preceded by chemical precipitation and some form of multimedia filtration, may be necessary.

Pilot Studies for Membrane Applications

Because every wastewater is unique with respect to its chemistry, it is difficult to predict a priori how a given membrane process will perform. As a result, the selection of the best membrane for a given application is usually based on the results of pilot studies. Membrane fouling indexes can be used to assess the need for pretreatment. In some

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situations, manufacturers of membranes will provide a testing service to identify the most appropriate membrane for a specific water or wastewater.

The elements that comprise a pilot plant include (1) the pretreatment system, (2) tankage for flow equalization and cleaning, (3) pumps for pressurizing the membrane, recirculation, and backflushing with appropriate controls (e.g., variable-frequency drives), (4) the membrane test module, (5) adequate facilities for monitoring the performance of the test module, and (6) an appropriate system for backflushing the membranes. The information collected should be sufficient to allow for the design of the full-scale system and should include as a minimum the following items:

Membrane operating parameters
- Pretreatment requirements including chemical dosages
- Transmembrane flux rate correlated to operating time
- Transmembrane pressure
- Washwater requirements
- Recirculation ratio
- Cleaning frequency including protocol and chemical requirements
- Posttreatment requirements

Typical water quality measurements may include:
- Turbidity
- Particle counts
- Total organic carbon
- Nutrients
- Heavy metals
- Organic priority pollutants
- Total dissolved solids
- pH
- Temperature
- Heterotrophic plate count (see Chap. 2)
- Other bacterial indicators
- The specific constituents that can limit recovery such as silica, barium, calcium, and sulfate
- Biotoxicity

Additional specific parameters selected for evaluation will depend on the final use to be made of the repurified water. Typical pilot scale test facilities used to evaluate various membrane treatment options in connection with the production of repurified water for indirect potable reuse are shown on Fig. 11–51.

Disposal of Concentrated Waste Streams

Disposal of the concentrated waste streams produced by membrane processes represents the major problem that must be dealt with in their applications. The principal methods now used for the disposal of the concentrated waste streams are reported in Table 11–29. While small facilities can dispose of small concentrated waste streams by blending with other wastewater flows, this approach is not suitable for large facilities. The concentrate from NF and RO facilities will contain hardness, heavy metals, high-molecular-weight organics, microorganisms, and often hydrogen sulfide gas. The pH is usually high due to the concentration of alkalinity, which increases the likelihood of metal precipitation in disposal wells. As a result, most of the large-scale desalination facilities are located along coastal regions, both in the United States and in other parts of the world. For inland

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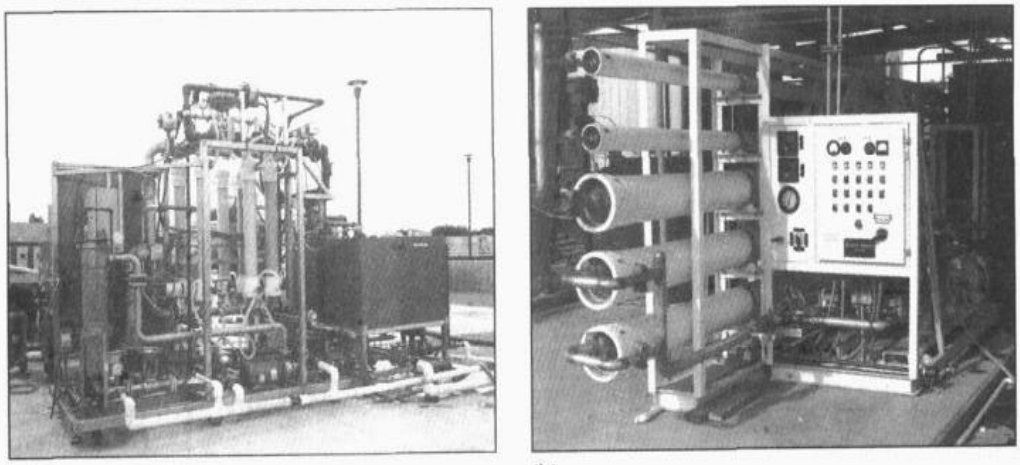
locations, long transmission lines to coastal regions are being considered. While controlled evaporation is technically feasible, because of high operating and maintenance costs this approach is used where no other alternatives are available, and the value of product water is high. The quality and quantity of the concentrated waste stream produced from nanofiltration, reverse osmosis, and electrodialysis can be estimated using simplified recovery and rejection computations as illustrated in Example 11–8.



(a) (b)

Figure 11–51
Typical pilot-plant facilities used to test membranes: (a) nanofiltration and (b) reverse osmosis.

Table 11–29
Disposal options for concentrated brine solutions from membrane processes

Disposal option	Description
Ocean discharge	The disposal option of choice for facilities located in the coastal regions of the United States. Typically, a brine line, with a deep ocean discharge, is used by a number of dischargers. Combined discharge with power-plant cooling water has been used in Florida. For inland locations truck, rail hauling, or pipeline is needed for transportation
Surface water discharge	Discharge of brines to surface waters is the most common method of disposal for concentrated brine solutions
Land application	Land application has been used for some low-concentration brine solutions
Discharge to wastewater-collection system	This option is suitable only for very small discharges such that the increase in TDS is not significant (e.g., less than 20 mg/L)
Deep-well injection	Depends on whether subsurface aquifer is brackish water or is otherwise unsuitable for domestic uses
Evaporation ponds	Large surface area required in most areas with the exception of some southern and western states
Controlled thermal evaporation	Although energy-intensive, thermal evaporation may be the only option available in many areas

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EXAMPLE 11–8 Estimate Quantity and Quality of Waste Streams from a Reverse Osmosis Facility Estimate quantity and quality of the waste stream, and the total quantity of water that must be processed, from a reverse osmosis facility that is to produce 4000 m^3/d of water to be used for industrial cooling operations. Assume that both the recovery and rejection rates are equal to 90 percent and that the concentration of the feed stream is 400 g/m^3.

Solution

1. Determine the flowrate of the concentrated waste stream and the total amount of water that must be processed.
   a. Combining Eqs. (11–41) and (11–39) results in the following expression for the concentrate stream flowrate:

$$Q_c = \frac{Q_p(1 - r)}{r}$$

   b. Determine the concentrate stream flowrate.

$$Q_c = \frac{(4000\ m^3/d)(1 - 0.9)}{0.9} = 444\ m^3/d$$

   c. Determine the total amount of water that must be processed to produce 4000 m^3/d of RO water. Using Eq. (11–41), the required amount of water is

$$Q_f = Q_p + Q_c = 4000\ m^3/d + 444\ m^3/d = 4444\ m^3/d$$

2. Determine the concentration of the permeate stream. The permeate concentration is obtained by writing Eq. (11–40) as follows:

$$C_p = C_f(1 - R) = 400\ g/m^3\ (1 - 0.9) = 40\ g/m^3$$

3. Determine the concentration of the concentrated waste stream. The required value is obtained by solving Eq. (11–42):

$$C_c = \frac{Q_f C_f - Q_p C_p}{Q_c}$$

$$C_c = \frac{(4444\ m^3/d)(400\ g/m^3) - (4000\ m^3/d)(40\ g/m^3)}{(444\ m^3/d)}$$

$$C_c = 3643\ g/m^3$$

Comment To reduce the volume of the waste stream that must be treated, a variety of concentration methods are currently under investigation.

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11-7 ADSORPTION

Adsorption is the process of accumulating substances that are in solution on a suitable interface. Adsorption, as noted in Chap. 4, is a mass transfer operation in that a constituent in the liquid phase is transferred to the solid phase. The *adsorbate* is the substance that is being removed from the liquid phase at the interface. The *adsorbent* is the solid, liquid, or gas phase onto which the adsorbate accumulates. Although adsorption is used at the air-liquid interface in the flotation process, only the case of adsorption at the liquid-solid interface will be considered in this discussion. The adsorption process has not been used extensively in wastewater treatment, but demands for a better quality of treated wastewater effluent, including toxicity reduction, have led to an intensive examination and use of the process of adsorption on activated carbon. Activated carbon treatment of wastewater is usually thought of as a polishing process for water that has already received normal biological treatment. The carbon in this case is used to remove a portion of the remaining dissolved organic matter. The purpose of this section is to introduce the basic concepts of adsorption and to consider carbon adsorption.

Types of Adsorbents

The principal types of adsorbents include activated carbon, synthetic polymeric, and silica-based adsorbents, although synthetic polymeric and silica-based adsorbents are seldom used for wastewater adsorption because of their high cost. Because activated carbon is used most commonly in advanced wastewater-treatment applications, the focus of the following discussion is on activated carbon. The nature of activated carbon, the use of granular carbon and powdered carbon for wastewater treatment, and carbon regeneration and reactivation are discussed below.

Activated Carbon. Activated carbon is prepared by first making a char from organic materials such as almond, coconut, and walnut hulls; other materials including woods, bone, and coal have also been used. The char is produced by heating the base material to a red heat (less than about 700°C) in a retort to drive off the hydrocarbons, but with an insufficient supply of oxygen to sustain combustion. The carbonization or char-producing process is essentially a pyrolysis process. The char particle is then *activated* by exposure to oxidizing gases such as steam and CO_2 at high temperatures, in the range from 800 to 900°C. These gases develop a porous structure, as illustrated previously on Fig. 4–32 in Chap. 4, in the char and thus creates a large internal surface area. The resulting pore sizes are defined as follows:

Macropores	>25 nm
Mesopores	>1 nm and <25 nm
Micropores	<1 nm

The surface properties that result are a function of both the initial material used and the preparation procedure, so that many variations are possible. The type of base material from which the activated carbon is derived may also affect the pore-size distribution and the regeneration characteristics. After activation, the carbon can be separated into, or prepared in, different sizes with different adsorption capacity. The two size classifications are *powdered activated carbon* (PAC), which typically has a diameter of less than 0.074 mm (200 sieve), and *granular activated carbon* (GAC), which has a diame-

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Table 11-30 Comparison of granular and powdered activated carbon

Parameter	Unit	Type of activated carbon[a] GAC	PAC
Total surface area	m^2/g	700–1300	800–1800
Bulk density	kg/m^3	400–500	360–740
Particle density, wetted in water	kg/L	1.0–1.5	1.3–1.4
Particle size range	mm (μm)	0.1–2.36	(5–50)
Effective size	mm	0.6–0.9	na
Uniformity coefficient	UC	≤1.9	na
Mean pore radius		16–30	20–40
Iodine number		600–1100	800–1200
Abrasion number	minimum	75–85	70–80
Ash	%	≤8	≤6
Moisture as packed	%	2–8	3–10

[a] Specific values will depend on the source material used for the production of the activated carbon.

ter greater than 0.1 mm (~140 sieve). The characteristics of granular and powdered activated carbon are summarized in Table 11-30.

Carbon Regeneration and Reactivation. Economical application of activated carbon depends on an efficient means of regenerating and reactivating the carbon after its adsorptive capacity has been reached. *Regeneration* is the term used to describe all of the processes that are used to recover the adsorptive capacity of the spent carbon, exclusive of reactivation, including: (1) chemicals to oxidize the adsorbed material, (2) steam to drive off the adsorbed material, (3) solvents, and (4) biological conversion processes. Typically some of the adsorptive capacity of the carbon (about 4 to 10 percent) is also lost in the regeneration process, depending on the compounds being adsorbed and the regeneration method used (Crittenden, 2000). In some applications, the capacity of the carbon following regeneration has remained essentially the same for years. A major problem with the use of powdered activated carbon is that the methodology for its regeneration is not well defined. The use of powdered activated carbon produced from recycled solid wastes may obviate the need to regenerate the spent carbon, and may be more economical. Additional details on carbon reactivation and regeneration may be found in Sontheimer et al. (1988).

Reactivation of granular carbon involves essentially the same process used to create the activated carbon from virgin material. Spent carbon is reactivated in a furnace by oxidizing the adsorbed organic material and, thus, removing it from the carbon surface. The following series of events occurs in the reactivation of spent activated carbon: (1) the carbon is heated to drive off the absorbed organic material (i.e., absorbate), (2) in the process of driving off the absorbed material some new compounds are formed that remain on the surface of the carbon, and (3) the final step in the reactivation process is to burn off the new compounds that were formed when the absorbed material was burned off. With effective process control, the adsorptive capacity of reactivated carbon

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will be essentially the same as that of the virgin carbon (Crittenden, 2000). For planning purposes, it is often assumed that a loss of 2 to 5 percent will occur in the reactivation process. It is important to note that most other losses of carbon occur through attrition due to mishandling. For example, right angle bends in piping cause attrition through abrasion and impact. The type of pumping facilities used will also affect the amount of attrition. In general, a 4 to 8 percent loss of carbon is assumed, due to handling. Replacement carbon must be available to make up the loss.

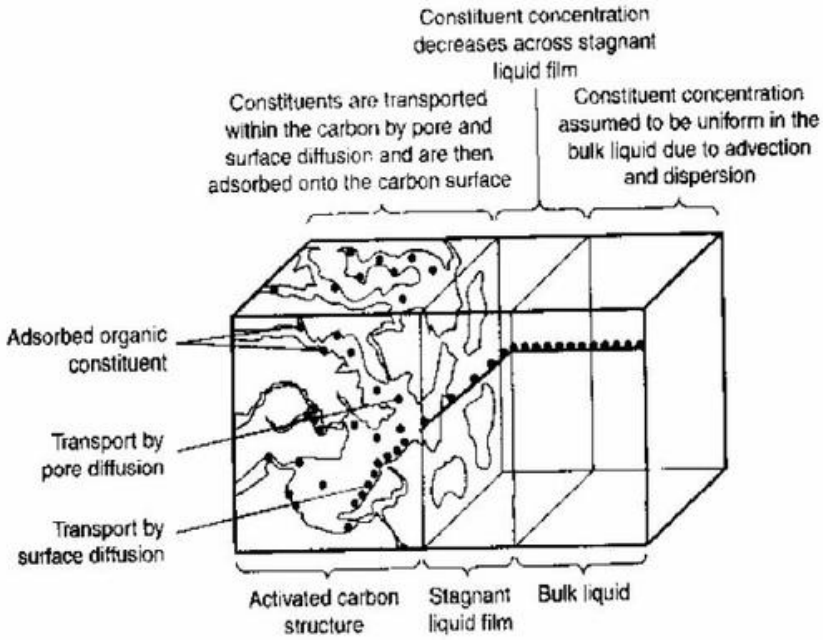
Fundamentals of Adsorption

The adsorption process, as illustrated on Fig. 11–52, takes place in four more or less definable steps: (1) bulk solution transport, (2) film diffusion transport, (3) pore transport, and (4) adsorption (or sorption). *Bulk solution transport* involves the movement of the organic material to be adsorbed through the bulk liquid to the boundary layer of fixed film of liquid surrounding the adsorbent, typically by advection and dispersion in carbon contactors. *Film diffusion transport* involves the transport by diffusion of the organic material through the stagnant liquid film to the entrance of the pores of the adsorbent. *Pore transport* involves the transport of the material to be adsorbed through the pores by a combination of molecular diffusion through the pore liquid and/or by diffusion along the surface of the adsorbent. *Adsorption* involves the attachment of the material to be adsorbed to adsorbent at an available adsorption site (Snoeyink and Summers, 1999). Adsorption can occur on the outer surface of the adsorbent and in the macropores, mesopores, micropores, and submicropores, but the surface area of the macro- and mesopores is small compared with the surface area of the micropores and submicropores and the amount of material adsorbed there is usually considered negligible. Adsorption forces include (Crittenden, 1999):

- Coulombic-unlike charges
- Point charge and a dipole
- Dipole-dipole interactions

Figure 11–52

Definition sketch for the adsorption of an organic constituent with activated carbon.



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- Point charge neutral species
- London or van der Waals forces
- Covalent bonding with reaction
- Hydrogen bonding

Additional details on the above adsorption forces may be found in Crittenden (1992). Because it is difficult to differentiate between chemical and physical adsorption, the term *sorption* is often used to describe the attachment of the organic material to the activated carbon.

Because the adsorption process occurs in a series of steps, the slowest step in the series is identified as the *rate limiting step*. In general, if physical adsorption is the principal method of adsorption, one of the diffusion transport steps will be the rate limiting, because the rate of physical adsorption is rapid. Where chemical adsorption is the principal method of adsorption, the adsorption step has often been observed to be rate limiting. When the rate of sorption equals the rate of desorption, equilibrium has been achieved and the capacity of the carbon has been reached. The theoretical adsorption capacity of the carbon for a particular contaminant can be determined by developing its adsorption isotherm as described below.

Development of Adsorption Isotherms. The quantity of adsorbate that can be taken up by an adsorbent is a function of both the characteristics and concentration of adsorbate and the temperature. The characteristics of the adsorbate that are of importance include: solubility, molecular structure, molecular weight, polarity, and hydrocarbon saturation (see Table 11-31). Generally, the amount of material adsorbed is determined as a function of the concentration at a constant temperature, and the resulting function is called an adsorption isotherm. Adsorption isotherms are developed by exposing a given amount of absorbate in a fixed volume of liquid to varying amounts of activated carbon. Typically, more than ten containers are used, and the minimum time allowed for the samples to equilibrate where powdered activated carbon is used is seven days. If granular activated carbon is used, it is usually powdered to minimize adsorption times. The test is illustrated previously on Fig. 4-34 in Chap. 4. At the end of the test period, the amount of absorbate remaining in solution is measured. The absorbent phase concentration after equilibrium is computed using Eq. (11-50). The absorbent phase concentration data computed using Eq. (11-50) are then used to develop adsorption isotherms as described below.

$$q_e = \frac{(C_o - C_e)V}{m} \tag{11-50}$$

where q_e = adsorbent (i.e., solid) phase concentration after equilibrium, mg adsorbate/g adsorbent
C_o = initial concentration of adsorbate, mg/L
C_e = final equilibrium concentration of adsorbate after absorption has occurred, mg/L
V = volume of liquid in the reactor, L
m = mass of adsorbent, g

Freundlich Isotherm. Equations that are often used to describe the experimental isotherm data were developed by Freundlich, Langmuir, and Brunauer, Emmet, and

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Table 11-31
Readily and poorly adsorbed organics[a]

Readily adsorbed organics	Poorly adsorbed organics
Aromatic solvents	Low-MW ketones, acids, and aldehydes
Benzene	Sugars and starches
Toluene	Very-high-MW or colloidal organics
Nitrobenzenes	Low-MW aliphatics
Chlorinated aromatics	
PCBs	
Chlorophenols	
Polynuclear aromatics	
Acenaphthene	
Benzopyrenes	
Pesticides and herbicides	
DDT	
Aldrin	
Chlordane	
Atrazine	
Chlorinated nonaromatics	
Carbon tetrachloride	
Chloroalkyl ethers	
Trichloroethene	
Chloroform	
Bromoform	
High-MW hydrocarbons	
Dyes	
Gasoline	
Amines	
Humics	

[a]From Froelich (1978).

Teller (BET isotherm) (Shaw, 1966). Of the three, the Freundlich isotherm is used most commonly to describe the adsorption characteristics of the activated carbon used in water and wastewater treatment. Derived empirically in 1912, the Freundlich isotherm is defined as follows:

$$\frac{x}{m} = K_f C_e^{1/n} \tag{11-51}$$

where x/m = mass of adsorbate adsorbed per unit mass of adsorbent, mg adsorbate/g activated carbon
K_f = Freundlich capacity factor, (mg absorbate/g activated carbon)(L water/mg adsorbate)$^{1/n}$
C_e = equilibrium concentration of adsorbate in solution after adsorption, mg/L
$1/n$ = Freundlich intensity parameter

The constants in the Freundlich isotherm can be determined by plotting log (x/m) versus log C_e and making use of Eq. (11-51) rewritten as:

$$\log\left(\frac{x}{m}\right) = \log K_f + \frac{1}{n} \log C_e \tag{11-52}$$

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The U.S. EPA has developed adsorption isotherms for a variety of toxic compounds, some of which are presented in Table 11–32 for reference (Dobbs and Cohen, 1980). As shown in Table 11–32, the variation in the Freundlich capacity factor for the various compounds is extremely wide (e.g., 14,000 for PCB to 6.8×10^{-5} for N-Dimethylnitrosamine). Because of the wide variation, the Freundlich capacity factor must be determined for each new compound. Application of the Freundlich adsorption isotherm is illustrated in Example 11–9.

Table 11–32 Freundlich adsorption isotherm constants for selected organic compounds[a,b]

Compound	pH	K_f(mg/g)(L/mg)$^{1/n}$	1/n
Benzene	5.3	1.0	1.6–2.9
Bromoform	5.3	19.6	0.52
Carbon tetrachloride	5.3	11	0.83
Chlorobenzene	7.4	91	0.99
Chloroethane	5.3	0.59	0.95
Chloroform	5.3	2.6	0.73
DDT	5.3	322	0.50
Dibromochloromethane	5.3	4.8	0.34
Dichlorobromomethane	5.3	7.9	0.61
1,2-Dichloroethane	5.3	3.6	0.83
Ethylbenzene	7.3	53	0.79
Heptachlor	5.3	1,220	0.95
Hexachloroethane	5.3	96.5	0.38
Methylene chloride	5.3	1.3	1.16
N-Dimethylnitrosamine	na	6.8×10^{-5}	6.60
N-Nitrosodi-n-propylamine	na	24	0.26
N-Nitrosodiphenylamine	3–9	220	0.37
PCB	5.3	14,100	1.03
PCB 1221	5.3	242	0.70
PCB 1232	5.3	630	0.73
Phenol	3–9	21	0.54
Tetrachloroethylene	5.3	51	0.56
Toluene	5.3	26.1	0.44
1,1,1-Trichloroethane	5.3	2–2.48	0.34
Trichloroethylene	5.3	28	0.62

[a] Adapted from Dobbs and Cohen (1980) and LaGrega et al. (2001).

[b] The adsorption isotherm constants reported in this table are meant to be illustrative of the wide range of values that will be encountered for various organic compounds. It is important to note that the characteristics of the activated carbon used as well as the analytical technique used for the analysis of the residual concentrations of the individual compounds will have a significant effect on the coefficient values obtained for specific organic compounds.

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Langmuir Isotherm. Derived from rational considerations, the Langmuir adsorption isotherm is defined as:

$$\frac{x}{m} = \frac{abC_e}{1 + bC_e} \tag{11-53}$$

where x/m = mass of adsorbate adsorbed per unit mass of adsorbent, mg adsorbate/g activated carbon
a, b = empirical constants
C_e = equilibrium concentration of adsorbate in solution after adsorption, mg/L

The Langmuir adsorption isotherm was developed by assuming: (1) a fixed number of accessible sites are available on the adsorbent surface, all of which have the same energy, and (2) adsorption is reversible. Equilibrium is reached when the rate of adsorption of molecules onto the surface is the same as the rate of desorption of molecules from the surface. The rate at which adsorption proceeds is proportional to the driving force, which is the difference between the amount adsorbed at a particular concentration and the amount that can be adsorbed at that concentration. At the equilibrium concentration, this difference is zero.

Correspondence of experimental data to the Langmuir equation does not mean that the stated assumptions are valid for the particular system being studied, because departures from the assumptions can have a canceling effect. The constants in the Langmuir isotherm can be determined by plotting $C_e/(x/m)$ versus C_e and making use of Eq. (11–53) rewritten as:

$$\frac{C_e}{(x/m)} = \frac{1}{ab} + \frac{1}{a}C_e \tag{11-54}$$

Application of the Langmuir adsorption isotherm is illustrated in Example 11–9.

EXAMPLE 11-9 Analysis of Activated-Carbon Adsorption Data Determine the Freundlich and Langmuir isotherm coefficients for the following GAC adsorption test data. The liquid volume used in the batch adsorption tests was 1 L. The initial concentration of the adsorbate in solution was 3.37 mg/L. Equilibrium was obtained after 7 days.

Mass of GAC, m, g	Equilibrium concentration of adsorbate in solution, C_e, mg/L
0.0	3.37
0.001	3.27
0.010	2.77
0.100	1.86
0.500	1.33

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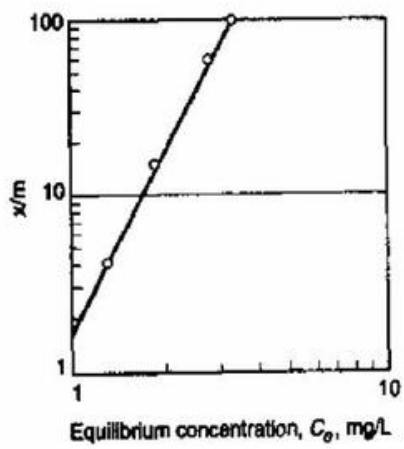
Solution

1. Derive the values needed to plot the Freundlich and Langmuir adsorption isotherms using the batch adsorption test data.

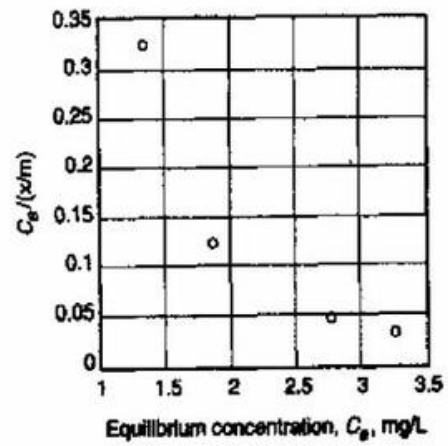
Adsorbate concentration, mg/L					
C_o	C_e	$C_o - C_e$	m, g	x/m,[a] mg/g	$C_e/(x/m)$
3.37	3.37	0.00	0.000	—	—
3.37	3.27	0.10	0.001	100	0.0327
3.37	2.77	0.60	0.010	60	0.0462
3.37	1.86	1.51	0.100	15.1	0.1232
3.37	1.33	2.04	0.500	4.08	0.3260

[a] $q_e = \frac{x}{m} = \frac{(C_o - C_e)V}{m}$

2. Plot the Freundlich and Langmuir adsorption isotherms using the data developed in Step 1 (see following figures).



(a)



(b)

3. Determine the adsorption isotherm coefficients.

   a. Freundlich

   When x/m versus C_e is plotted on log-log paper, the intercept when $C_e = 1.0$ is the value of (x/m) and the slope of the line is equal to $1/n$. Thus, $x/m = 1.55$, and $K_f = 1.55$. When $x/m = 1.0$, $C_e = 0.9$, and $1/n = 0.26$. Thus,

$$\frac{x}{m} = 1.5\, C_e^{0.26}$$

   b. Langmuir

   Because the plot for the Langmuir isotherm is curvilinear, use of the Langmuir adsorption isotherm is inappropriate.

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Adsorption of Mixtures. In the application of adsorption to wastewater treatment, mixtures of organic compounds are always encountered. Typically, there is a depression of the adsorptive capacity of any individual compound in a solution of many compounds, but the total adsorptive capacity of the adsorbent may be larger than the adsorptive capacity with a single compound. The amount of inhibition due to competing compounds is related to the size of the molecules being adsorbed, their adsorptive affinities, and their relative concentrations. It is important to note that adsorption isotherms can be determined for a heterogeneous mixture of compounds including total organic carbon (TOC), dissolved organic carbon (DOC), chemical oxygen demand (COD), dissolved organic halogen (DOH), UV absorbance, and fluorescence (Snoeyink and Summers, 1999). The adsorption from mixtures is considered further in Crittenden et al. (1987b, 1987c, 1985) and Sontheimer (1988).

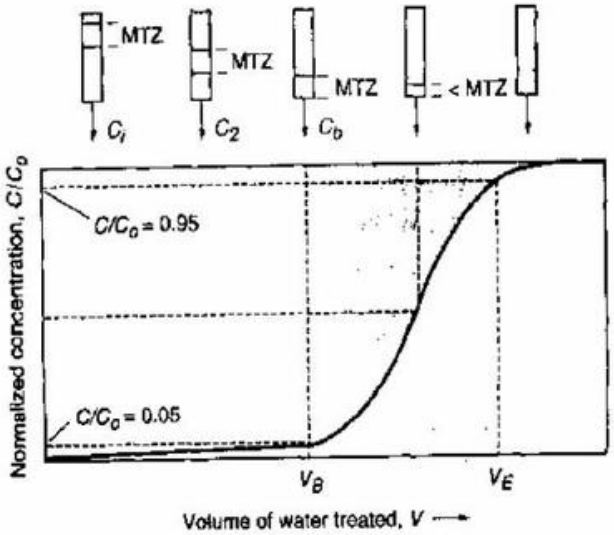
Activated Carbon Adsorption Kinetics

As noted previously, both granular carbon (in downflow and upflow columns) and powdered activated carbon are used for wastewater treatment. The analysis procedures for both types are described briefly in the following discussion.

Mass Transfer Zone. The area of the GAC bed in which sorption is occurring is called the mass transfer zone (MTZ) (see Fig. 11–53). After the water containing the constituent to be removed passes through a region of the bed whose depth is equal to the MTZ, the concentration of the contaminant in the water will have been reduced to its minimum value. No further adsorption will occur within the bed below the MTZ. As the top layers of carbon granules become saturated with organic material, the MTZ will move down in the bed until breakthrough occurs. Typically, breakthrough is said to have occurred when the effluent concentration reaches 5 percent of the influent value. Exhaustion of the adsorption bed is assumed to have occurred when the effluent concentration is equal to 95 percent of the influent concentration. The length of the MTZ is typically a function of the hydraulic loading rate applied to the column and the characteristics of the activated carbon. In the extreme, if the loading rate is too great the

Figure 11–53

Typical breakthrough curve for activated carbon showing movement of mass transfer zone (MTZ) with throughput volume.



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height of the MTZ will be larger than the GAC bed depth, and the adsorbable constituent will not be removed completely by the carbon. At complete exhaustion, the effluent concentration is equal to the influent concentration.

In addition to the applied hydraulic loading rate, the shape of the breakthrough curve will also depend on whether the applied liquid contains nonadsorbable and biodegradable constituents. The impact of the presence of nonadsorbable and biodegradable organic constituents on the shape of the breakthrough curve is illustrated on Fig. 11-54. As shown on Fig. 11-54, if the liquid contains nonadsorbable constituents, the nonadsorbable constituents will appear in the effluent as soon as the carbon column is put into operation. If adsorbable and biodegradable constituents are present in the applied liquid, the breakthrough curve will not reach a C/C_o value of 1.0, but will be depressed, and the observed C/C_o value will depend on the biodegradability of the influent constituents, because biological activity continues even though the adsorption capacity has been utilized. If the liquid contains nonadsorbable and biodegradable constituents, the observed breakthrough curve will not start at zero and will not terminate at a value of 1.0 (Snoeyink and Summers, 1999). The above effects are observed quite commonly in wastewater adsorption applications, especially with respect to the removal of COD.

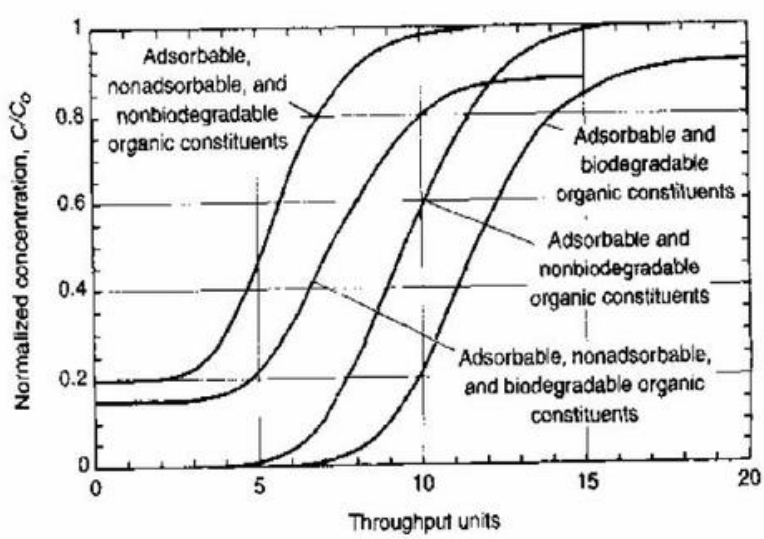
The height of the MTZ (see Fig. 11-53) varies with the flowrate because dispersion, diffusion, and channeling in a granular medium are directly related to the flowrate. For a symmetrical breakthrough curve it has been shown that the height of the mass transfer zone, H_{MTZ}, is related to the column height Z and the throughput volumes V_B and V_E, as follows (Michaels, 1952; Weber, 1972).

$$H_{MTZ} = Z\left[\frac{V_E - V_B}{V_E - 0.5(V_E - V_B)}\right] \tag{11-55}$$

where H_{MTZ} = length of mass transfer zone, m
Z = height of the adsorption column, m
V_E = throughput volume to exhaustion, L, m^3
V_B = throughput volume to breakthrough L, m^3

Figure 11-54

Impact of the presence of adsorbable, nonadsorbable, and biodegradable organic constituents on the shape of the activated carbon breakthrough curve. *(Adapted from Snoeyink and Summers, 1999.)*



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The area above the breakthrough curve represents the mass of adsorbate adsorbed within the column and is equal to:

$$\int (C_o - C)dV$$

integrated from $V = 0$ to $V = V_b$ the throughput volume at breakthrough. At complete exhaustion, $C = C_o$.

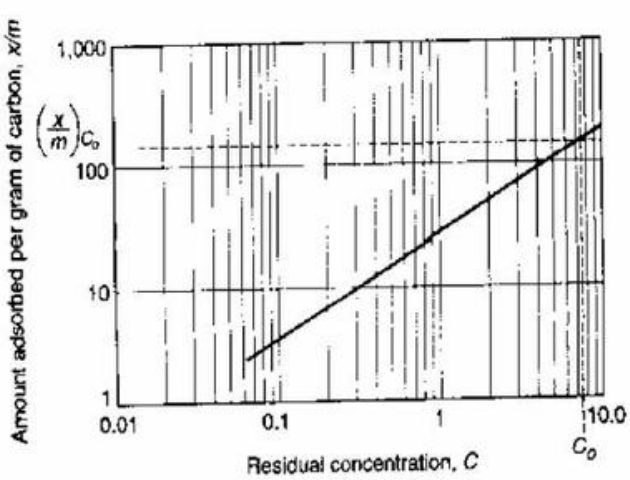
In practice, the only way to use the capacity at the bottom of the carbon adsorption column is to have two or more columns in series and switch them as they are exhausted, or to use multiple columns in parallel so that breakthrough in one column does not affect effluent quality. The optimum flowrate and bed depth, as well as the operating capacity of the carbon, must be established to determine the dimensions and the number of columns necessary for continuous treatment. These parameters can be determined from dynamic column tests, as discussed below.

Carbon Adsorption Capacity. The adsorptive capacity of a given carbon is estimated from isotherm data as follows. If isotherm data are plotted, the resulting isotherm will be as shown on Fig. 11–55. Referring to Fig. 11–55, the adsorptive capacity of the carbon can be estimated by extending a vertical line from the point on the horizontal axis corresponding to the initial concentration C_o, and extrapolating the isotherm to intersect this line. The $q_e = (x/m)_{C_o}$ value at the point of intersection can be read from the vertical axis. The $(q_e)_{C_o}$ value represents the amount of constituent adsorbed per unit weight of carbon when the carbon is at equilibrium with the initial concentration of constituent C_o. The equilibrium condition generally exists in the upper section of a carbon bed during column treatment, and it therefore represents the ultimate capacity of the carbon for a particular waste.

Breakthrough Adsorption Capacity. In the field, the breakthrough adsorption capacity $(x/m)_b$ of the GAC in a full-scale column is some percentage of the theoretical adsorption capacity found from the isotherm. The $(x/m)_b$ of a single column can be assumed to be approximately 25 to 50 percent of the theoretical capacity, $(x/m)_o$. The value of $(x/m)_b$ can be determined using the small-scale column test described later in

Figure 11–55

Typical activated carbon adsorption isotherm.



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this section. Once $(x/m)_b$ is known, the time to breakthrough can be approximated by solving the following equation for t_b.

$$\left(\frac{x}{m}\right)_b = \frac{x_b}{m_{GAC}} = Q\left(C_o - \frac{C_b}{2}\right)\frac{t_b}{m_{GAC}} \tag{11-56}$$

Where $(x/m)_b$ = field breakthrough adsorption capacity, g/g
x_b = mass of organic material adsorbed in the GAC column at breakthrough, g
m_{GAC} = mass of carbon in the column, g
Q = flowrate, m^3/d
C_o = influent organic concentration, g/m^3
C_b = breakthrough organic concentration, g/m^3
t_b = time to breakthrough, d

Equation (11-56) was developed assuming that C_o is constant and that the effluent concentration increases linearly with time from 0 to C_b (see Fig. 11-53). The term $(C_o - C_b/2)$ represents the average concentration of the organic matter adsorbed up to the breakthrough point. Rearranging Eq. (11-56), the time to breakthrough can be calculated using the following relationship:

$$t_b = \frac{(x/m)_b m_{GAC}}{Q(C_o - C_b/2)} \tag{11-57}$$

Over the years, a number of equations have been developed to describe the breakthrough curve, including those by Bohart and Adams (1920) and Crittenden et al. (1987*a*).

However, as noted previously, because of the breakthrough phenomenon (see Fig. 11-55), the usual practice is either to use *two or more columns in series and rotate them as they become exhausted, or to use multiple columns in parallel so that breakthrough in a single column will not significantly affect the effluent quality.* With proper sampling from points within the column, constituent (e.g., TOC) breakthrough can be anticipated.

Activated Carbon Treatment Process Applications

Carbon adsorption is used principally for the removal of refractory organic compounds, as well as residual amounts of inorganic compounds such as nitrogen, sulfides, and heavy metals. The removal of taste and odor compounds from wastewater is another important application, especially in reuse applications. Both powdered and granular activated carbon are used and appear to have a low adsorption affinity for low molecular weight polar organic species. If biological activity is low in the carbon contactor or in other biological unit processes, these species are difficult to remove with activated carbon. Typical compounds that can be removed by carbon adsorption are reported in Table 11-31. Under normal conditions, after treatment with carbon, the effluent BOD ranges from 2 to 7 mg/L, and the effluent COD ranges from 10 to 20 mg/L. Under optimum conditions, it appears that the effluent COD can be reduced to less than 10 mg/L.

Treatment with Granular Activated Carbon (GAC). Treatment with GAC involves passing a liquid to be treated through a bed of activated carbon held in a reactor (sometimes called a contactor). Several types of activated carbon contactors are used

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for advanced wastewater treatment. Typical systems may be either pressure or gravity type, and may be downflow or upflow fixed-bed units having two or three columns in series, or expanded bed upflow-countercurrent type. Typical schematic diagrams of carbon contactors are shown on Fig. 11–56.

Fixed-Bed A fixed-bed column is used most commonly for contacting wastewater with GAC. Fixed-bed columns can be operated singly, in series, or in parallel. A typical granular activated carbon contactor is shown on Fig. 11–57*a*. Granular-medium filters are commonly used upstream of the activated carbon contactors to remove the organics associated with the suspended solids present in secondary effluent. The water to be treated is applied to the top of the column and withdrawn at the bottom. The carbon is held in place with an underdrain system at the bottom of the column. Provision for backwashing and surface washing is often provided in wastewater applications to limit the headloss buildup due to the removal of particulate suspended solids within the carbon column. Unfortunately, backwashing has the effect of destroying the adsorption front as discussed later.

The advantage of a downflow design is that adsorption of organics and filtration of suspended solids are accomplished in a single step. Although upflow fixed-bed reactors have been used, downflow beds are used more commonly to lessen the chance of accumulating particulate material in the bottom of the bed, where the particulate material would be difficult to remove by backwashing. If soluble organic removal is not maintained at a high level, more frequent regeneration of the carbon may be required. Lack of consistency in pH, temperature, and flowrate may also affect performance of carbon contactors. A typical column installation used for the treatment of wastewater is shown on Fig. 11–57*b*.

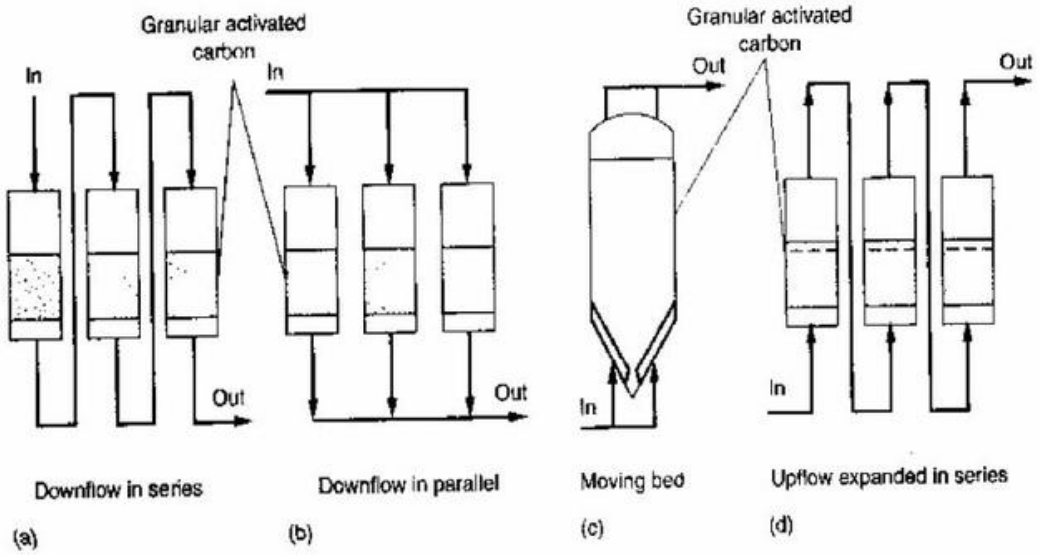

Figure 11–56

Types of activated carbon contactors: (*a*) downflow in series, (*b*) downflow in parallel, (*c*) moving bed, and (*d*) upflow expanded in series. *(Adapted from Calgon Carbon Corp.)*

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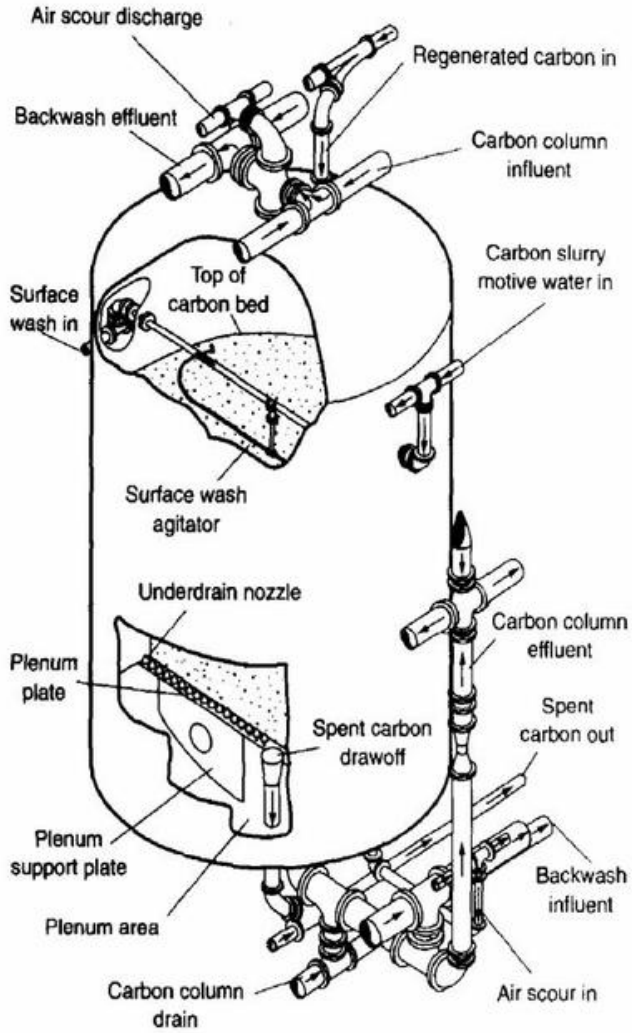
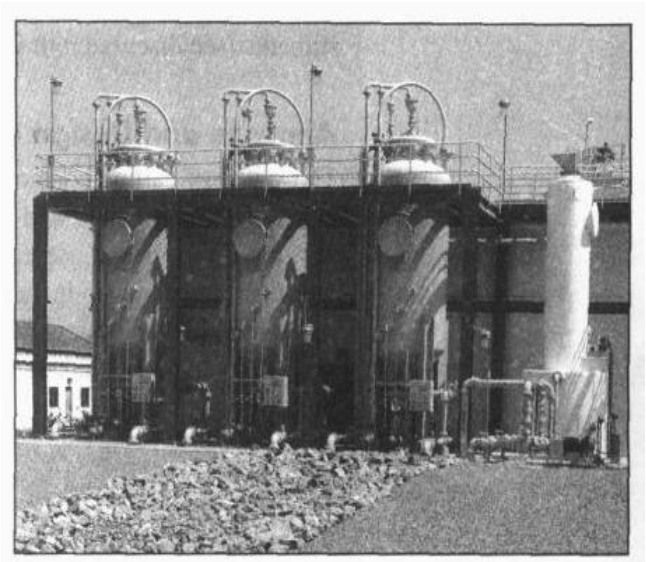


(a)



(b)

Figure 11-57

Activated carbon adsorbers: (a) Typical activated-carbon contactor in a pressure vessel and (b) view of granular activated carbon contactors operated in parallel, used for the treatment of filtered secondary effluent.

Expanded-Bed Expanded-bed, moving-bed, and pulsed-bed carbon contactors have also been developed to overcome the problems associated with headloss buildup. In the expanded-bed system, the influent is introduced at the bottom of the column and the activated carbon is allowed to expand, much as a filter bed expands during backwash. When the adsorptive capacity of the carbon at the bottom of the column is exhausted, the bottom portion of carbon is removed, and an equivalent amount of regenerated or virgin carbon is added to the top of the column. In such a system, headloss does not build up with time after the operating point has been reached. In general, expanded-bed upflow contactors may have more carbon fines in the effluent than downflow contactors because bed expansion leads to the creation of fines as the carbon particles collide and abrade, and allows the fines to escape through passageways created by the expanded bed. At present, few, if any, expanded bed contactors are used for the treatment of wastewater.

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Treatment with Powdered Activated Carbon (PAC). An alternative means of achieving adsoption is through the application of powdered activated carbon (PAC). Powdered activated carbon can be applied to the effluent from biological treatment processes, directly to the various biological treatment processes, and in physical-chemical treatment process flow diagrams. In the case of biological treatment plant effluent, PAC is added to the effluent in a contacting basin. After a certain amount of time for contact, the carbon is allowed to settle to the bottom of the tank, and the treated water is then removed from the tank. Because carbon is very fine, a coagulant, such as a polyelectrolyte, may be needed to aid the removal of the carbon particles, or filtration through rapid sand filters may be required. The addition of PAC directly to the aeration basin of an activated-sludge treatment process has proved to be effective in the removal of a number of soluble refractory organics. In physical-chemical treatment processes, PAC is used in conjunction with chemicals used for the precipitation of specific constituents (see discussion in Chap. 6).

Analysis and Design of Granular Activated Carbon Contactor

The sizing of carbon contactors is based on four factors: contact time, hydraulic loading rate, carbon depth, and number of contactors. Typical design information for the first three factors is presented in Table 11–33. Typical specifications for granular activated carbon are given in Table 11–34. A minimum of two parallel carbon contactors is recommended for design. Multiple units permit one or more units to remain in operation while one unit is taken out of service for removal and regeneration of spent carbon, or for maintenance.

Table 11–33 Typical design values for GAC contactors[a]

Parameter	Symbol	Unit	Value
Volumetric flowrate	V	m^3/h	50–400
Bed volume	V_b	m^3	10–50
Cross-sectional area	A_b	m^2	5–30
Length	D	m	1.8–4
Void fraction	α	m^3/m^3	0.38–0.42
GAC density	ρ	kg/m^3	350–550
Approach velocity	V_f	m/h	5–15
Effective contact time	t	min	2–10
Empty bed contact time	EBCT	min	5–30
Operation time	t	d	100–600
Throughput volume	V_L	m^3	10–100
Specific throughput	V_{sp}	m^3/kg	50–200
Bed volumes	BV	m^3/m^3	2,000–20,000

[a]Adapted from Sontheimer et al. (1998).

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For the case where the mass transfer rate is fast and the mass transfer zone is a sharp wave front, a steady-state mass balance around a carbon contactor (see Fig. 11–58) reactor yields:

Accumulation = inflow − outflow − amount absorbed

$$0 = QC_o t - QC_e t - m_{GAC} q_e \tag{11-58}$$

where Q = volumetric flowrate, L/h
C_o = initial concentration of adsorbate, mg/L
t = time, h
C_e = final equilibrium concentration of adsorbate, mg/L
m_{GAC} = mass of adsorbent, g
q_e = adsorbent phase concentration after equilibrium, mg adsorbate/g adsorbent

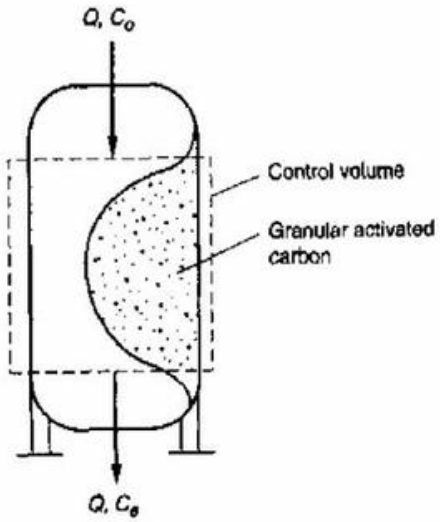
Table 11–34 Typical specifications for GAC

Parameter	Unit	Value
Total surface area	m^2/g	700–1300[a]
Bulk density	kg/m^3	400–440[a]
Particle density, wetted in water	kg/L	1.3–1.5[a]
Effective size	mm	0.8–0.9[a]
Uniformity coefficient	UC	≤1.9
Mean particle diameter	mm	1.5–1.7
Iodine number		850 min
Abrasion number		70 min
Ash	%	8 max
Moisture	%	4–6 max

[a] Depends on the source material for the carbon.

Figure 11–58

Definition sketch for the analysis of activated-carbon adsorption.



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From Eq. (11–58), the adsorbent usage rate is defined as

$$\frac{m_{GAC}}{Qt} = \frac{C_o - C_e}{q_e} \tag{11–59}$$

If it is assumed that the mass of the adsorbate in the pore space is small compared to the amount adsorbed, then the term QC_et in Eq. (11–58) can be neglected without serious error and the absorbent usage rate is given by:

$$\frac{m_{GAC}}{Qt} \approx \frac{C_o}{q_e} \tag{11–60}$$

To quantify the operational performance of GAC contactors, the following terms have been developed and are used commonly.

1. Empty-bed contact time (EBCT):

$$\text{EBCT} = \frac{V_b}{Q} = \frac{A_bD}{v_fA_b} = \frac{D}{v_f} \tag{11–61}$$

where EBCT = empty bed contact time, h
V_b = volume of GAC in contactor, m^3
Q = volumetric flowrate, m^3/h
A_b = cross-sectional area of GAC filter bed, m^2
D = length of GAC in contactor, m
v_f = linear approach velocity, m/h

2. Activated carbon density:
The density of the activated carbon is defined as

$$\rho_{GAC} = \frac{m_{GAC}}{V_b} \tag{11–62}$$

where ρ_{GAC} = density of granular activated carbon, g/L
m_{GAC} = mass of granular activated carbon, g
V_b = volume of GAC filter bed, L

3. Specific throughput, expressed as m^3 of water treated per gram of carbon:

$$\text{Specific throughput, } m^3/g = \frac{Qt}{m_{GAC}} = \frac{V_bt}{\text{EBCT} \times m_{GAC}} \tag{11–63}$$

Using Eq. (11–62), Eq. (11–63) can be written as

$$\text{Specific throughput} = \frac{V_bt}{\text{EBCT}(\rho_{GAC} \times V_b)} = \frac{t}{\text{EBCT}\rho_{GAC}} \tag{11–64}$$

4. Carbon usage rate (CUR) expressed as gram of carbon per m^3 of water treated:

$$\text{CUR, } g/m^3 = \frac{m_{GAC}}{Qt} = \frac{1}{\text{Specific throughput}} \tag{11–65}$$

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5. Volume of water treated for a given EBCT, expressed in liters, L:

$$\text{Volume of water treated, L} = \frac{\text{Mass of GAC for given EBCT}}{\text{GAC usage rate}} \tag{11-66}$$

6. Bed life, expressed in days, d:

$$\text{Bed life, d} = \frac{\text{Volume of water treated for given EBCT}}{Q} \tag{11-67}$$

The application of these terms is illustrated in Example 11–10.

EXAMPLE 11–10 Estimation of Activated Carbon Adsorption Breakthrough Time A fixed-bed activated carbon adsorber has a fast mass transfer rate, and the mass transfer zone is essentially a sharp wavefront. Assuming the following data apply, determine the carbon requirements to treat a flow of 1000 L/min, and the bed life.

1. Compound = trichloroethylene (TCE)
2. Initial concentration, C_o = 1.0 mg/L
3. Final concentration C_e = 0.005 mg/L
4. GAC density = 450 g/L
5. Freundlich capacity factor K_f = 28 (mg/g)(L/mg)$^{1/n}$ (see Table 11–32)
6. Freundlich intensity parameter, $1/n$ = 0.62 (see Table 11–32)
7. EBCT = 10 min

Solution

1. Estimate the GAC usage rate for toluene. The GAC usage rate is estimated using Eq. (11–59).

$$\frac{m_{GAC}}{Qt} = \frac{C_o - C_e}{q_e} = \frac{C_o - C_e}{K_f C_o^{1/n}}$$

$$= \frac{(1.0\ \text{mg/L})}{28\,(\text{mg/g})\,(\text{L/mg})^{0.62}(1.0\ \text{mg/L})^{0.62}}$$

$$= 0.036\ \text{g GAC/L}$$

2. Determine the mass of carbon required for a 10 min EBCT.

The mass of GAC in the bed = $V_b\rho_{GAC}$ = EBCT × Q × ρ_{GAC}

Carbon required = 10 min (1000 L/min) (450 g/L) = 4.5×10^6 g

3. Determine the volume of water treated using a 10 min EBCT.

$$\text{Volume of water treated} = \frac{\text{mass of GAC for given EBCT}}{\text{GAC usage rate}}$$

$$\text{Volume of water treated} = \frac{4.5 \times 10^6\ \text{g}}{(0.036\ \text{g GAC/L})} = 1.26 \times 10^8\ \text{L}$$

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4. Determine the bed life.

$$\text{Bed life} = \frac{\text{volume of water treated for given EBCT}}{Q}$$

$$\text{Bed life} = \frac{1.26 \times 10^8\ \text{L}}{(1000\ \text{L/min})(1440\ \text{min/d})} = 87.5\ \text{d}$$

Comment In this example, the full capacity of the carbon in the contactor was utilized based on the assumption that two columns in series will be used. If a single column is to be used, a breakthough curve must be used to arrive at the bed life. Biological activity within the column was not considered.

Small-Scale Column Tests

Over the years, a number of small-scale column tests have been developed to simulate the results obtained with full-scale reactors. One of the early column tests was the *high-pressure minicolumn* (HPMC) technique developed by Rosene et al. (1983), and later modified by Bilello and Beaudet (1983). In the HPMC test procedure, a high-pressure liquid chromatography column loaded with activated carbon is used. Typically the HPMC test procedure is used to determine the capacity of activated carbon for the adsorption of volatile organic compounds. The principal advantage of the HPMC test procedure is that it allows for the rapid determination of the GAC adsorptive capacity under conditions similar to those encountered in the field.

An alternative procedure known as the *rapid small-scale column test* (RSSCT) has been developed by Crittenden et al. (1991). The test procedure allows for the scaling of data obtained from small columns (see Fig. 11-59) to predict the performance of pilot or full-scale carbon columns. In developing the procedure, mathematical models were used to define the relationships between the breakthrough curve for small and large columns. In adsorption columns, the mass transfer mechanisms that are responsible for the spreading of the mass transfer zone are (1) dispersion, (2) film diffusion, and (3) interparticle diffusion. Two different design relationships were developed, one for constant diffusivity and one for proportional diffusivity. In the constant diffusivity model, it is assumed that dispersion is negligible because the hydraulic loading rate is high in the RSSCT, and that mass transfer occurs as a result of film diffusion. Further, it is assumed that the interparticle diffusivity is the same for both the small and large columns. In the proportional diffusivity model, it is assumed that dispersion is negligible because the hydraulic loading rate is high in the RSSCT, and that mass transfer occurs as a result of interparticle diffusion. The relationships for the two cases can be generalized as follows:

$$\frac{\text{EBCT}_{SC}}{\text{EBCT}_{LC}} = \left[\frac{d_{SC}}{d_{LC}}\right]^{2-x} = \frac{t_{SC}}{t_{LC}} \tag{11-68}$$

$$\frac{V_{SC}}{V_{LC}} = \frac{d_{LC}}{d_{SC}} \tag{11-69}$$

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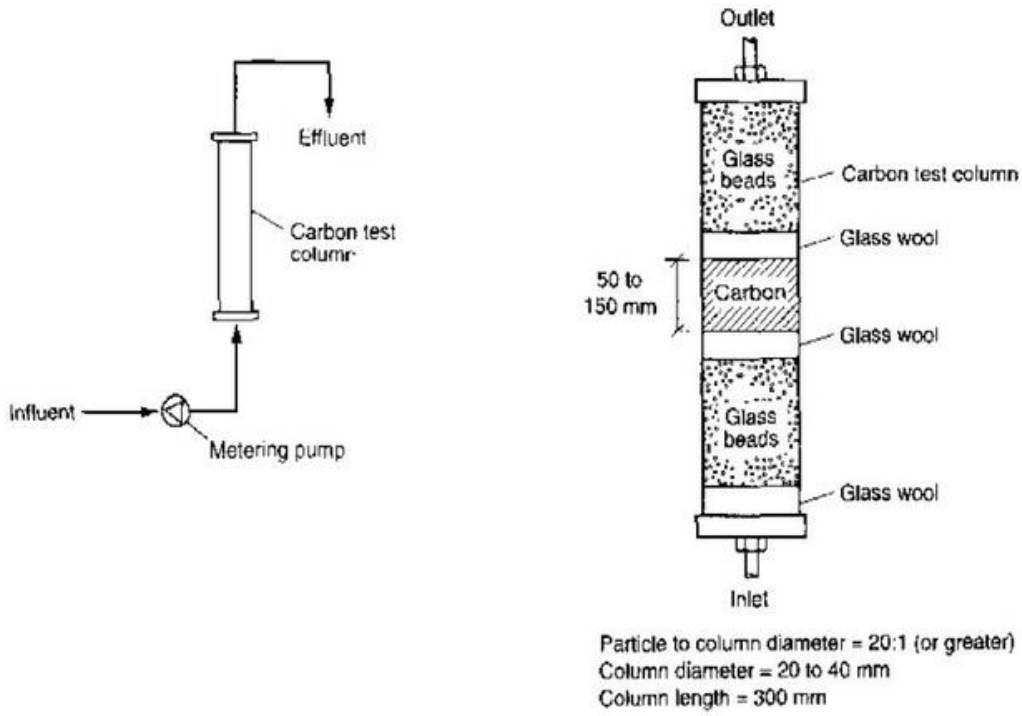
Figure 11-59

Small-scale column used to develop data for pilot- or full-scale carbon columns. *(Adopted from Crittenden et al., 1991.)*



where d_{SC} = diameter of particle in short column, mm
d_{LC} = diameter of particle in long column, mm
t_{SC} = time in short column, min
t_{LC} = time in long column, min
V_{SC} = superficial velocity in short column, m/h
V_{LC} = superficial velocity in long column, m/h

For constant and proportional diffusivity, the value of x in the exponent in Eq. (11–68) is 0 and 1, respectively. The application of the above equations is illustrated in Example 11–11.

EXAMPLE 11–11 **Comparison of Rapid Small-Scale Column Test Parameters to Pilot-Scale Parameters** Determine the corresponding parameters for an RSSCT based on the following data proposed for a pilot-scale column. Assume that film diffusion is the controlling mechanism.

Parameter	Unit	Pilot column (LC)	RSSCT (SC)
Particle diameter	mm	0.5	0.1
Carbon density	g/L	450	450
EBCT	min	10	
Loading rate	m/h	5.0	
Flowrate	mL/min	200	

(continued)

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(Continued)

Parameter	Unit	Pilot column (LC)	RSSCT (SC)
Column diameter	mm	75	10[a]
Column length	mm	1000	
Mass adsorbent	g		
Time of operation	d	100	
Water volume	L	28,800	

[a] Assumed value for small column.

Solution

1. Estimate the EBCT for the RSSCT.

$$\text{EBCT}_{SC} = \text{EBCT}_{LC}\left[\frac{d_{SC}}{d_{LC}}\right]^2$$

$$\text{EBCT}_{SC} = 10\text{ min}\left[\frac{0.1}{0.5}\right]^2 = 0.4\text{ min}$$

2. Estimate the loading rate for the RSSCT.

$$V_{SC} = V_{LC}\frac{d_{LC}}{d_{SC}}$$

$$V_{SC} = 5\text{ m/h}\,\frac{0.5}{0.1} = 25\text{ m/h}$$

3. Estimate flowrate for the RSSCT.

$$A = \frac{\pi}{4}d_{SC}^2 = \frac{\pi}{4}(10\text{ mm})^2 = 78.5\text{ mm}^2$$

$$Q_{SC} = (V_{SC})(A)$$

$$Q_{SC} = \frac{(25\text{ m/h})(10^3\text{ mm/m})(78.5\text{ mm}^2)}{(60\text{ min/h})(10^3\text{ mm}^3/\text{mL})} = 32.7\text{ mL/min}$$

4. Estimate column length for the RSSCT.

$$L_{SC} = \frac{Q_{SC}\times\text{EBCT}_{SC}}{A} = \frac{(32.7\text{ mL/min})(0.4\text{ min})}{(1\text{ mL}/10^3\text{ mm}^3)(78.5\text{ mm}^2)} = 166.7\text{ mm}$$

5. Estimate mass of adsorbent required for the RSSCT.

$$M_{SC} = \text{EBCT}_{LC}\left[\frac{d_{SC}}{d_{LC}}\right]^2(Q_{SC})(\rho_{SC})$$

$$M_{SC} = 10\text{ min}\left[\frac{0.1\text{ mm}}{0.5\text{ mm}}\right]^2\left[\frac{(32.7\text{ mL/min})(450\text{ g/L})}{10^3\text{ mL/L}}\right] = 5.9\text{ g}$$

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6. Estimate time of operation for the RSSCT.

$$t_{SC} = t_{LC}\frac{\text{EBCT}_{SC}}{\text{EBCT}_{LC}}$$

$$t_{SC} = 100\text{ d}\,\frac{0.4\text{ min}}{10\text{ min}} = 4\text{ d}$$

7. Estimate volume of water required for the RSSCT.

$$V_W = Q_{SC} \times t_{SC}$$

$$V_W = \frac{(32.7\text{ mL/min})(4\text{ d})(1440\text{ min/d})}{(10^3\text{ mL/L})} = 188.4\text{ L}$$

8. Summarize the findings for the RSSCT.

Parameter	Unit	Pilot column	RSSCT
Particle radius	mm	0.5	0.1
Carbon density	g/L	450	450
EBCT	min	10	0.4
Loading rate	m/h	5.0	25.0
Flowrate	mL/min	200	32.7
Column diameter	mm	75	10[a]
Column length	mm	1000	166.7
Mass adsorbent	g		5.9
Time of operation	d	100	4
Water volume	L	28,800	188.4

[a] Assumed value for small column.

Comment The time savings in conducting the RSSCT versus the pilot column is apparent. Furthermore, many more tests can be conducted to test alternative configurations and carbon types.

Analysis and Design of Powdered Activated Carbon Contactor

For a powdered activated carbon (PAC) application, the isotherm adsorption data can be used in conjunction with a materials mass balance analysis to obtain an approximate estimate of the amount of carbon that must be added as illustrated below. Here again, because of the many unknown factors involved, column and bench scale tests are recommended to develop the necessary design data.

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If a mass balance is written around the contactor (i.e., batch reactor) after equilibrium has been reached, the resulting expression is (Note: the following derivation was presented previously in Chap. 4, but is repeated here for continuity):

$$\text{Amount adsorbed} = \begin{matrix}\text{initial amount}\\ \text{of absorbate}\\ \text{present}\end{matrix} - \begin{matrix}\text{final amount}\\ \text{of absorbate}\\ \text{present}\end{matrix} \tag{11-70}$$

$$q_e m = VC_o - VC_e \tag{11-71}$$

where q_e = adsorbent phase concentration after equilibrium, mg adsorbate/g adsorbent
m = mass of adsorbent, g
V = volume of liquid in the reactor, L
C_o = initial concentration of adsorbate, mg/L
C_e = final equilibrium concentration of adsorbate after absorption has occurred, mg/L

It should be noted that q_e is in equilibrium with C. If Eq. (11–71) is solved for q_e, the following expression is obtained.

$$q_e = \frac{V(C_o - C_e)}{m} \tag{11-72}$$

If Eq. (11–72) is rewritten as follows:

$$\frac{V}{m} = \frac{q_e}{C_o - C_e} \tag{11-73}$$

The term V/m is defined as the *specific volume* and represents the volume of liquid that can be treated with a given amount of carbon. The graphical representation of the specific volume was presented previously on Fig. 4–33 in Chap. 4. The reciprocal value of the specific volume corresponds to the dose of adsorbent that must be used.

EXAMPLE 11-12 Estimation of Powdered Activated Carbon (PAC) Adsorption Dose and Cost

A treated wastewater with a flowrate of 1000 L/min is to be treated with PAC to reduce the concentration of residual organics measured as TOC from 5 to 1 mg/L. The Freundlich adsorption isotherm parameters were developed as discussed previously. Assuming the following data apply, determine the PAC requirements to treat the wastewater flow. If PAC costs \$0.50/kg, estimate the annual cost for treatment, assuming the PAC will not be regenerated.

1. Compound = mixed organics
2. Initial concentration, C_o = 5.0 mg/L
3. Final concentration, C_e = 1.0 mg/L
4. GAC density = 450 g/L
5. Freundlich capacity factor, K_f = 150 (mg/g)(L/mg)$^{1/n}$
6. Freundlich intensity parameter, $1/n$ = 0.5

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Solution

1. Estimate the PAC dose based on the isotherm data. The PAC dose can be estimated by writing Eq. (11–71) as follows:

$$\frac{m}{V} = \frac{(C_o - C_e)}{q_e}$$

Substituting Eq. (11–15) for q_e and using the given values in the above expression yields

$$\frac{m}{V} = \frac{[(5 - 1)\ \text{mg/L}]}{(150\ \text{mg/g})(\text{L/mg})^{0.5}(1.0\ \text{mg/L})^{0.5}} = 0.0267\ \text{g/L}$$

2. Estimate the annual cost for the PAC treatment.

Annual cost

$$= \frac{(0.0267\ \text{g/L})(1000\ \text{L/min})(1440\ \text{min/d})(365\ \text{d/yr})(\$0.50/\text{kg})}{(10^3\ \text{g/kg})}$$

Annual cost = \$7008/year

Comment For small wastewater flows, it is not usually cost effective to plan for carbon regeneration.

Activated Sludge with Powdered Activated Carbon Treatment

A proprietary process, "PACT," combines the use of powdered activated carbon with the activated-sludge process (see Fig. 11–60). In this process, when the activated carbon is added directly to the aeration tank, biological oxidation and physical adsorption occur simultaneously. A feature of this process is that it can be integrated into existing activated sludge systems at nominal capital cost. The addition of powdered activated carbon has several process advantages, including (1) system stability during shock loads, (2) reduction of refractory priority pollutants, (3) color and ammonia removal, and (4) improved sludge settleability. In some industrial waste applications, where nitrification is inhibited by toxic organics, the application of powdered activated carbon may reduce or limit this inhibition.

The dosage of powdered activated carbon and the mixed liquor–powdered activated carbon suspended solids concentration are related to the SRT as follows:

$$X_p = \frac{X_i \text{SRT}}{\tau} \tag{11–74}$$

where X_p = equilibrium powdered activated carbon-MLSS content, mg/L
X_i = powdered activated carbon dosage, mg/L
SRT = solids retention time, d
τ = hydraulic retention time, d

Carbon dosages typically range from 20 to 200 mg/L. With higher SRT values, the organic removal per unit of carbon is enhanced, thereby improving the process efficiency.

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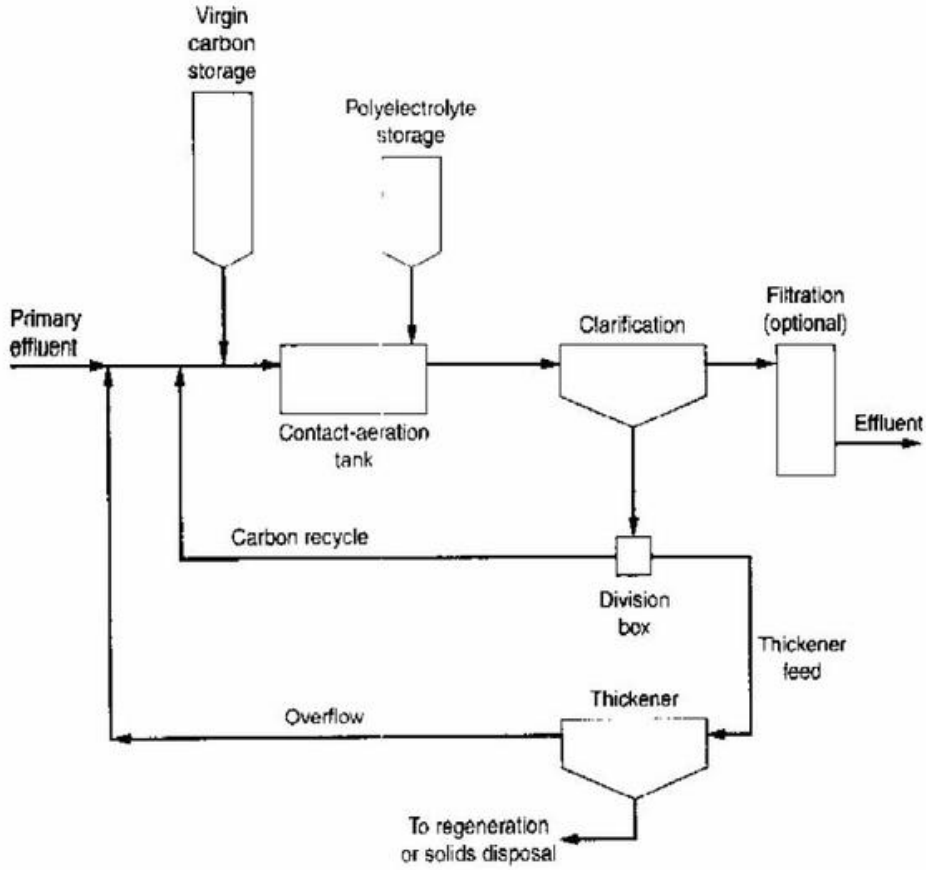
Figure 11–60

Definition sketch for the application of powdered activated carbon.



Reasons cited for this phenomenon include (1) additional biodegradation due to decreased toxicity, (2) degradation of normally nondegradable substances due to increased exposure time to the biomass through adsorption on the carbon, and (3) replacement of low-molecular-weight compounds with high-molecular-weight compounds, resulting in improved adsorption efficiency and lower toxicity.

11–8 GAS STRIPPING

Gas stripping involves the mass transfer of a gas from the liquid phase to the gas phase. The transfer is accomplished by contacting the liquid containing the gas that is to be stripped with a gas (usually air) that does not contain the gas initially. The removal of dissolved gases from wastewaters by gas (usually air) stripping has received considerable attention, especially for the removal of ammonia and odorous gases and volatile organic compounds (VOCs). Early work on the air stripping of ammonia from wastewater was conducted at Lake Tahoe, California (Culp and Slechta, 1966; Slechta and Culp, 1967). The removal of VOCs by aeration was considered earlier in Sec. 5–13.

The purpose of this section is to introduce the fundamental principles involved in gas stripping and to illustrate the general application of these principles. A design procedure is also presented. The material presented in this section is applicable to the removal of ammonia (NH_3), carbon dioxide (CO_2), oxygen (O_2), hydrogen sulfide (H_2S), and a variety of VOCs. The focus of the discussion in this section is on the analysis of facilities designed specifically for the removal of gaseous constituents as opposed

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to the removal of odorous gases (see Sec. 15–3, Chap. 15) and VOCs in aeration systems designed for the biological treatment of wastewater (see Sec. 5–13, Chap. 5).

Analysis of Gas Stripping

Important factors that must be considered in the analysis of gas stripping include (1) the characteristics of the compound(s) to be stripped, (2) the type of contactor to be used and the required number of stages, (3) the materials mass balance analysis of the stripping tower, and (4) the required physical features and dimensions of the required stripping tower.

Characteristics of the Compound(s) to Be Stripped. As noted above, the removal of volatile dissolved compounds by stripping involves contacting the liquid with a gas that does not contain the compound initially. The compound that is to be stripped will come out of solution and enter the gas phase to satisfy the Henry's law equilibrium as discussed in Chap. 2. Compounds such as benzene, toluene, and vinyl chloride that have Henry's law constants greater than 500 atm (mol H_2O/mol air) are readily strippable, compounds such as ammonia 0.75 atm (mol H_2O/mol air) and sulfur dioxide 38 atm (mol H_2O/mol air) are marginally strippable, and compounds such as acetone and methyl ethyl ketone with Henry's law constants less than 0.1 atm (mol H_2O/mol air) are essentially not strippable.

The air stripping of ammonia from wastewater requires that the ammonia be present as a gas. Ammonium ions in wastewater exist in equilibrium with gaseous ammonia, as shown in Eq. (2–38):

$$NH_4^+ \leftrightarrow NH_3 + H^+ \tag{2-38}$$

As the pH of the wastewater is increased above 7, the equilibrium is shifted to the left and the ammonium ion is converted to ammonia, which may be removed by gas stripping. The amount of lime required to raise the pH of wastewater as a function of the alkalinity is given on Fig. 6–11 in Chap. 6.

Methods Used to Contact Phases. In practice, two methods are used to achieve contact between phases so that mass transfer can occur: (1) continuous contact and (2) staged contact. As shown on Fig. 11–61, three flow patterns are used in practice: (1) countercurrent, (2) cocurrent, and (3) cross-flow. In addition, the contact medium may be fixed or mobile (Crittenden, 1999). The most common flow pattern in mass transfer operations is the countercurrent mode. A schematic and photograph of a typical gas stripping tower is shown on Fig. 11–62.

Mass-Balance Analysis for a Continuous Stripping Tower. A steady-state materials balance for the lower portion of a countercurrent continuous stripping tower used for the removal of a dissolved gas from wastewater (see Fig. 11–63) is given by

1. General word statement:

$$\begin{matrix}\text{Moles of solute}\\ \text{entering in}\\ \text{liquid stream}\end{matrix} + \begin{matrix}\text{Moles of solute}\\ \text{entering in}\\ \text{gas stream}\end{matrix} = \begin{matrix}\text{Moles of solute}\\ \text{leaving in}\\ \text{liquid stream}\end{matrix} + \begin{matrix}\text{Moles of solute}\\ \text{leaving in}\\ \text{gas stream}\end{matrix} \tag{11-75}$$

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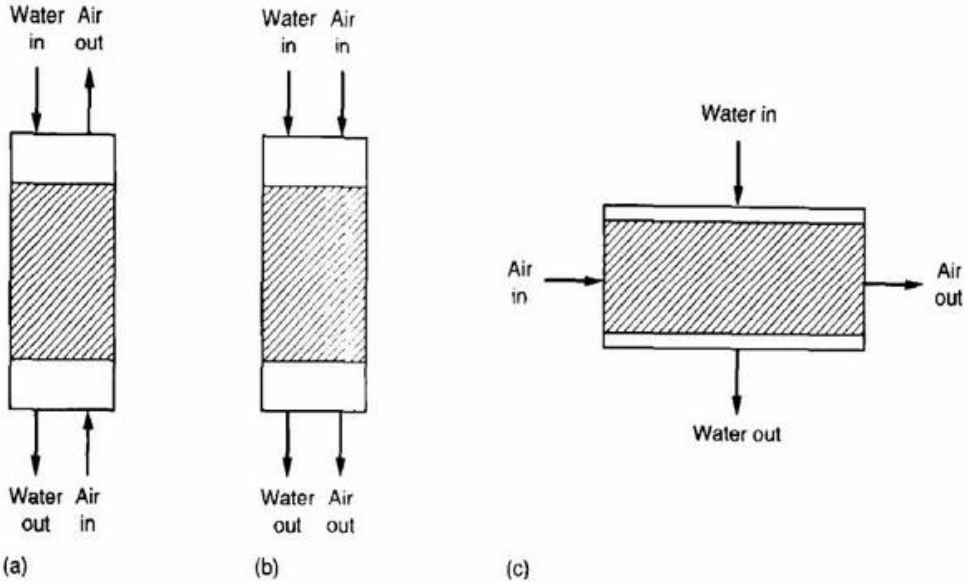
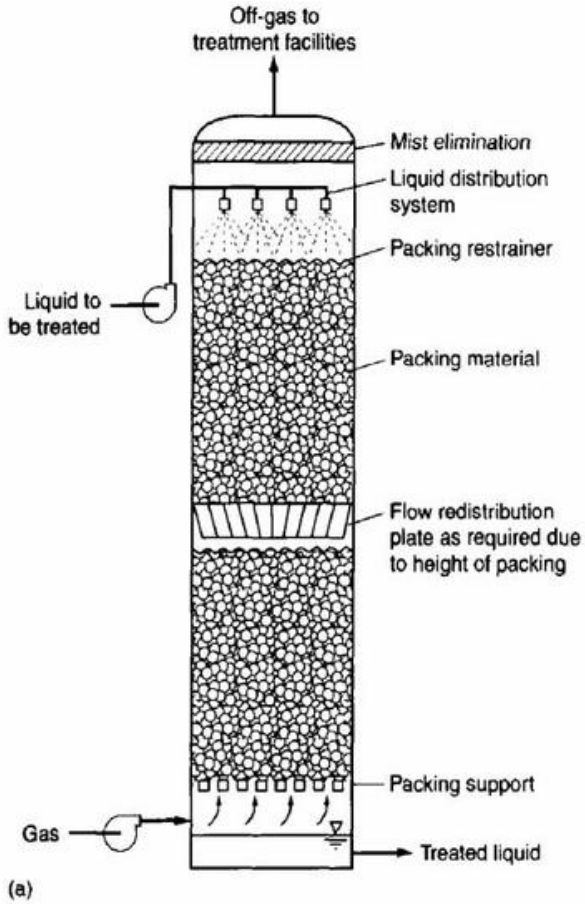
Figure 11-61

Typical water and airflow patterns for gas stripping towers: (a) countercurrent flow, (b) cocurrent flow, and (c) cross flow.





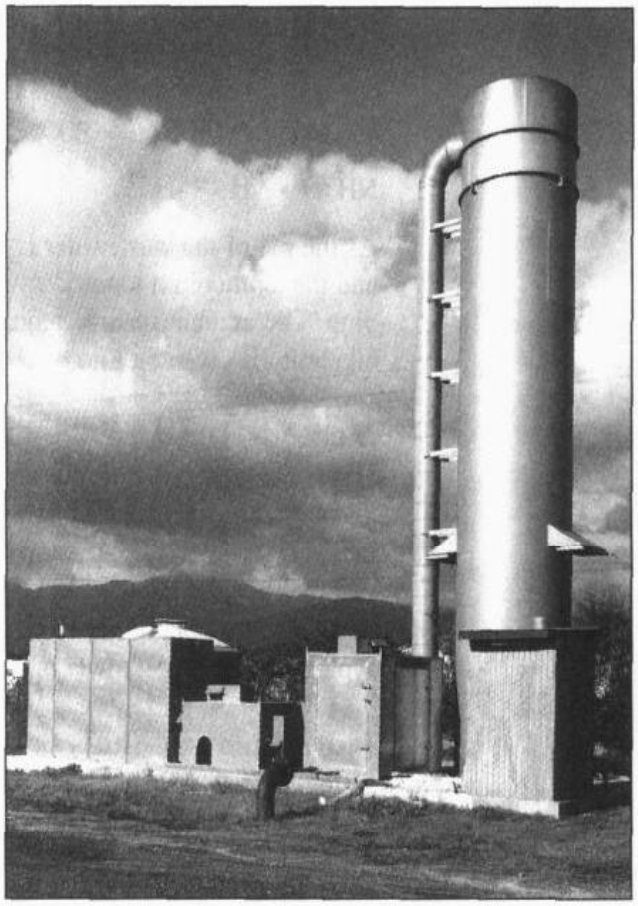


Figure 11-62

Typical stripping towers for the removal of volatile gases from water: (a) schematic and (b) typical operating facility.

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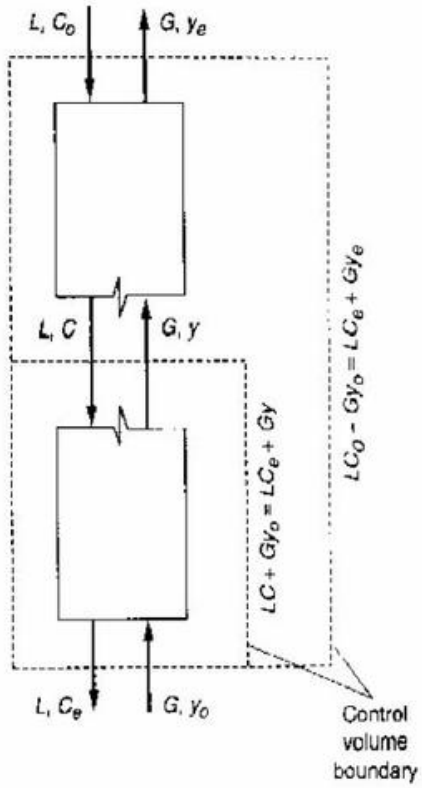
Figure 11-63

Definition sketch for the analysis of a continuous countercurrent flow gas stripping tower.



2. Simplified word statement:

Inflow = outflow (11-76)

3. Symbolic representation (refer to Fig. 11-63):

$$LC + Gy_o = LC_e + Gy \quad (11\text{-}77)$$

where L = moles of incoming liquid (i.e., wastewater) per unit time
C = concentration of solute in liquid at point within the tower, moles of solute per mole of liquid
G = moles of incoming gas per unit time
y_o = concentration of solute in gas entering the bottom of the tower, moles of solute per mole of solute-free gas
C_e = concentration of solute in liquid leaving the bottom of the tower, moles of solute per mole of liquid
y = concentration of solute at a point within the tower, moles of solute per mole of solute-free gas

Combining terms, Eq. (11-77) can be written as

$$(y_o - y) = L/G(C_e - C) \quad (11\text{-}78)$$

If the overall tower is considered, Eq. (11-77) can be written as

$$LC_o + Gy_o = LC_e + Gy_e \quad (11\text{-}79)$$

Combining terms, Eq. (11-79) can be written as

$$(y_o - y_e) = L/G(C_e - C_o) \quad (11\text{-}80)$$

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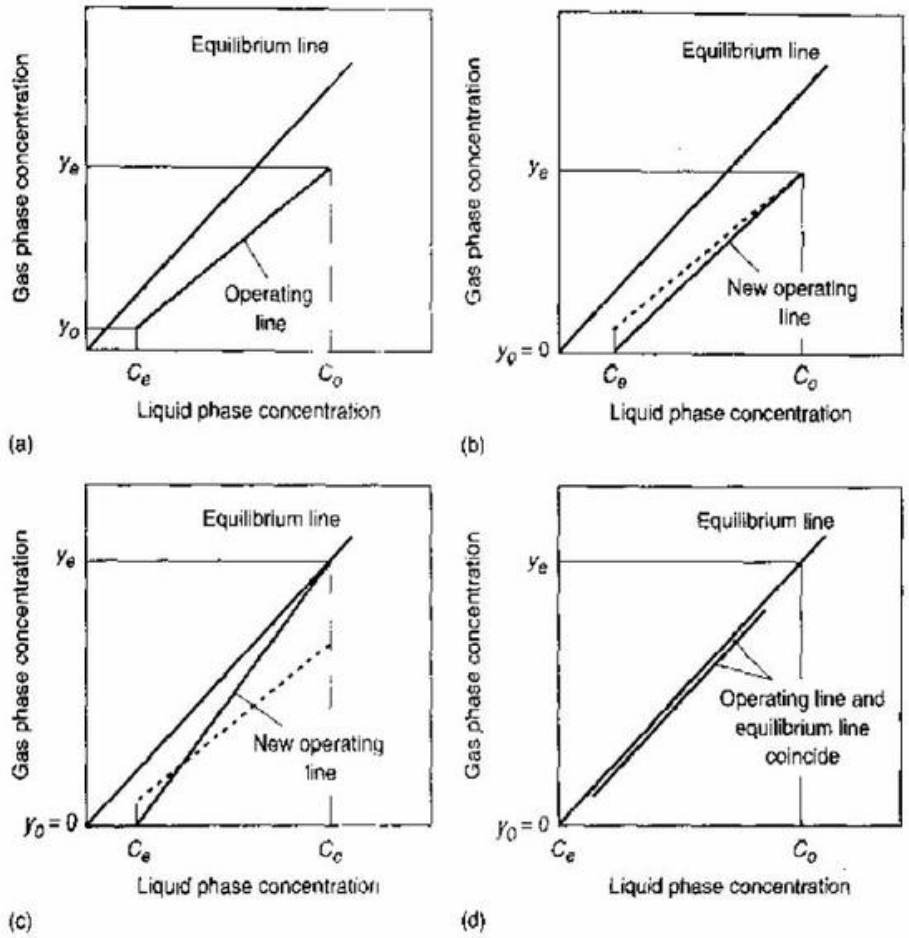
Figure 11–64

Operating lines for various gas stripping conditions: (a) general case, (b) condition when $y_o = 0$, (c) condition when $y_o = 0$ and y_e is in equilibrium with C_o, the constituent concentration in the incoming water, and (d) condition when $y_o = 0$, $C_e = 0$, and y_e is in equilibrium with C_o.



where C_o = concentration of solute in liquid entering at the top of the tower, moles of solute per mole of liquid

y_e = concentration of solute in gas leaving the top of the tower, moles of solute per mole of air

Because Eq. (11–80) is derived solely from a consideration of the equality of input and output, it holds regardless of the internal equilibria that may control the mass transfer. Equation (11–80) represents the equation of a straight line with slope L/G, which passes through the point (C_o, y_o) and point (C_e, y_e). The line passed through these two points (see Fig. 11–64*a*) is known as the *operating line* and represents the conditions at any point within the column. The equilibrium line is based on Henry's law. For example, equilibrium lines defined by Henry's law for ammonia as a function of temperature are presented on Fig. 11–65. It should be noted that when a gas is being stripped from solution, the operating line will lie below the equilibrium line. If a gas is being absorbed into solution the operating line will lie above the equilibrium line.

If it is assumed that the air entering the bottom of the tower contains no solute (i.e, $y_o = 0$), then Eq. (11–80) can be written as

$$y_e = L/G(C_o - C_e) \tag{11–81}$$

The operating line for the condition defined by Eq. (11–81) is shown on Fig. 11–64*b*.

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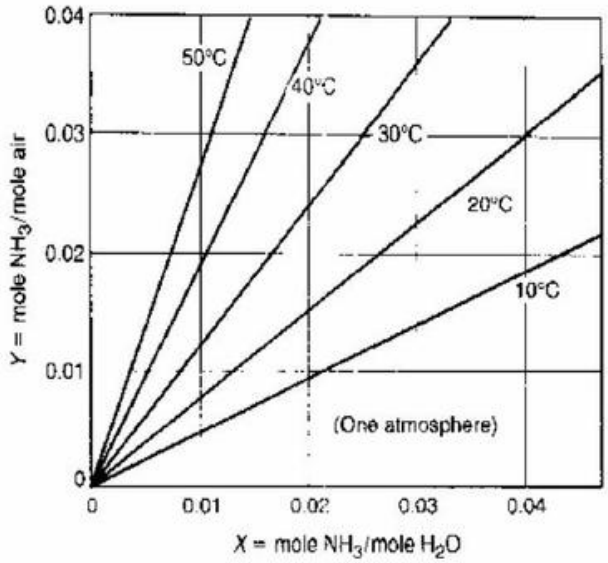
Figure 11-65

Equilibrium curves for ammonia in water as a function of temperature based on Henry's law (see also Table 2-8 in Chap. 2).



Using Henry's law [see Eq. (2-46)], y_e is defined as follows:

$$y_e = \frac{H}{P_T} C_o' \tag{11-82}$$

where y_e = concentration of solute in gas leaving the top of the tower, moles of solute per mole of air

H = Henry's law constant, $\dfrac{\text{atm (mole gas/mole air)}}{\text{(mole gas/mole water)}}$

P_T = total pressure, usually 1.0 atm

C_o' = concentration of solute in liquid that is in equilibrium with the gas leaving the tower, moles of solute per mole of liquid

Using Eq. (11-82), Eq. (11-81) can be written as follows:

$$C_o' = \frac{L}{G} \times \frac{P_T}{H}(C_o - C_e) \tag{11-83}$$

If it is assumed that the concentration of solute in the liquid entering the tower is in equilibrium with the gas leaving the tower, Eq. (11-83) can be written as

$$\frac{G}{L} = \frac{P_T}{H} \times \frac{C_o - C_e}{C_o} \tag{11-84}$$

The operating line for the condition defined by Eq. (11-84) is shown on Fig. 11-64*c*. The value of G/L (air to liquid ratio) defined by Eq. (11-84) represents the minimum amount of air that can be used for stripping for the given conditions (i.e, $y_o = 0$ and $y_e = HC_o/P_T$). In practice, from one and a half to three times the theoretical minimum air-to-liquid ratio is used to achieve effective stripping of most constituents. The application of this relationship is illustrated in Example 11-13.

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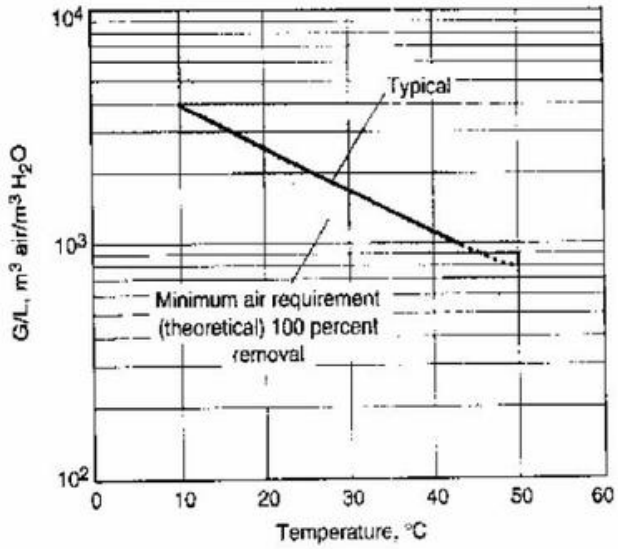
Figure 11-66

Air requirements for ammonia stripping as a function of temperature.



If it is assumed further that the liquid leaving and the air entering the bottom of the tower contains no solute, then Eq. (11–84) can be written as

$$\frac{G}{L} = \frac{P_T \times C_o}{H \times C_o} = \frac{P_T}{H} \tag{11-85}$$

The value of G/L for this condition corresponds to the equilibrium line defined by Henry's law (see Fig. 11–64*d*), and represents the theoretical minimum amount of air that can be used for stripping for the given conditions (i.e., $y_o = 0$, $C_e = 0$, and $y_e = HC_o/P_T$). The theoretical air-to-liquid ratio for stripping ammonia from wastewater for various temperatures is plotted on Fig. 11–66. The theoretical ratio is derived by assuming the process to be 100 percent efficient with a stripping tower of infinite height—obviously unachievable in practice.

EXAMPLE 11–13 **Air Requirements for Ammonia Stripping** Determine the theoretical amount of air required at 20°C to reduce the ammonia concentration from 40 to 1 mg/L in a treated wastewater with a flowrate of 4000 m³/d. Assume the pH of the wastewater has been increased to a value of 11 (see Fig. 2–13 in Chap. 2), the Henry's constant for ammonia at 20°C is 0.75 atm, and the air entering the bottom of the tower does not contain any ammonia.

Solution

1. Determine the influent and effluent mole fractions of ammonia in the liquid using Eq. (2–3).

$$x_B = \frac{n_B}{n_A + n_B}$$

where x_B = mole fraction of solute B
n_B = number of moles of solute B
n_A = number of moles of solute A

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$$C_o = \frac{[(40 \times 10^{-3})/17]}{[55.5 + (40 \times 10^{-3})/17]} = 4.24 \times 10^{-5} \text{ mole } NH_3/\text{mole } H_2O$$

$$C_e = \frac{[(1 \times 10^{-3})/17]}{[55.5 + (1 \times 10^{-3})/17]} = 1.06 \times 10^{-6} \text{ mole } NH_3/\text{mole } H_2O$$

2. Determine the effluent mole fraction of ammonia in the air leaving the tower using Eq. (11–82).

$$y_e = \frac{H}{P_T} C_o$$

$$H = \frac{(0.75 \text{ atm})(\text{mole } NH_3/\text{mole air})}{(\text{mole } NH_3/\text{mole } H_2O)} = (0.75 \text{ atm})\left(\frac{\text{mole } H_2O}{\text{mole air}}\right)$$

$$y_e = \frac{H}{P_T} \times C_o = \frac{0.75 \text{ atm}}{1.0 \text{ atm}}\left(\frac{\text{mole } H_2O}{\text{mole air}}\right) \times (4.24 \times 10^{-5} \text{ mole } NH_3/\text{mole } H_2O)$$

$$= 3.18 \times 10^{-5} \frac{\text{mole } NH_3}{\text{mole air}}$$

3. Determine the gas-to-liquid ratio using Eq. (11–84) rearranged as follows:

$$\frac{G}{L} = \frac{P_T}{H} \times \frac{(C_o - C_e)}{C_o} = \frac{(C_o - C_e)}{y_e}$$

$$\frac{G}{L} = \frac{[(4.24 \times 10^{-5} - 0.106 \times 10^{-5}) \text{ mole } NH_3/\text{mole } H_2O]}{(3.18 \times 10^{-5} \text{ mole } NH_3/\text{mole air})} = 1.3 \frac{\text{mole air}}{\text{mole } H_2O}$$

4. Convert the moles of air and water to liters of air and water.

For air at 20°C:

1.3 mole × 24.1 L/mole = 31.33 L

For water:

$$\frac{(1.0 \text{ mole } H_2O)(18 \text{ g/mole } H_2O)}{(10^3 \text{ g/L})} = 0.018 \text{ L}$$

$$\frac{G}{L} = \frac{31.33 \text{ L}}{0.018 \text{ L}} = 1741 \text{ L/L} = 1741 \text{ m}^3/\text{m}^3$$

5. Determine the total quantity of air required based on ideal conditions.

$$\text{Air required} = \frac{(1741 \text{ m}^3/\text{m}^3)(4000 \text{ m}^3/\text{d})}{(1440 \text{ min/d})} = 4835 \text{ m}^3/\text{min}$$

Comment The approximate amount of lime required to raise the pH of wastewater to a value of 11 is given on Fig. 6–11 in Chap. 6. The procedure followed to determine the height of the stripping tower is illustrated in Example 11–14, presented later in this section.

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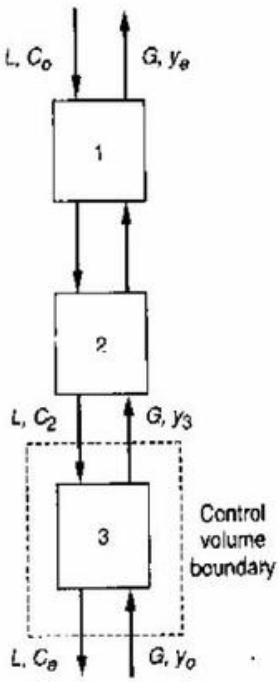


Figure 11–67

Definition sketch for the analysis of a three-stage countercurrent flow gas stripping tower.

Mass-Balance Analysis of a Multistage Stripping Tower. In the analysis of stripping towers, reference is often made to the number of ideal stages required for stripping. The analysis for the number of stages is analogous to the simulation of plug flow with a series of complete-mix reactors as detailed in Chap. 4. Separate stages are used to improve performance of stripping towers. Equilibrium conditions are assumed in each stage of the tower. A steady-state materials balance for the lower portion of a countercurrent staged stripping tower (see Fig. 11–67) is given by

$$\text{Inflow} = \text{outflow}$$

$$LC_2 + Gy_o = LC_e + Gy_3 \tag{11–86}$$

or

$$(y_3 - y_o) = L/G(C_2 - C_e) \tag{11–87}$$

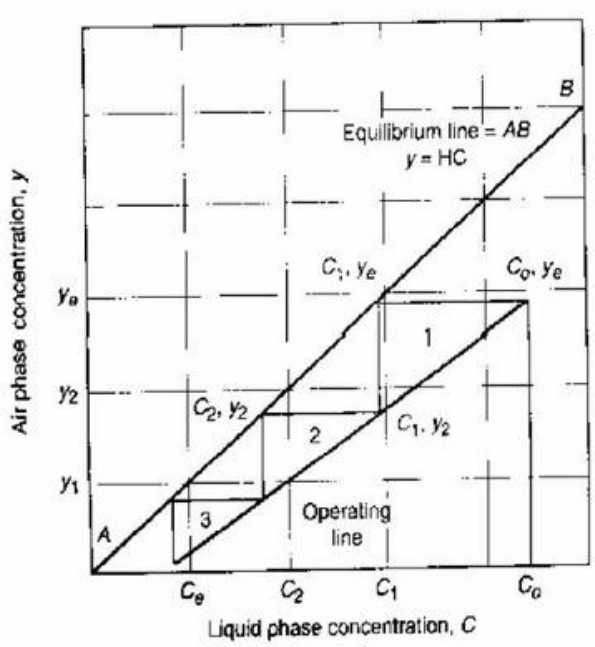
If an overall mass balance is performed around the tower, the resulting equation is the same as that derived above for the continuous stripping tower [see Eq. (11–79)].

In the 1920s, McCabe and Thiele (1925) developed a graphical procedure for determining the required number of ideal stages. The method is illustrated on Fig. 11–68 for a stripping tower comprised of three stages. The operating line for the three stages is also shown on Fig. 11–68. The number of ideal stages required for stripping of a constituent is obtained as follows. The point C_o, y_e represents the airstream leaving and the water stream entering the top of the stripping column. The composition of the liquid in equilibrium with the constituent concentration in the airstream is found by extending a horizontal line from the point C_o, y_e to the point C_1, y_e. From the point C_1, y_e the value of y_2, the air entering stage 1 from stage 2, is obtained from the equation of the operating line. If a materials mass balance is performed between stages 1 and 2, the resulting expression for y_2 is

$$y_2 = \frac{L}{G} C_1 + \frac{Gy_e - LC_o}{G} \tag{11–88}$$

Figure 11–68

Operating line for three-stage countercurrent flow gas stripping tower.



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The value of y_2 is obtained by drawing a vertical line from point C_1, y_e to the operating line at point C_1, y_2 as shown on Fig. 11–68. In a similar manner the value of C_2 is obtained by drawing a horizontal line from the point C_1, y_2 to the equilibrium line. This procedure is repeated until the point C_n, y_{n-1} is reached. The number of ideal stages is typically a fractional number (e.g., 4.2, 5.6). In practice, the number of stages is rounded to the next whole number.

Determination of Height of Stripping Tower. The purpose of the following analysis is to illustrate how the height of the packing in the stripping tower is determined, based on an analysis of the mass transfer occurring within the tower. A mass balance performed on the liquid phase within the stripping tower shown on Fig. 11–69 is as follows.

Simplified word statement:

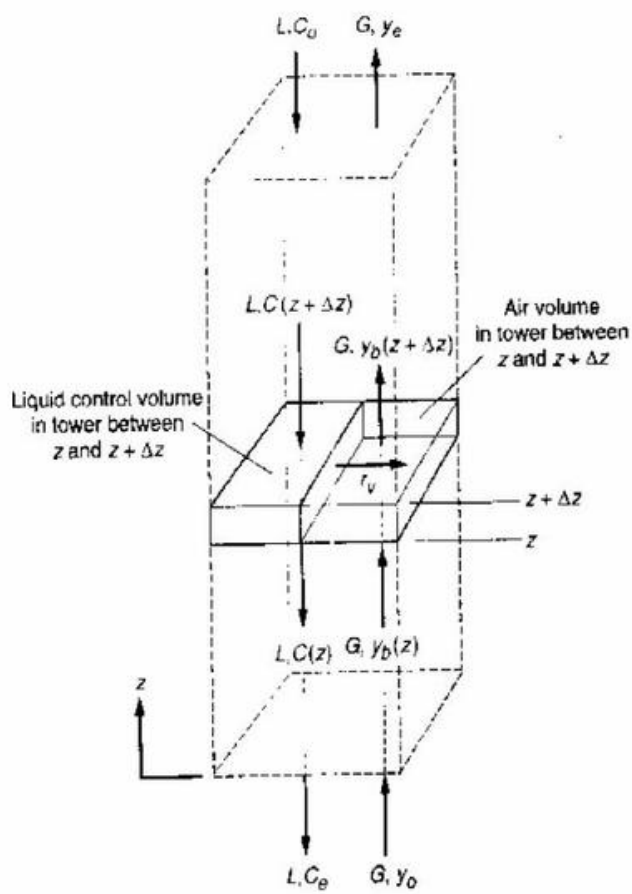
Accumulation = inflow − outflow + volatilization

$$\frac{\partial C}{\partial t}\Delta V = LC|_{z-\Delta z} - LC|_z + r_v \Delta V \tag{11–89}$$

where $\partial C/\partial t$ = change in concentration with time, g/m^3·s
ΔV = differential volume, m^3
Δz = differential height, m
L = liquid volumetric flowrate, m^3/s
C = concentration of constituent C, g/m^3
r_v = rate of mass transfer per unit volume per unit time C, g/m^3·s

Figure 11–69

Definition sketch for the analysis of mass transfer from the liquid phase to the gas phase within a stripping tower. *Note:* the packing material is not shown. *(Adapted from Hand et al., 1999.)*



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Substituting the area times the differential height ($A\Delta z$) for the differential volume (ΔV) and writing the differential form for the term $LC|_{z+\Delta z}$ in Eq. (11–89) results in the following expression:

$$\frac{\partial C}{\partial t} A\Delta z = L\left(C + \frac{\Delta C}{\Delta z}\Delta z\right) - LC + r_v A\Delta z \tag{11–90}$$

Simplifying Eq. (11–90) and taking the limit as Δz approaches zero yields

$$\frac{\partial C}{\partial t} = \frac{L}{A}\frac{\partial C}{\partial z} + r_v \tag{11–91}$$

The rate of mass transfer due to volatilization as described in Chap. 4 [see Eq. (4–127)] is given by:

$$r_v = -K_L a(C_b - C_s) \tag{11–92}$$

where r_v = rate of mass transfer per unit volume per unit time, g/m^3·s
$K_L a$ = volumetric mass transfer coefficient which depends on water-quality characteristics and temperature, 1/s
C_b = concentration in liquid bulk phase at time t, g/m^3
C_s = concentration in liquid in equilibrium with gas as given by Henry's law, g/m^3

Assuming steady-state conditions within the tower ($\partial C/\partial t = 0$) and substituting Eq. (11–92) for r_v, Eq. (11–91) can now be written as

$$\frac{dC_b}{dz} = \frac{K_L aA}{L}(C_b - C_s) \tag{11–93}$$

The height of the tower packing can be obtained by integrating the above expression:

$$\int_0^Z dz = \frac{L}{K_L aA}\int_{C_o}^{C_e}\frac{dC_b}{(C_b - C_s)} \tag{11–94}$$

To integrate the right-hand side of the above equation, a relationship must be found between C_b and C_s because C_s is changing continuously throughout the height of the tower. From Henry's law, the value of C_s is given by:

$$C_s = \frac{P_T}{H}y \tag{11–95}$$

Substituting a modified form of Eq. (11–81) for y in Eq. (11–95) yields:

$$C_s = \frac{L}{G} \times \frac{P_T}{H}(C_b - C_e) \tag{11–96}$$

If Eq. (11–96) is substituted into Eq. (11–94) and the resulting expression is integrated, the following expression is obtained (Hand et al., 1999):

$$Z = \frac{L}{K_L aA}\left[\frac{C_o - C_e}{C_o - C_e - C_o'}\right]\ln\left[\frac{C_o - C_o'}{C_e}\right] \tag{11–97}$$

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where

$$C_o' = \frac{L}{G} \times \frac{P_T}{H}(C_o - C_e) \tag{11-98}$$

It should be noted that if $C_o' = C_o$, Eq. (11-98) is the same as Eq. (11-84).

Design Equations for Stripping Towers. Utilizing the above equations, a number of process models and design equations have been developed. Equations that can be used to determine the height of a stripping tower are as follows:

$$Z = \text{HTU} \times \text{NTU} \tag{11-99}$$

where Z = height of stripping tower packing, m
HTU = height of a transfer unit, m
NTU = number of transfer units

The height of a transfer unit is defined as

$$\text{HTU} = \frac{L}{K_L a A} \tag{11-100}$$

where L = liquid volumetric flowrate, m^3/s
$K_L a$ = volumetric mass transfer coefficient, 1/s
A = cross-sectional area of tower, m^2

The HTU is a measure of the mass transfer characteristics of the packing medium.

The number of transfer units is defined as

$$\text{NTU} = \left[\frac{C_o - C_e}{C_o - C_e - C_o'}\right] \ln\left[\frac{C_o - C_o'}{C_e}\right] \tag{11-101}$$

Substituting Eq. (11-98) in Eq. (11-101) yields

$$\text{NTU} = \left[\frac{S}{S-1}\right] \ln\left[\frac{(C_o/C_e)(S-1)+1}{S}\right] \tag{11-102}$$

where S is known as the stripping factor and is defined as

$$S = \frac{G}{L} \times \frac{H}{P_T} \tag{11-103}$$

A value of $S = 1$ corresponds to the minimum amount of air required for stripping. When $S > 1$ the amount of air is in excess and complete stripping is possible given a tower of infinite height. When $S < 1$, there is insufficient air for stripping. In practice, stripping factors vary from 1.5 to 5.0.

Values for $K_L a$ for specific compounds are best obtained from pilot-plant studies or by using empirical correlations such as given in Eq. (5-62) in Chap. 5, and repeated here for convenience as Eq. (11-104). It should be noted that many other relationships have been proposed in the literature (Sherwood and Hollaway, 1940, and Onda et al., 1968).

$$K_L a_{\text{VOC}} = K_L a_{O_2}\left(\frac{D_{\text{VOC}}}{D_{O_2}}\right)^n \tag{11-104}$$

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where $K_L a_{VOC}$ = system mass transfer coefficient, 1/h
$K_L a_{O_2}$ = system oxygen mass transfer coefficient, 1/h
D_{VOC} = diffusion coefficient of VOC in water, cm^2/s
D_{O_2} = diffusion coefficient of oxygen in water, cm^2/s
n = coefficient (0.5 for stripping towers)

Air and water temperature are significant factors in the design of stripping towers because of their effect on air and water viscosities, Henry's law constants, and volumetric mass transfer coefficients. The effect of temperature on the Henry's law constant is illustrated on Fig. 11–65. The value of $K_L a$ can be adjusted for temperature effects using Eq. (2–25) with a theta value of 1.024.

Design of Stripping Towers

In its simplest form a stripping tower consists of a tower (usually circular), a support plate for the packing material, a distribution system for the liquid to be stripped, located above the packing material, and an air supply located at the bottom of the stripping tower (see Fig. 11–62). The process design variables include (1) the type of packing material, (2) the stripping factor, (3) the cross-sectional area of the tower, and (4) the height of the stripping tower. The cross-sectional area will depend on the pressure drop through the packing. Representative design values for stripping of VOCs and ammonia are presented in Table 11–35. The significant difference in the amount of air required for stripping is a clear illustration of the importance of the Henry's constant.

The headloss through the packing is determined using a generalized gas pressure-drop relationship such as that plotted on Fig. 11–70 (Eckert, 1975). The pressure drop is expressed in newtons per square meter per meter of depth $(N/m^2)/m$. The upper line on Fig. 11–70 labeled *approximate flooding* represents the condition that occurs when the amount of water and air applied is so great that the pore spaces fill to the point where water starts to flood within the tower. The units for the x and y axes are as follows:

x axis:

$$x = \frac{L'}{G'}\left(\frac{\rho_G}{\rho_L - \rho_G}\right)^{1/2} \approx \frac{L'}{G'}\left(\frac{\rho_G}{\rho_L}\right)^{1/2} \tag{11–105}$$

y axis:

$$y = \frac{(G')^2(C_f)(\mu_L)^{0.1}}{(\rho_G)(\rho_L - \rho_G)} \tag{11–106}$$

which can be rewritten as follows:

$$G' = \left[\frac{(\text{value from } y \text{ axis})(\rho_G)(\rho_L - \rho_G)}{(C_f)(\mu_L)^{0.1}}\right]^{1/2} \tag{11–107}$$

where L' = liquid loading rate, $kg/m^2 \cdot s$
G' = gas loading rate, $kg/m^2 \cdot s$
ρ_G = density of gas, kg/m^3
ρ_L = density of liquid, kg/m^3
C_f = packing factor for packing material, unitless
μ_L = viscosity of liquid, $N/m^2 \cdot s$

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Table 11-35
Typical design parameters for stripping towers for the removal of VOC and ammonia[a]

Item	Symbol	Unit	VOC removal[b]	Ammonia removal[c]
Liquid loading rate		$L/m^2 \cdot min$	600–1800	40–80
Air-to-liquid ratio[d]	G/L	m^3/m^3	20–60:1	2000–6000:1
Stripping factor	S	unitless	1.5–5.0	1.5–5.0
Allowable air pressure drop	ΔP	$(N/m^2)/m$	100–400	100–400
Height-to-diameter ratio	Z/D	m/m	≤10:1	≤10:1
Packing depth[e]	Z	m	1–6	2–6
Factor of safety	SF	%D, %Z	20–50	20–50
Wastewater pH	pH	Unitless	5.5–8.5	10.8–11.5
Approximate packing factors				
Pall rings, Intalox saddles	C_f	12.5 mm[f]	180–240	180–240
	C_f	25 mm[f]	30–60	30–60
	C_f	50 mm[f]	20–25	20–25
Berl saddles, Raschig rings	C_f	12.5 mm[f]	300–600	300–600
	C_f	25 mm[f]	120–160	120–160
	C_f	50 mm[f]	45–60	45–60

[a] Adapted in part from Eckert (1970, 1975), Kavanaugh and Trussell (1980), and Hand et al. (1999).
[b] Typical data for VOCs with Henry's constants greater than 500 atm (mole H_2O/mole air).
[c] Ammonia with a Henry's constant of 0.75 atm (mole H_2O/mole air) is considered only marginally strippable, which accounts for the low loading rate and high air-to-liquid ratio.
[d] Ratio is highly temperature-dependent.
[e] For packing depths greater than 5 to 6 m, redistribution of the liquid flow is recommended (see Fig. 11-62a).
[f] Size of packing material.

The packing factor C_f depends on the type and size of the packing. Typical ranges for packing factors that can be used for preliminary assessments are reported in Table 11-35. For more detailed design calculations, current values should be obtained from manufacturers.

To use Fig. 11-70, select a value for L'/G' and compute the corresponding x value. Enter the plot at the computed value of x and move vertically upward to a preselected pressure-drop line. Move horizontally from the point of intersection to the y axis and note the value on the y axis. Using the y-axis value, determine the gas loading rate G' using Eq. (11-107) and the corresponding liquid loading rate L'. To determine the required cross-sectional area the liquid flowrate is divided by the liquid loading rate.

A generalized analysis procedure is as follows:

1. Select a packing material and its corresponding packing factor for use in Eq. (11-105).
2. Select several stripping factors for successive trials (e.g., 2.5, 3.0, 4.0)

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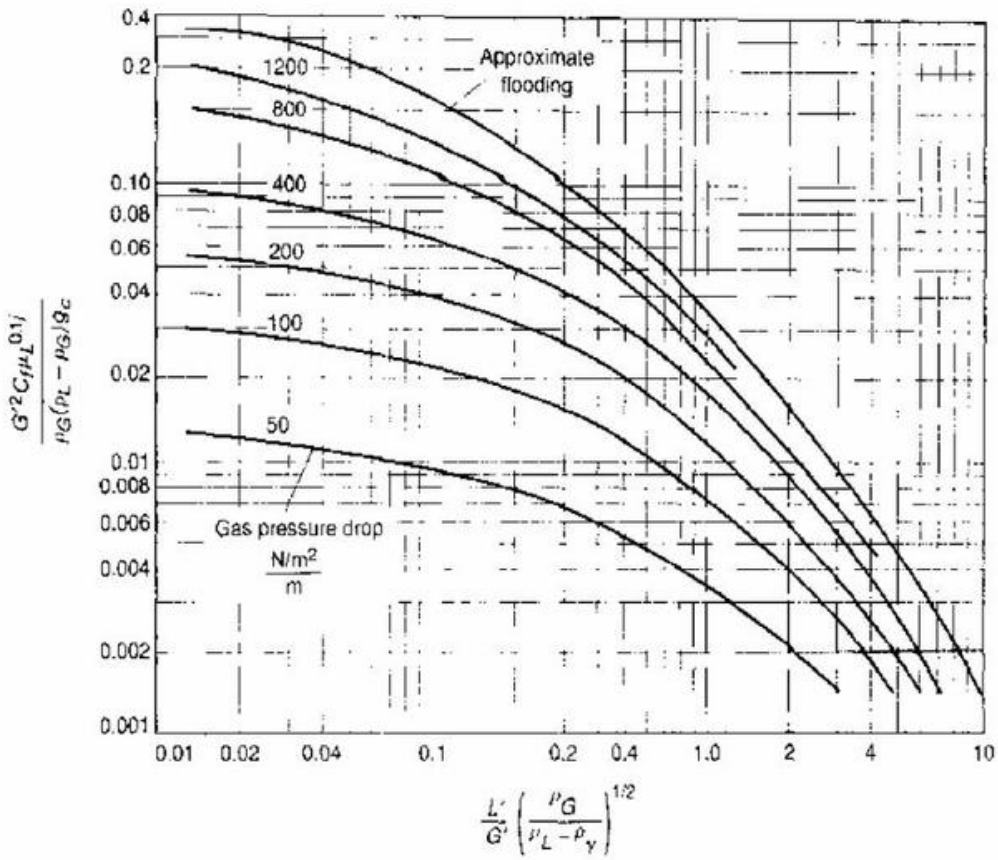
Figure 11-70

Generalized pressure-drop curves for packed-bed stripping towers. (From Eckert, 1975.)



3. Select acceptable pressure drop ΔP (typically a function of the packing material selected).
4. Determine the cross-sectional area of the tower, based on the allowable pressure drop using the data presented on Fig. 11–70 or other appropriate relationships.
5. Determine the height of the transfer units using Eq. (11–100). To apply Eq. (11–100), the value of $K_l a$ must be known or estimated using Eq. (11–104).
6. Determine the number of transfer units using Eq. (11–102).
7. Determine the height of the stripping tower using Eq. (11–99).

The design procedure outlined above is illustrated in Example 11–14. Representative design values for stripping towers are given in Table 11–35. To evaluate the stripping process more thoroughly, any of the commercially available software packages can be used.

EXAMPLE 11–14 Determination of Height of Stripping Tower for the Removal of Ammonia

Determine the diameter and height of the stripping tower required to treat the wastewater in Example 11–13 using the data given in Table 11–35. Assume 25 mm Pall rings will be used as the packing material in the stripping tower. The ammonia concentration in a treated wastewater from a flow of 4000 m³/d is to be reduced from 40 to 1 mg/L. Assume that the Henry's constant for ammonia at 20°C is 0.75 atm, and the air entering the bottom of the tower does not contain any ammonia. Assume the $K_L a$ value for ammonia is 0.0125 s⁻¹.

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Solution

1. Select a packing material and corresponding packing factor. For 25 mm Pall rings (specified), assume a packing factor of 50 (see Table 11–35).
2. Select a stripping factor. Assume a stripping factor of 3 (see Table 11–35).
3. Select an acceptable pressure drop. Assume a pressure drop of 200 (N/m^2)/m (see Table 11–35).
4. Determine the cross-sectional area of the tower stripping using the pressure drop plot given on Fig. 11–71.
   a. Determine the value of the ordinate for a stripping factor of 3.

$$S = \frac{G}{L} \times \frac{H}{P_T} = \frac{\text{G mole air}}{\text{L mole water}} \times \frac{0.75 \text{ atm}}{1.0 \text{ atm}} = \frac{\text{G mole air}}{\text{L mole water}} \times 0.75$$

$$S = 0.75 \times \left(\frac{\text{G mole air}}{\text{L mole water}}\right)\left(\frac{28.8 \text{ g}}{\text{mole air}}\right)\left(\frac{\text{mole water}}{18 \text{ g}}\right)$$

$$= 1.2 \frac{G \text{ g}}{L \text{ g}} = 1.2 \frac{G' \text{ kg}}{L' \text{ kg}}$$

$$\frac{L'}{G'} = \frac{3}{1.2 \text{ kg/kg}} = 2.5$$

$$\frac{L'}{G'}\left(\frac{\rho_G}{\rho_L - \rho_G}\right)^{1/2} \approx \frac{L'}{G'}\left(\frac{\rho_G}{\rho_L}\right)^{1/2} = (2.5 \text{ kg/kg})\left[\frac{(1.204 \text{ kg/m}^3)}{(998.2 \text{ kg/m}^3)}\right]^{1/2} = 0.087$$

   b. Determine the corresponding value on the abscissa. For an ordinate value of 0.087 and a pressure drop of 200 (N/m^2)/m, the ordinate value is 0.04.
   c. Using an ordinate value 0.04, determine the loading rate using Eq. (11–107).

$$G' = \left[\frac{(\text{value from } y \text{ axis})(\rho_G)(\rho_L - \rho_G)}{(C_f)(\mu_L)^{0.1}}\right]^{1/2}$$

$$G' = \left[\frac{(0.04)(1.204)(998.2 - 1.204)}{(50)(0.001)^{0.1}}\right]^{1/2} = 1.38 \text{ kg/m}^2\text{·s}$$

$$L' = 2.5\, G' = 2.5 \times 1.38 \text{ kg/m}^2\text{·s} = 3.45 \text{ kg/m}^2\text{·s}$$

   d. Substitute known values and solve for the diameter of the tower.

$$D = \left[\frac{(4/\pi)(4000 \text{ m}^3/\text{d})(998.2 \text{ kg/m}^3)(1 \text{ d}/86{,}400 \text{ s})}{(4.8 \text{ kg/m}^2\text{·s})}\right]^{1/2} = 4.13 \text{ m}$$

5. Determine the height of the transfer unit using Eq. (11–100).

$$\text{HTU} = \frac{L}{K_L a A}$$

$$\text{HTU} = \frac{L}{K_L a A} = \frac{(4000 \text{ m}^3/\text{d})(1 \text{ d}/86{,}400 \text{ s})}{(0.0125/\text{s})[(3.14/4)(4.13 \text{ m})^2]} = 0.28 \text{ m}$$

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6. Determine the number of transfer units using Eq. (11–102).

$$\text{NTU} = \left[\frac{S}{S-1}\right] \ln \left[\frac{(C_o/C_e)(S-1)+1}{S}\right]$$

$$\text{NTU} = \left[\frac{3}{3-1}\right] \ln \left[\frac{(40/1)(3-1)+1}{3}\right] = 4.94$$

7. Determine the height of the stripping tower packing using Eq. (11–99).

$$Z = \text{HTU} \times \text{NTU} = 0.28 \times 4.94 = 1.38 \text{ m}$$

Comment In this example, the value of K_La for ammonia was known. Quite often the required K_La value must be determined in the field, using pilot-scale facilities. Alternatively Eq. (11–104) can be used to estimate a value for K_La. In some cases, data from the literature or from manufacturers may be used to obtain preliminary sizing. To optimize the design, various stripping ratios must be evaluated. Optimization is best accomplished using one of the commercially available software packages.

Application

As noted previously, air stripping is used to remove ammonia (NH_3), carbon dioxide (CO_2), oxygen (O_2), hydrogen sulfide (H_2S), and a variety of VOCs. The application of air stripping for the removal of ammonia is considered in the following discussion. The removal and treatment of VOCs in aeration systems is considered in Sec. 5–13 in Chap. 5. The VOC management alternatives considered in Chap. 5 are also applicable for the treatment of VOCs removed with a stripping tower. The removal and treatment of odorous compounds are considered in Sec. 15–3 in Chap. 15.

Removal of Ammonia from Wastewater. In general, removal efficiency depends on the temperature, size, and proportions of the facility, and the efficiency of the air-water contact. If the removal of ammonia is unsatisfactory, the tower has not been designed correctly or is overloaded. In this case, additional air volume may improve operation. As the temperature decreases, the amount of air required increases significantly for the same degree of removal (see Fig. 11–66). A typical flow diagram for the removal of ammonia from wastewater by air stripping is shown on Fig. 11–71. The ammonia stripped from the wastewater is converted to the ammonium ion using an acid scrubber.

In most cases where ammonia stripping has been applied, a number of operating problems have developed, the most serious being (1) maintaining the required pH for effective stripping, (2) calcium carbonate scaling within the tower and feed lines, and (3) poor performance during cold weather operation. Maintaining the required pH is a control problem that can be managed with multiple sensors. The amount and nature

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(soft to extremely hard) of the calcium carbonate scale formed varies with the characteristics of the wastewater and local environmental conditions, and cannot be predicted a priori. Under conditions of icing, the liquid-air contact geometry in the tower is altered, which further reduces the overall efficiency. The best solution for cold weather conditions is to enclose the stripper.

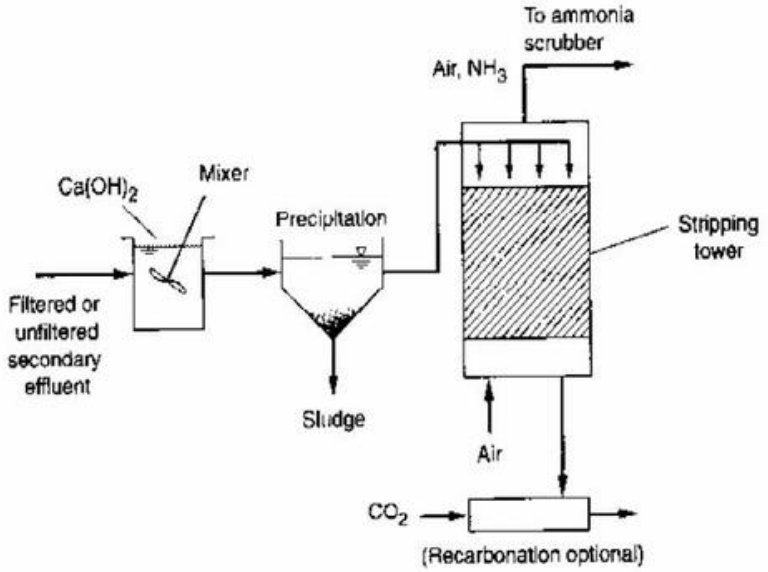
Removal of Ammonia from Digester Supernatant by Steam Stripping. Steam stripping of ammonia is similar to air stripping, with the exception that the process requires temperatures in excess of 95°C (200°F). Although costly, steam stripping has been proposed for the removal of ammonia from digester supernatant in large treatment plants, where separate facilities are used for the treatment of return flows (see discussion in Chap. 14). The flow diagram for the removal of ammonia from dewatering centrate is shown on Fig. 11-72. The ammonia stripped from the supernatant is converted to the ammonium ion by passing the off-gas through an acid bath. Acid scrubbers are also used.

In a large-scale test, two processes were evaluated for the removal of ammonia from centrate: steam stripping and hot air stripping. Steam stripping proved to be the most cost-effective. Based on the results of recent full-scale testing, it has been found that pretreatment of the centrate is essential for effective steam stripping of ammonia. Pretreatment consisting of settling followed by straining proved to be effective. Using a gas-to-liquid ratio of 300 to 1, it was possible to achieve ammonia removals of about 88 percent, at pH values of 7 to 7.5. The influent ammonia concentrations values were on the order of 500 to 600 mg/L (Gopalakrishnan et al., 2000).

Operating problems that have been encountered with steam stripping include (1) extensive fouling (iron deposits, for example) within the heat exchanger and in the stripper due to the presence of waste constituents at elevated temperatures, (2) maintaining the required pH for effective stripping, (3) controlling the steam flow, and

Figure 11-71

Typical flow diagram for the air stripping of ammonia from wastewater.



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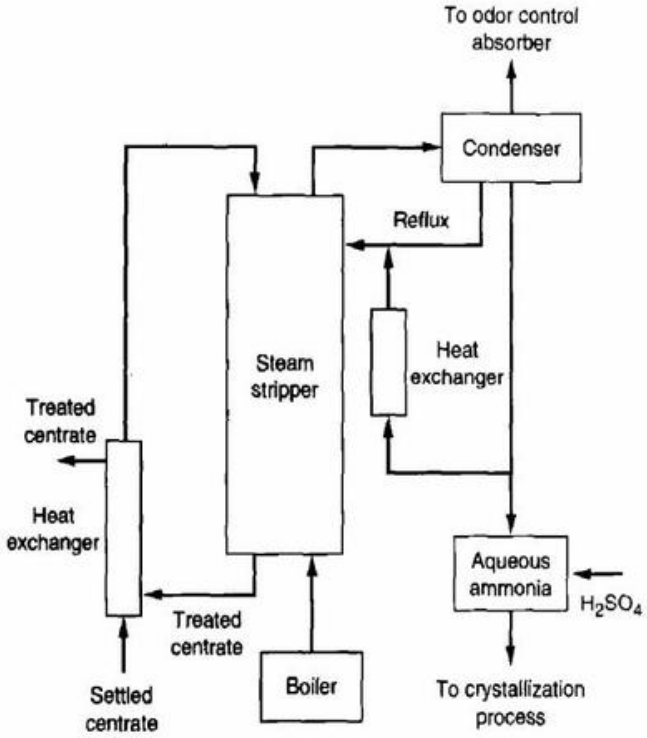

(a)

(b)

Figure 11-72

Stream stripping of ammonia from centrate from dewatering operations: (a) typical flow diagram, and (b) view of pilot-plant facility. (Courtesy of The New York City Department of Environmental Protection.)

(4) maintaining the stripping tower temperature. Because of the importance of temperature, steam stripping should be carried out in enclosed facilities. Spiral-wound heat exchangers have proved to be effective. In Europe, an acid wash is used to clean the piping, the heat exchanger, and the stripping column.

11-9 ION EXCHANGE

Ion exchange is a unit process in which ions of a given species are displaced from an insoluble exchange material by ions of a different species in solution. The most widespread use of this process is in domestic water softening, where sodium ions from a cationic-exchange resin replace the calcium and magnesium ions in the treated water,

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thus reducing the hardness. Ion exchange has been used in wastewater applications for the removal of nitrogen, heavy metals, and total dissolved solids.

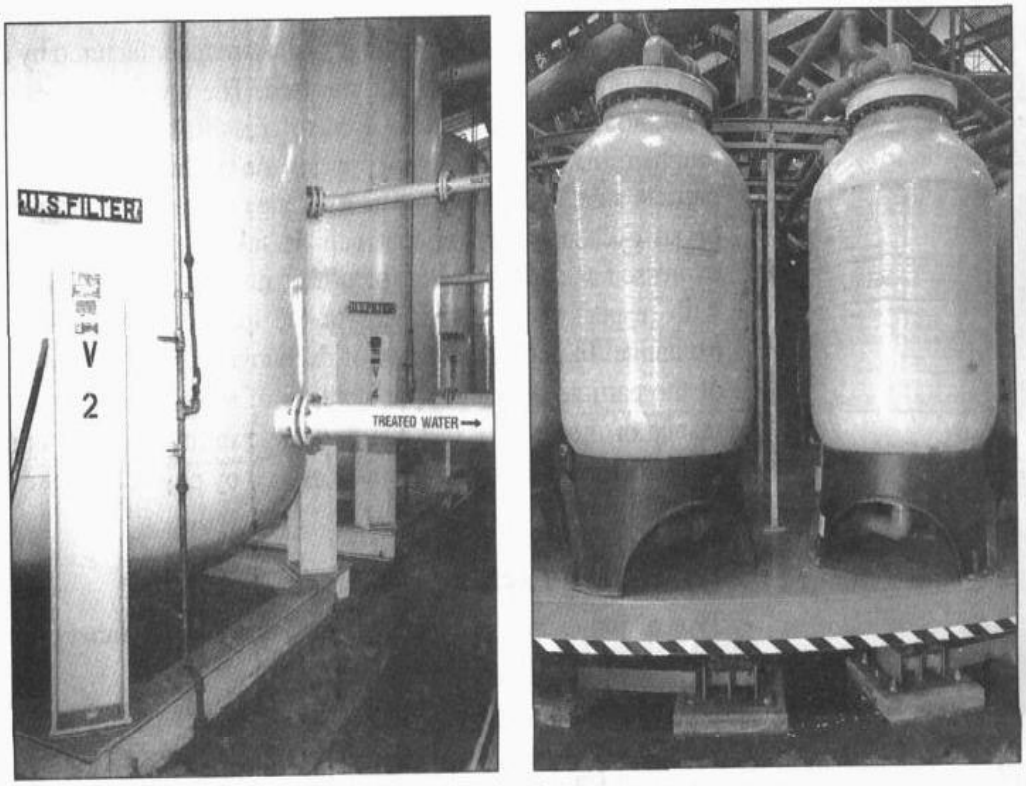
Ion-exchange processes can be operated in a batch or continuous mode. In a batch process, the resin is stirred with the water to be treated in a reactor until the reaction is complete. The spent resin is removed by settling and subsequently is regenerated and reused. In a continuous process, the exchange material is placed in a bed or a packed column, and the water to be treated is passed through it. Continuous ion exchangers are usually of the downflow, packed-bed column type. Wastewater enters the top of the column under pressure, passes downward through the resin bed, and is removed at the bottom. When the resin capacity is exhausted, the column is backwashed to remove trapped solids and is then regenerated. Two examples of commercial ion-exchange reactors are shown on Fig. 11-73.

Ion-Exchange Materials

Naturally occurring ion-exchange materials, known as zeolites, are used for water softening and ammonium ion removal. Zeolites used for water softening are complex aluminosilicates with sodium as the mobile ion. Ammonium exchange is accomplished using a naturally occurring zeolite clinoptilolite. Synthetic aluminosilicates are manufactured, but most synthetic ion-exchange materials are resins or phenolic polymers.

Figure 11-73

Two examples of full-scale ion-exchange installations: (*a*) large downflow packed-bed columns, and (*b*) ion-exchange canisters on a rotating platform. The canisters are rotated so that one canister can be regenerated while the others are in operation.



(a) (b)

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Table 11-36 Classification of ion-exchange resins[a]

Type of resin	Characteristics
Strong-acid cation resins	Strong-acid resins behave in a manner similar to a strong acid, and are highly ionized in both the acid ($R—SO_3H$) and salt ($R—SO_3Na$) form, over the entire pH range
Weak-acid cation resins	Weak-acid cation exchangers have a weak-acid functional group (—COOH), typically a carboxylic group. These resins behave like weak organic acids that are weakly dissociated
Strong-base anion resins	Strong-base resins are highly ionized, having strong-base functional groups such as (OH), and can be used over the entire pH range. These resins are used in the hydroxide (OH) form for water deionization
Weak-base anion resins	Weak-base resins have weak-base functional groups in which the degree of ionization is dependent on pH
Heavy-metal selective chelating resins	Chelating resins behave like weak-acid cation resins but exhibit a high degree of selectivity for heavy-metal cations. The functional group in most of these resins is EDTA, and the resin structure in the sodium form is R—EDTA—Na

[a] Adapted in part from Eckenfelder (2000).

Five types of synthetic ion-exchange resins are in use: (1) strong-acid cation, (2) weak-acid cation, (3) strong-base anion, (4) weak-base anion, and (5) heavy-metal selective chelating resins. The properties of these resins are summarized in Table 11-36.

Most synthetic ion-exchange resins are manufactured by a process in which styrene and divinylbenzene are copolymerized. The styrene serves as the basic matrix of the resin, and divinylbenzene is used to cross-link the polymers to produce an insoluble tough resin. Important properties of ion-exchange resins include exchange capacity, particle size, and stability. The exchange capacity of a resin is defined as the quantity of an exchangeable ion that can be taken up. The exchange capacity of resins is expressed as eq/L or eq/kg (meq/L or meq/g). The particle size of a resin is important with respect to the hydraulics of the ion-exchange column and the kinetics of ion exchange. In general, the rate of exchange is proportional to the inverse of the square of the particle diameter. The stability of a resin is important to the long-term performance of the resin. Excessive osmotic swelling and shrinking, chemical degradation, and structural changes in the resin caused by physical stresses are important factors that may limit the useful life of a resin.

Typical Ion-Exchange Reactions

Typical ion-exchange reactions for natural and synthetic ion-exchange materials are given below.

For natural zeolites (Z):

$$ZNa_2 + \begin{bmatrix} Ca^{2+} \\ Mg^{2+} \\ Fe^{2+} \end{bmatrix} \leftrightarrow Z \begin{bmatrix} Ca^{2-} \\ Mg^{2-} \\ Fe^{2-} \end{bmatrix} + 2Na^{+} \qquad (11\text{-}108)$$

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For synthetic resins (R):

Strong-acid cation exchange:

$$RSO_3H + Na^+ \leftrightarrow RSO_3Na + H^+ \quad (11\text{-}109)$$

$$2RSO_3Na + Ca^{2+} \leftrightarrow (RSO_3)_2Ca + 2Na^+ \quad (11\text{-}110)$$

Weak-acid cation exchange:

$$RCOOH + Na^+ \leftrightarrow RCOONa + H^+ \quad (11\text{-}111)$$

$$2RCOONa + Ca^{2+} \leftrightarrow (RCOO)_2Ca + 2Na^+ \quad (11\text{-}112)$$

Strong-base anion exchange:

$$RR'_3NOH + Cl^- \leftrightarrow RR'_3NCl + OH^- \quad (11\text{-}113)$$

Weak-base anion exchange:

$$RNH_3OH + Cl^- \leftrightarrow RNH_3Cl + OH^- \quad (11\text{-}114)$$

$$2RNH_3Cl + SO_4^{2-} \leftrightarrow (RNH_3)_2SO_4 + 2Cl^- \quad (11\text{-}115)$$

Exchange Capacity of Ion-Exchange Resins

Reported exchange capacities vary with the type and concentration of regenerant used to restore the resin. Typical synthetic resin exchange capacities are in the range of 2 to 10 eq/kg of resin; zeolite cation exchangers have exchange capacities of 0.05 to 0.1 eq/kg. Exchange capacity is measured by placing the resin in a known form. A cationic resin would be washed with a strong acid to place all of the exchange sites on the resin in the H^+ form or washed with a strong NaCl brine to place all of the exchange sites in the Na^+ form. A solution of known concentration of an exchangeable ion (e.g., Ca^{2+}) can then be added until exchange is complete and the amount of exchange capacity can be measured, or in the acid case, the resin is titrated with a strong base. Determination of the capacity of an ion resin based on titration is illustrated in Example 11–15.

Exchange capacities for resins often are expressed in terms of grams $CaCO_3$ per cubic meter of resin (g/m^3) or gram equivalents per cubic meter (g eq/m^3). Conversion between these two units is accomplished using the following expression:

$$\frac{1\ \text{eq}}{\text{m}^3} = \frac{(1\ \text{eq})(50\ \text{g CaCO}_3/\text{eq})}{\text{m}^3} = 50\ \text{g CaCO}_3/\text{m}^3 \quad (11\text{-}116)$$

Calculation of the required resin volume for an ion-exchange process is also illustrated in Example 11–15.

EXAMPLE 11–15 Determination of Ion-Exchange Capacity for a New Resin A column study was conducted to determine the capacity of a cation-exchange resin. In conducting the study, 0.1 kg of resin was washed with NaCl until the resin was in the R-Na form. The column was then washed with distilled water to remove the chloride ion (Cl^-) from the interstices of the resin. The resin was then titrated with a solution of calcium chloride ($CaCl_2$), and

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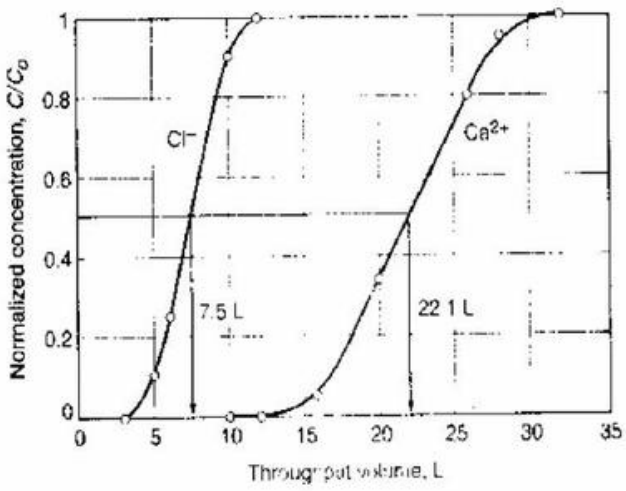
the concentrations of chloride and calcium were measured at various throughput volumes. The measured concentrations of Cl^- and Ca^{2+} and the corresponding throughput volumes are as given below. Using the data given below, determine the exchange capacity of the resin and the mass and volume of a resin required to treat 4000 m^3 of water containing 18 mg/L of ammonium ion NH_4^+. Assume the density of the resin is 700 kg/m^3.

Throughput volume, L	Constituent, mg/L	
	Cl^-	Ca^{2+}
2	0	0
3	Trace	0
5	7	0
6	18	0
10	65	0
12	71	Trace
20	71	13
26	71	32
28	71	38
32	$C_o = 71$	$C_o = 40$

Solution

1. Prepare a plot of the normalized concentrations of Cl^- and Ca^{2+} as a function of the throughput volume. The required plot is given below.



2. Determine the exchange capacity. The exchange capacity (EC) of the resin in meq/kg is

$$EC = \frac{VC_o}{R}$$

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where V = throughput volume between the Cl^- and Ca^{2+} breakthrough curves at $C/C_o = 0.5$

C_o = calcium concentration in meq/L

R = amount of resin in kg

$$EC = \frac{(22.1\ L - 7.5\ L)\left[\frac{(40\ mg/L)}{(20\ mg/meq)}\right]}{0.1\ kg\ of\ resin} = 292\ meq/kg\ of\ resin$$

3. Determine the mass and volume of resin required to treat 4000 m^3 of water containing 18 mg/L of ammonium ion NH_4^+.

   a. Determine the meq of NH_4^+.

$$NH_4^+,\ meq/L = \frac{(18\ mg/L\ as\ NH_4^+)}{(18\ mg/meq)} = 1\ meq/L$$

   b. The required exchange capacity is equal to

$$(1.0\ meq/L)(4000\ m^3)(10^3\ L/m^3) = 4 \times 10^6\ meq$$

   c. The required mass of resin is

$$R_{mass},\ kg = \frac{4 \times 10^6\ meq}{(292\ meq/kg\ of\ resin)} = 13{,}700\ kg\ of\ resin$$

   d. The required volume of resin is

$$R_{vol},\ m^3 = \frac{13{,}700\ kg\ of\ resin}{(700\ kg/m^3)} = 19.6\ m^3\ of\ resin$$

Comment In practice, because of leakage and other operational and design limitations, the required volume of resin will usually be about 1.1 to 1.4 times that computed on the basis of exchange capacity. Also, the above computation is based on the assumption that the entire capacity of the resin is utilized.

Ion-Exchange Chemistry

The chemistry of the ion-exchange process may be represented by the following equilibrium expression for the reaction of constituent A on a cation-exchange resin and constituent B in solution.

$$nR^-A^+ + B^{+n} \leftrightarrow R_n^-B^{+n} + nA^+ \qquad (11\text{-}117)$$

where R^- is the anionic group attached to an ion-exchange resin and A and B are cations in solution. The generalized form of the equilibrium expression for the above reaction is

$$\frac{[A^+]_S^n[R_n^-B^{+n}]_R}{[R^-A^+]_R^n[B^{+n}]_S} = K_{A^+ \to B^{+n}} \qquad (11\text{-}118)$$

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where $K_{A^+ \to B^{++}}$ = selectivity coefficient
$[A^+]_S$ = concentration of A in solution
$[R^-A^+]_R$ = concentration A on the exchange resin

The reactions for the removal of sodium (Na^+) and calcium (Ca^{2+}) ions from water using a strong-acid synthetic cationic-exchange resin *R*, and the regeneration of the exhausted resins with hydrochloric acid (HCl) and sodium chloride (NaCl) are as follows:

Reaction:

$$R^-H^+ + Na^+ \rightarrow R^-Na^+ + H^+ \tag{11-119}$$

$$2R^-Na^+ + Ca^{2+} \rightarrow R_2^-Ca^{2+} + 2Na^+ \tag{11-120}$$

Regeneration:

$$R^-Na^+ + HCl \rightarrow R^-H^+ + NaCl \tag{11-121}$$

$$R_2^-Ca^{2+} + 2NaCl \rightarrow 2R^-Na^+ + CaCl_2 \tag{11-122}$$

The corresponding equilibrium expressions for sodium and calcium are as follows:

For sodium:

$$\frac{[H^+][R^-Na^+]}{[R^-H^+][Na^+]} = K_{H \to Na} \tag{11-123}$$

For calcium:

$$\frac{[Na^+]^2[R^-Ca^{2+}]}{[R^-Na^+]^2[Ca^{2+}]} = K_{Na \to Ca} \tag{11-124}$$

The selectivity coefficient depends primarily on the nature and valence of the ion, the type of resin and its saturation, and the ion concentration in wastewater and typically is valid over a narrow pH range. In fact, for a given series of similar ions, exchange resins have been found to exhibit an order of selectivity or affinity for the ions. Approximate selectivity coefficients for cationic and anionic resins are given in Tables 11–37 and 11–38, respectively. The use of the selectivity coefficients given in these tables is illustrated in Example 11–16.

For synthetic cationic and anionic exchange resins, typical series are

$$Li^+ < H^+ < Na^+ < NH_4^+ < K^+ < Rb^+ < Ag^+ \tag{11-125}$$

$$Mg^{2+} < Zn^{2+} < Co^{2+} < Cu^{2+} < Ca^{2+} < Sr^{2+} < Ba^{2+} \tag{11-126}$$

$$OH^- < F^- < HCO^- < CL^- < Br^- < NO_3^- < ClO_4^- \tag{11-127}$$

In practice, the selectivity coefficients are determined by measurement in the laboratory and are valid only for the conditions under which they were measured. At low concentrations, the value of the selectivity coefficient for the exchange of monovalent ions by divalent ions is, in general, larger than the exchange of monovalent ions by monovalent ions. This fact has, in many cases, limited the use of synthetic resins for the removal of certain substances in wastewater, such as ammonia in the form of the ammonium ion. There are, however, certain natural zeolites that favor NH_4^+ or Cu^{2+}.

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Table 11-37 Approximate selectivity scale for cations on 8 percent cross-linked strong-acid ion-exchange resins[a]

Cation	Selectivity	Cation	Selectivity
Li^+	1.0	Co^{2+}	3.7
H^+	1.3	Cu^{2+}	3.8
Na^+	2.0	Cd^{2+}	3.9
NH_4^+	2.6	Be^{2+}	4.0
K^+	2.9	Mn^{2+}	4.1
Rb^+	3.2	Ni^{2+}	3.9
Cs^+	3.3	Ca^{2+}	5.2
Ag^+	8.5	Sr^{2+}	6.5
Mg^{2+}	3.3	Pb^{2+}	9.9
Zn^{2+}	3.5	Ba^{2+}	11.5

[a] Adapted from Bonner and Smith (1957); see also Slater (1991).

Table 11-38 Approximate selectivity scale for anions on strong-base ion-exchange resins[a]

Cation	Selectivity	Cation	Selectivity
HPO_4^{2-}	0.01	BrO_3^-	1.0
CO_3^{2-}	0.03	Cl^-	1.0
OH^- (Type I)	0.06	CN^-	1.3
F^-	0.1	NO_2^-	1.3
SO_4^{2-}	0.15	HSO_4^-	1.6
CH_3COO^-	0.2	Br^-	3.0
HCO_3^-	0.4	NO_3^-	3.0–4.0
OH (Type II)	0.5–0.65	I^-	18.0

[a] Adapted from Peterson (1953) and Bard (1966).

Anderson (1975) in a classic paper developed a method that can be used to evaluate the effectiveness of a proposed ion-exchange process using strong ionic resins. In the development proposed by Anderson, it is assumed that at 100 percent leakage, the effluent concentration of a constituent is equal to the influent concentration (i.e., equilibrium has been reached). The equilibrium condition can be assumed to be either the limiting operating exchange capacity of the resin or the capacity corresponding to the maximum regeneration level that can be attained. Using this assumption, Eq. (11-118) is converted from concentration units to units of equivalent fractions by making the following substitutions:

$$X_{A^+} = \frac{[A^+]_S}{C} \quad \text{and} \quad X_{B^+} = \frac{[B^+]_S}{C} \tag{11-128}$$

$$X_{A^+} + X_{B^+} = 1 \tag{11-129}$$

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where X_{A^+} and X_{B^+} are the equivalent fractions of A and B in solution and C is the total cationic or anionic concentration in solution.

$$\bar{X}_{A^+} = \frac{[R^- A^+]_R}{\bar{C}} \quad \text{and} \quad \bar{X}_{B^+} = \frac{[R^- B^+]_R}{\bar{C}} \tag{11-130}$$

$$\bar{X}_{A^+} + \bar{X}_{B^+} = 1 \tag{11-131}$$

where $\bar{X}_A$ and $\bar{X}_B$ are the equivalent fractions of A and B in the resin and $\bar{C}$ is the total ionic concentration in the resin (i.e., the total resin capacity in cq/L). Substituting the above terms into Eq. (11-118) and simplifying results in the following expression:

$$\frac{\bar{X}_{B^+} X_{A^+}}{\bar{X}_{A^+} X_{B^+}} = K_{A^+ \to B^+} \tag{11-132}$$

Substituting for X_A and $\bar{X}_A$ in Eq. (11-132) results in

$$\frac{\bar{X}_{B^+}}{1 - \bar{X}_{B^+}} = [K_{A^+ \to B^+}] \left[\frac{X_B}{1 - X_{B^+}} \right] \tag{11-133}$$

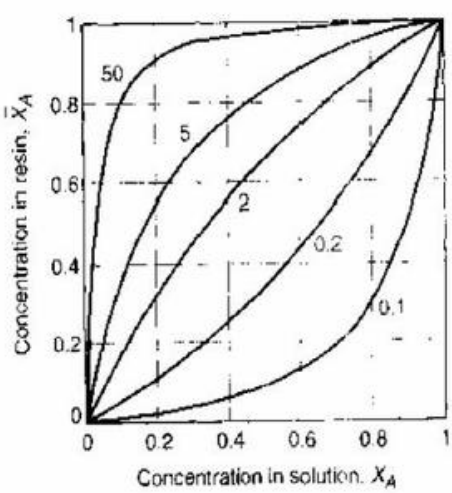
It should be noted that Eq. (11-133) is only valid for exchanges between monovalent ions on fully ionized exchange resins. The distribution of a single monovalent ion A between the solution and the resin for different values of the selectivity coefficient is presented on Fig. 11-74. The distribution curves can be used to assess the effectiveness of a resin for the removal of a given ion, based on the selectivity coefficient.

The following three attributes of Eq. (11-133) were identified by Anderson (1975).

1. The term $\bar{X}_B/(1 - \bar{X}_B)$ corresponds to the state of the resin in an exchange column when the influent and effluent concentrations are the same.
2. The term $\bar{X}_B$ corresponds to the extent to which the resin can be converted to the B^+ form when the resin is in equilibrium with a solution of composition X_B.
3. The term $\bar{X}_B$ also corresponds to the maximum extent of regeneration that can be achieved with a regenerant composition of X_B.

Figure 11-74

Distribution curves for a single monovalent ion A between the solution and the resin for different values of the selectivity coefficient.



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The corresponding equation for exchanges between monovalent and divalent ions on a fully ionized exchange resin is

$$\frac{\bar{X}_{B^{2+}}}{(1-\bar{X}_{B^{2+}})^2} = K_{A^+\to B^{2+}}\left[\frac{\bar{C}}{C}\right]\frac{X_{B^{2+}}}{(1-X_{B^{2+}})^2} \tag{11-134}$$

The application of these equations is illustrated in Example 11-16.

Application of Ion Exchange

As noted previously, ion exchange has been used in wastewater applications for the removal of nitrogen, heavy metals, and total dissolved solids.

For Nitrogen Control. For nitrogen control, the ions typically removed from the waste stream are ammonium, NH_4^+, and nitrate, NO_3^-. The ion that the ammonium displaces varies with the nature of the solution used to regenerate the bed. Although both natural and synthetic ion-exchange resins are available, synthetic resins are used more widely because of their durability. Some natural resins (zeolites) have found application in the removal of ammonia from wastewater. Clinoptilolite, a naturally occurring zeolite, has proved to be one of the best natural exchange resins. In addition to having a greater affinity for ammonium ions than other ion-exchange materials, it is relatively inexpensive when compared to synthetic media. One of the novel features of this zeolite is the regeneration system employed. Upon exhaustion, the zeolite is regenerated with lime $Ca(OH)_2$, and the ammonium ion removed from the zeolite is converted to ammonia because of the high pH. A flow diagram for this process is shown on Fig. 11-75. The stripped liquid is collected in a storage tank for subsequent reuse. A problem that must be solved is the formation of calcium carbonate precipitates within the zeolite exchange bed and in the stripping tower and piping appurtenances. As indicated on Fig. 11-76, the zeolite bed is equipped with backwash facilities to remove the carbonate deposits that form within the filter.

Figure 11-75

Typical flow diagram for the removal of ammonia by zeolite exchange. *Note:* The ammonia removed is removed in ammonia scrubber.

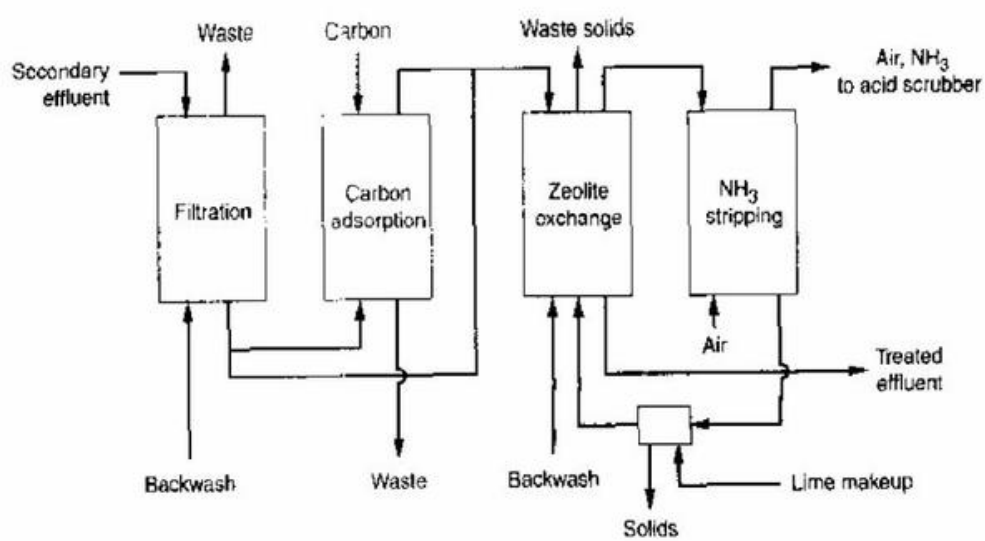

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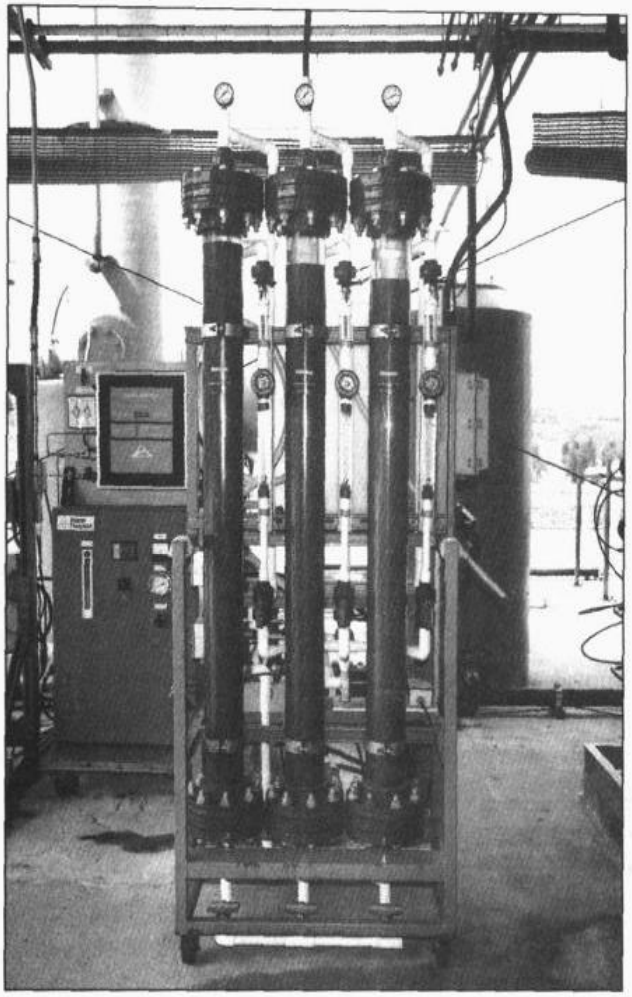
Figure 11-76

Typical ion-exchange test columns used to study the removal of nitrate from water that has been processed with reverse osmosis.



When using conventional synthetic ion-exchange resins for the removal of nitrate, two problems are encountered. First, while most resins have a greater affinity for nitrate over chloride or bicarbonate, they have a significantly lower affinity for nitrate as compared to sulfate, which limits the useful capacity of the resin for the removal of nitrate. The impact of the presence of sulfate on the nitrate removal capacity of conventional resins is illustrated in Example 11-16. Second, because of the lower affinity for nitrate over sulfate, a phenomenon known as *nitrate dumping* can occur. Nitrate dumping occurs when an ion-exchange column is operated past the nitrate breakthrough, at which point the sulfate in the feedwater will displace the nitrate on the resin, causing a release of nitrate. To overcome the problems associated with low affinity and nitrate breakthrough, new types of resins have been developed in which the affinities for nitrate and sulfate have been exchanged. When significant amounts of sulfate are present (i.e., typically greater than 25 percent of the total of the sum of the sulfate and nitrate expressed in meq/L), the use of nitrate-selective resins is advantageous. Because the performance of nitrate-selective resins will vary with the composition of the treated wastewater, pilot testing will usually be required (McGarvey et al., 1989; Dimotsis and McGarvey, 1995). Typical ion-exchange test columns used to study the removal of nitrate from water which has been processed with reverse osmosis are shown on Fig. 11-76.

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EXAMPLE 11-16 Ion Exchange for the Removal of Nitrate without and with Sulfate Present in the Water Nitrate is to be removed from two different treated wastewaters with the compositions given below. For the purpose of illustration, assume a conventional ion-exchange resin will be used.

Wastewater A

Cation	Conc., mg/L	mg/meq	meq/L	Anion	Conc., mg/L	mg/meq	meq/L
Ca^{2+}	82.2	20.04	4.10	HCO_3^-	305.1	61.02	5.00
Mg^{2+}	17.9	12.15	1.47	SO_4^{2-}	0.00	48.03	0.00
Na^+	46.4	23.00	2.02	Cl^-	78.0	35.45	2.20
K^+	15.5	39.10	0.40	NO_3^-	50.0	62.01	0.91
		Σ cations	7.99			Σ anions	8.11

Wastewater B

Cation	Conc., mg/L	mg/meq	meq/L	Anion	Conc., mg/L	mg/meq	meq/L
Ca^{2+}	82.2	20.04	4.10	HCO_3^-	220	61.02	3.61
Mg^{2+}	17.9	12.15	1.47	SO_4^{2-}	79.2	48.03	1.65
Na^+	46.4	23.00	2.02	Cl^-	78.0	35.45	2.20
K^+	15.5	39.10	0.40	NO_3^-	50.0	62.01	0.81
		Σ cations	7.99			Σ anions	8.27

Determine the maximum amount of water that can be processed per liter of a strong-base anion-exchange resin with an exchange capacity of 2.0 eq/L.

Solution:
Wastewater A

1. Estimate the selectivity coefficient (see Table 11–38). To apply Eq. (11–133) the system must be reduced to two components. For this purpose, HCO_3^- and Cl^- are combined into a single component. Using a selectivity value of 4 for nitrate, the selectivity coefficient is estimated as follows:

$$K_{HCO_3^- \to NO_3^-} = \frac{4.0}{0.4} = 10.0$$

$$K_{Cl^- \to NO_3^-} = \frac{4.0}{1.0} = 4.0$$

$$K_{\{(HCO_3^-)(Cl^-)\} \to NO_3^-} = 7.0 \text{ (estimated)}$$

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2. For the equilibrium condition ($C_e/C_o = 1.0$), estimate the nitrate equivalent fraction in solution.

$$X_{NO_3^-} = \frac{0.91}{8.11} = 0.112$$

3. Compute the equilibrium resin composition using Eq. (11–133).

$$\frac{\bar{X}_{B^+}}{1 - \bar{X}_{B^+}} = [K_{A^+ \to B^+}]\left[\frac{X_{B^+}}{1 - X_{B^+}}\right]$$

$$\frac{\bar{X}_{NO_3^-}}{1 - \bar{X}_{NO_3^-}} = 7.0\left[\frac{0.112}{1 - 0.112}\right]$$

$$\bar{X}_{NO_3^-} = 0.47$$

Thus, 47 percent of the exchange sites on the resin can be used for the removal of nitrate.

4. Determine the limiting operating capacity of the resin for the removal of nitrate.

Limiting operating capacity = (2 eq/L of resin)(0.47) = 0.94 eq/L of resin

5. Determine the volume of water that can be treated during a service cycle.

$$\text{Vol} = \frac{(\text{nitrate removal capacity of resin, eq/L of resin})}{(\text{nitrate in solution, eq/L of water})}$$

$$= \frac{(0.94 \text{ eq/L of resin})}{(0.91 \times 10^{-3} \text{ eq/L of water})} = 1033 \frac{\text{L of water}}{\text{L of resin}}$$

Solution: Wastewater B

1. Estimate the selectivity coefficient (see Table 11–38). To apply Eq. (11–134) the system must be reduced to two components. For this purpose, HCO_3^-, Cl^-, and NO_3^- are combined into a single monovalent component. The selectivity coefficient is estimated as follows:

$$K_{HCO_3^- \to SO_4^{2-}} = \frac{0.15}{0.4} = 0.4$$

$$K_{Cl^- \to SO_4^{2-}} = \frac{0.15}{1.0} = 0.15$$

$$K_{NO_3^- \to SO_4^{2-}} = \frac{0.15}{4.0} = 0.04$$

$$K_{[(NO_3^-)(HCO_3^-)(Cl^-)] \to SO_4^{2-}} = 0.2 \text{ (estimated)}$$

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2. For the equilibrium condition ($C_e/C_o = 1.0$), estimate the sulfate equivalent fraction in solution.

$$X_{SO_4^{2-}} = \frac{1.65}{8.27} = 0.2$$

3. Compute the equilibrium resin composition using Eq. (11–134).

$$\frac{\bar{X}_{B^{2+}}}{(1-\bar{X}_{B^{2+}})^2} = [K_{A^+\to B^{2+}}]\frac{\bar{C}}{C}\left[\frac{X_{B^{2+}}}{(1-X_{B^{2+}})^2}\right]$$

$$\frac{\bar{X}_{SO_4^{2-}}}{(1-\bar{X}_{SO_4^{2-}})^2} = 0.2\,\frac{2}{0.00827}\left[\frac{0.2}{(1-0.2)^2}\right]$$

$$\bar{X}_{SO_4^{2-}} = 0.68$$

Thus, 68 percent of the exchange sites on the resin will be in the divalent form at equilibrium. The relative amount of NO_3^- can be estimated by assuming that the remaining 32 percent of the resin sites are in equilibrium with a solution of NO_3^-, HCO_3^-, and Cl^- with the same relative concentration as the feed.

The equivalent fraction of nitrate in the solution will then be

$$X_{NO_3^-} = \frac{0.81}{6.62} = 0.12$$

The selectivity coefficient for the monovalent system is estimated:

$$K_{HCO_3^-\to NO_3^-} = \frac{4.0}{0.4} = 10.0$$

$$K_{Cl^-\to NO_3^-} = \frac{4.0}{1.0} = 4.0$$

$$K_{[(HCO_3^-)(Cl^-)]\to NO_3^-} = 7.0 \text{ (estimated)}$$

Compute the equilibrium resin composition using Eq. (11–133).

$$\frac{\bar{X}'_{B^+}}{1-\bar{X}'_{B^+}} = [K_{A^+\to B^+}]\left[\frac{X'_{B^+}}{1-X'_{B^+}}\right]$$

$$\frac{\bar{X}'_{NO_3^-}}{1-\bar{X}'_{NO_3^-}} = 7.0\left[\frac{0.12}{1-0.12}\right]$$

$$\bar{X}'_{NO_3^-} = 0.5$$

The fraction of the total resin capacity in the nitrate form is then computed.

$$\bar{X}_{NO_3^-} = (1-\bar{X}_{SO_4^{2-}})(\bar{X}'_{NO_3^-}) = (0.32)(0.5) = 0.16$$

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4. Determine the limiting operating capacity of the resin for the removal of nitrate.

   Limiting operating capacity = (2 eq/L of resin)(0.16) = 0.32 eq/L of resin

5. Determine the volume of water that can be treated during a service cycle.

$$\text{Vol} = \frac{(\text{nitrate removal capacity of resin, eq/L of resin})}{(\text{nitrate in solution, eq/L of water})}$$

$$= \frac{(0.32\ \text{eq/L of resin})}{(0.81 \times 10^{-3}\ \text{eq/L of water})} = 395\ \frac{\text{L of water}}{\text{L of resin}}$$

Comment As illustrated in this example, the ionic composition of the wastewater can have a significant effect on the amount of water that can be treated per unit volume of resin, especially where nitrate is to be removed. Because the sulfate is more than 25 percent of the sum of the sulfate and nitrate, the use of a nitrate-selective resin would be advantageous in this application. The approximate nature of these calculations also demonstrates the importance of conducting pilot-plant tests to establish actual throughput volumes.

Removal of Heavy Metals. Metal removal may be required as a pretreatment before discharge to a municipal sewer system. Because of the potential accumulation and toxicity of these metals, it is desirable to remove them from wastewater effluents before release to the environment. Ion exchange is one of the most common forms of treatment used for the removal of metals. Facilities and activities that may discharge wastewater containing high concentrations of metals include metal processing, electronics industries (semiconductors, printed circuit boards), metal plating and finishing, pharmaceuticals and laboratories, and vehicle service shops. High metal concentrations can also be found in leachate from landfills, and stormwater runoff.

Where industries produce effluents with widely fluctuating metal concentrations, flow equalization may be required to make ion exchange feasible, making process design difficult. The economic feasibility of using ion-exchange processes for metal removal greatly improves when the process is used for the removal and recovery of valuable metals. Because it is now possible to manufacture resins for specific applications, the use of resins that have a high selectivity for the desired metal(s) also improves the economics of ion exchange.

Materials used for the exchange of metals include zeolites, weak and strong anion and cation resins, chelating resins, and microbial and plant biomass. Biomass materials are generally more abundant and therefore less expensive when compared to other commercially available resins. Natural zeolites, clinoptilolite (selective for Cs) and chabazite (mixed metals background Cr, Ni, Cu, Zn, Cd, Pb) (Ouki and Kavannagh 1999), have been used to treat wastewater with mixed metal backgrounds. Chelating resins, such as aminophosphonic and iminodiacetic resins, have been manufactured to have a high selectivity for specific metals, such as Cu, Ni, Cd, and Zn.

Ion-exchange processes are highly pH-dependent. Solution pH has a significant impact on the metal species present and the interaction between exchanging ions and

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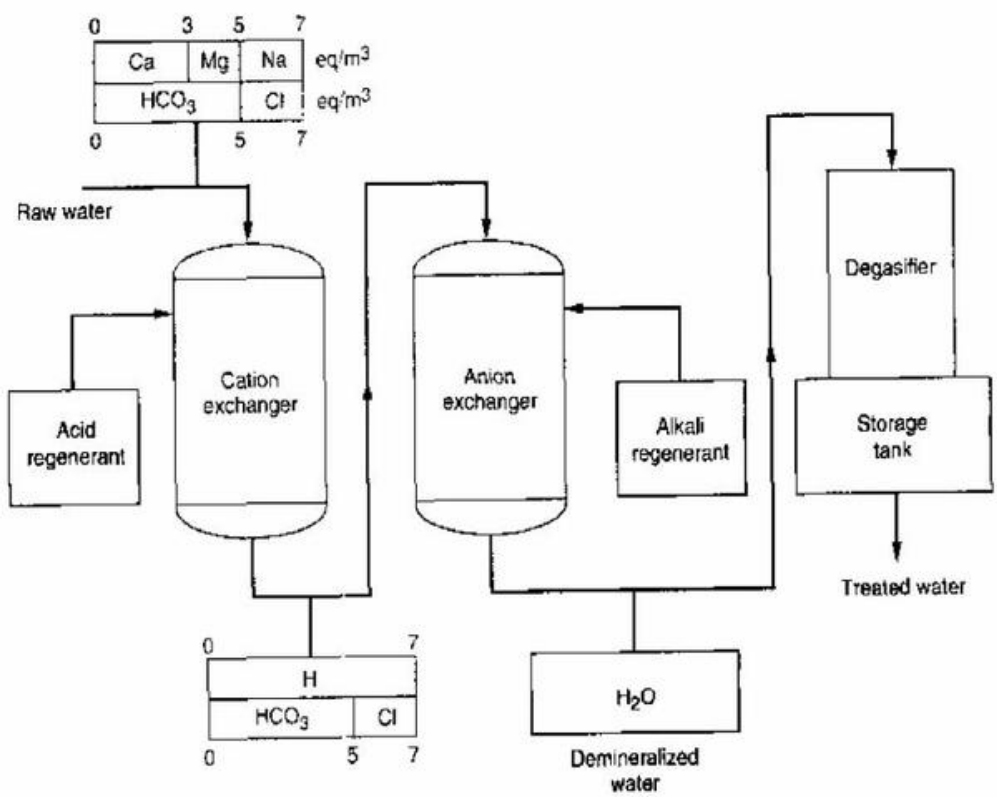
the resin. Most metals bind better at higher pH, due to less competition from protons for adsorption sites. Operating and wastewater conditions determine selectivity of resin, pH, temperature, other ionic species, and chemical background. The presence of oxidants, particles, solvents, and polymers may affect the performance of ion-exchange resins. The quantity and quality of regenerate produced and subsequently requiring management must also be considered.

Removal of Total Dissolved Solids. For the reduction of the total dissolved solids, both anionic- and cationic-exchange resins must be used (see Fig. 11–77). The wastewater is first passed through a cation exchanger where the positively charged ions are replaced by hydrogen ions. The cation-exchanger effluent is then passed over an anionic-exchange resin where the anions are replaced by hydroxide ions. Thus, the dissolved solids are replaced by hydrogen and hydroxide ions that react to form water molecules.

Total dissolved solids removal can take place in separate exchange columns arranged in series, or both resins can be mixed in a single reactor. Wastewater application rates range from 0.20 to 0.40 m^3/m^2·min (5 to 10 gal/ft^2·min). Typical bed depths are 0.75 to 2.0 m (2.5 to 6.5 ft). In reuse applications, treatment of a portion of the wastewater by ion exchange, followed by blending with wastewater not treated by ion exchange, would possibly reduce the dissolved solids to acceptable levels. In some situations, it appears that ion exchange may be as competitive, if not more so, with reverse osmosis.

Figure 11–77

Typical flow diagram for the removal of hardness and for the complete demineralization of water using ion-exchange resins.



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Operational Considerations

To make ion exchange economical for advanced wastewater treatment, it would be desirable to use regenerants and restorants that would remove both the inorganic anions and the organic material from the spent resin. Chemical and physical restorants found to be successful in the removal of organic material from resins include sodium hydroxide, hydrochloric acid, methanol, and bentonite. To date, ion exchange has had limited application because of the extensive pretreatment required, concerns about the life of the ion-exchange resins, and the complex regeneration system required.

High concentrations of influent TSS can plug the ion-exchange beds, causing high headlosses and inefficient operation. Resin binding can be caused by residual organics found in biological treatment effluents. Some form of chemical treatment and clarification is required before ion-exchange demineralization. This problem has been solved partially by prefiltering the wastewater or by using scavenger exchange resins before application to the exchange column.

11-10 ADVANCED OXIDATION PROCESSES

Advanced oxidation processes (AOPs) are used to oxidize complex organic constituents found in wastewater that are difficult to degrade biologically into simpler end products. When chemical oxidation is used, it may not be necessary to oxidize completely a given compound or group of compounds. In many cases, partial oxidation is sufficient to render specific compounds more amenable to subsequent biological treatment or to reduce their toxicity. The oxidation of specific compounds may be characterized by the extent of degradation of the final oxidation products as follows (Rice, 1996):

1. *Primary degradation.* A structural change in the parent compound.
2. *Acceptable degradation (defusing).* A structural change in the parent compound to the extent that toxicity is reduced.
3. *Ultimate degradation (mineralization).* Conversion of organic carbon to inorganic CO_2.
4. *Unacceptable degradation (fusing).* A structural change in the parent compound resulting in increased toxicity.

Theory of Advanced Oxidation

Advanced oxidation processes typically involve the generation and use of the hydroxyl free radical ($HO^{\bullet}$) as a strong oxidant to destroy compounds that cannot be oxidized by conventional oxidants such as oxygen, ozone, and chlorine. The relative oxidizing power of the hydroxyl radical, along with other common oxidants, is summarized in Table 11-39. As shown, with the exception of fluorine, the hydroxyl radical is one of the most active oxidants known. The hydroxyl radical reacts with the dissolved constituents, initiating a series of oxidation reactions until the constituents are completely mineralized. Nonselective in their mode of attack and able to operate at normal temperature and pressures, hydroxyl radicals are capable of oxidizing almost all reduced materials present without restriction to specific classes or groups of compounds, as compared to other oxidants.

Advanced oxidation processes differ from the other treatment processes discussed (such as ion exchange or stripping) because wastewater compounds are degraded rather

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Table 11-39
Comparison of oxidizing potential of various oxidizing agents[a]

Oxidizing agent	Electrochemical oxidation potential (EOP), V	EOP relative to chlorine
Fluorine	3.06	2.25
Hydroxyl radical	2.80	2.05
Oxygen (atomic)	2.42	1.78
Ozone	2.08	1.52
Hydrogen peroxide	1.78	1.30
Hypochlorite	1.49	1.10
Chlorine	1.36	1.00
Chlorine dioxide	1.27	0.93
Oxygen (molecular)	1.23	0.90

[a] From Ozonia (1977).

Table 11-40
Examples of technologies used to produce the reactive hydroxyl free radical, $HO^{\bullet}$[a]

Ozone-based processes	Non-ozone-based processes
Ozone at elevated pH (8 to >10)	H_2O_2 + UV[b]
Ozone + UV_{254} (also applicable in the gas phase)[b]	H_2O_2 + UV + ferrous salts (Fenton's reagent)
Ozone + H_2O_2[b]	Electron-beam irradiation
Ozone + UV_{254} + H_2O_2[b]	Electrohydraulic cavitation
Ozone + TiO_2	Ultrasonics
Ozone + TiO_2 + H_2O_2	Nonthermal plasmas
Ozone + electron-beam irradiation	Pulsed corona discharges
Ozone + ultrasonics	Photocatalysis (UV + TiO_2)
	Gamma radiolysis
	Catalytic oxidation
	Supercritical water oxidation

[a] Adopted in part from Rice (1996).
[b] Processes currently (2001) being used on a commercial scale.

than concentrated or transferred into a different phase. Because secondary waste materials are not generated, there is no need to dispose of or regenerate materials. Additional details on AOPs may be found in Singer and Reckhow (1999).

Technologies Used to Produce Hydroxyl Radicals ($HO^{\bullet}$)

At the present time, a variety of technologies are available to produce $HO^{\bullet}$ in the aqueous phase. The various technologies are summarized in Table 11-40, according to

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whether ozone is used in the reaction. Of the technologies reported in Table 11–40, only ozone/UV, ozone/hydrogen peroxide, ozone/UV/hydrogen peroxide, and hydrogen peroxide/UV are being used on a commercial scale (Rice, 1996).

Ozone/UV. Production of the free radical HO˙ with UV light can be illustrated by the following reactions for the photolysis of ozone (Glaze et al., 1987; Glaze and Kang, 1990):

$$O_3 + UV \text{ (or } h\nu, \lambda < 310 \text{ nm)} \rightarrow O_2 + O(^1D) \quad (11\text{–}135)$$

$$O(^1D) + H_2O \rightarrow HO^{\bullet} + HO^{\bullet} \text{ (in wet air)} \quad (11\text{–}136)$$

$$O(^1D) + H_2O \rightarrow HO^{\bullet} + HO^{\bullet} \rightarrow H_2O_2 \text{ (in water)} \quad (11\text{–}137)$$

where O_3 = ozone
UV = ultraviolet radiation (or $h\nu$ = energy)
O_2 = oxygen
$O(^1D)$ = excited oxygen atom. The symbol (^1D) is a spectroscopic notation used to specify the atomic and molecular configuration (also known as a singlet oxygen)
$HO^{\bullet}$ = hydroxyl radical. The dot (•) that appears next to the hydroxyl and other radicals is used to denote the fact that these species have an unpaired electron.

As shown in Eq. (11–127), the photolysis of ozone in wet air results in the formation of hydroxyl radicals. In water, the photolysis of ozone leads to the formation of hydrogen peroxide [see Eq. (11–128)]. Because the photolysis of ozone in water leads to the formation of hydrogen peroxide, which is subsequently photolyzed to form hydroxyl radicals, the use of ozone in this application is generally not cost-effective. In air, the ozone/UV process can degrade compounds through direct ozonation, photolysis, or reaction with the hydroxyl radical. The ozone/UV process is more effective when the compounds of interest can be degraded through the absorption of the UV irradiation as well as through the reaction with the hydroxyl radicals. A schematic flow diagram of the processes is illustrated on Fig. 11–78.

Figure 11–78

Schematic representation of advanced oxidation process involving the use of ozone and UV radiation.

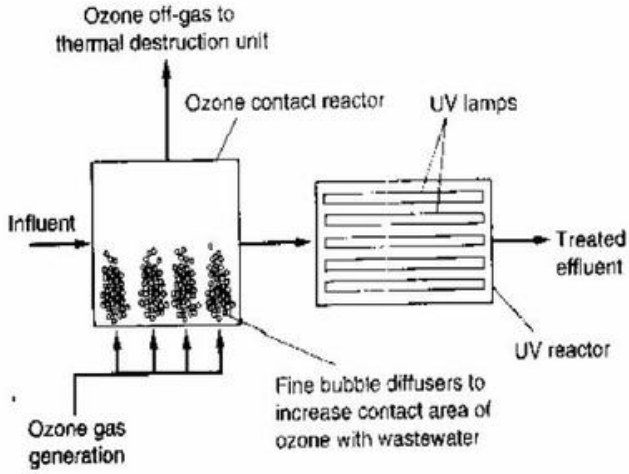

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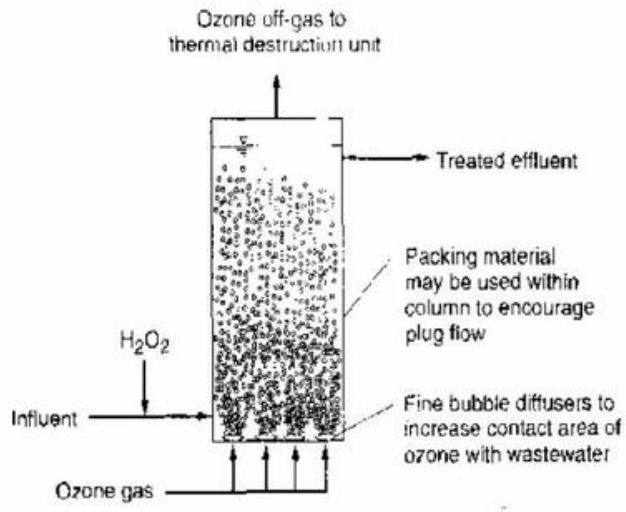
Figure 11-79

Schematic representation of advanced oxidation process involving the use of ozone and hydrogen peroxide.



Ozone/Hydrogen Peroxide. For compounds that do not adsorb UV, AOPs involving ozone/H_2O_2 may be more effective. A schematic flow diagram of the processes is illustrated on Fig. 11-79. Compounds in water such as trichloroethylene (TCE) and perchloroethylene (PCE) have been reduced significantly with AOPs using hydrogen peroxide and ozone to generate HO$^\bullet$ (Karimi et al., 1997). The overall reaction for the production of hydroxyl radicals using hydrogen peroxide and ozone is as follows:

$$H_2O_2 + 2O_3 \rightarrow HO^\bullet + HO^\bullet + 3O_2 \qquad (11\text{-}138)$$

Hydrogen Peroxide/UV. Hydroxyl radicals are also formed when water containing H_2O_2 is exposed to UV light (200 to 280 nm). The following reaction can be used to describe the photolysis of H_2O_2:

$$H_2O_2 + \text{UV (or } h\nu, \lambda \approx 200\text{–}280 \text{ nm}) \rightarrow HO^\bullet + HO^\bullet \qquad (11\text{-}139)$$

In some cases the use of the hydrogen peroxide/UV process has not been feasible because H_2O_2 has a small molar extinction coefficient, requiring high concentrations of H_2O_2 and not using the UV energy efficiently. A schematic flow diagram and a typical installation of the hydrogen peroxide/UV process is shown on Fig. 11-80.

Most recently, the hydrogen peroxide/UV process has been applied to the oxidation of trace constituents found in treated water. The process has been studied for the removal of N-Nitrosodimethylamine (NDMA) and other compounds of concern in treated wastewater including (1) sex and steroidal hormones, (2) human prescription and nonprescription drugs, (3) veterinary and human antibiotics, and (4) industrial and household wastewater products (see Table 2-20 in Chap. 2). At the relatively low concentration range of these compounds encountered in treated water (typically in the μg/L range), their oxidation appears to follow first-order kinetics. The electrical energy required for their oxidation is expressed in EE/O units, defined as the electrical energy input per unit volume per log order of reduction (Bolton et al., 1999). In equation form, the EE/O is given by

$$EE/O = \frac{EE_f}{V[\log(C_i/C_f)]} \qquad (11\text{-}140)$$

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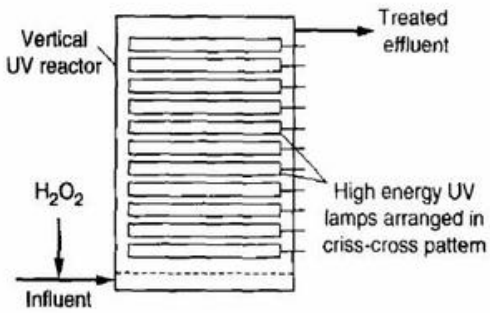
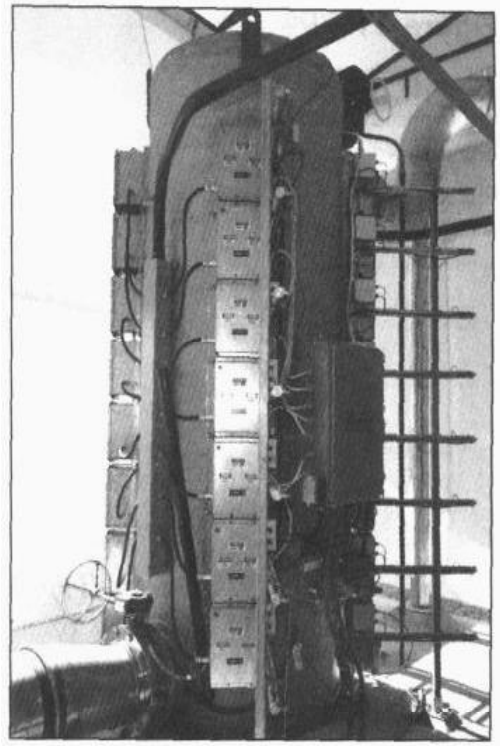
Figure 11-80

Hydrogen peroxide and UV radiation advanced oxidation process: (a) schematic flow diagram. The hydrogen peroxide feed system and storage container are placed in a contained area and (b) photograph of typical vertical-flow UV reactor used to treat a flowrate of about 4000 m^3/d. Fifteen high-energy UV lamps (20 kW each) are employed in the reactor.



(a)



(b)

where EE/O = electrical energy input per log reduction, $kWh/m^3 \cdot$log order of reduction

EE_i = electrical energy input, kWh

V = volume of liquid treated, m^3

C_i = initial concentration, ng

C_f = final concentration, ng

Based on currently available technology (2001), the required EEO value for a one-log order of reduction (i.e., 100 to 10) of NDMA is on the order of 21 to 265 kWh/10^3 $m^3 \cdot$log order (0.08 to 1.0 kWh/10^3 gal·log order) with a 5 to 6 mg/L dose of H_2O_2, although in some cases it did not appear that the peroxide was necessary (Soroushian et al., 2001). The required EE/O value will vary significantly with the characteristics of the treated wastewater.

Other Processes. Other reactions that yield $HO^{\bullet}$ include the reactions of H_2O_2 and UV with Fenton's reagent and the adsorption of UV by semiconductor metal oxides such as TiO_2 suspended in water, which acts as a catalyst. Still others are currently under development.

Applications

Based on numerous studies, it has been found that combined AOPs are more effective than any of the individual agents (e.g., ozone, UV, hydrogen peroxide). AOPs are usually applied to low COD wastewaters because of the cost of ozone and/or H_2O_2 required

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to generate the hydroxyl radicals. Material that was previously resistant to degradation may be transformed into compounds that will require further biological treatment. The application of AOPs for the disinfection of treated wastewater and for the treatment of refractory organic compounds is considered below. Some operational problems are also identified.

Disinfection. Because it was recognized that free radicals generated from ozone were more powerful oxidants than ozone alone, it was reasoned that the hydroxyl free radicals could be used effectively to oxidize microorganisms and refractory organic materials in water and wastewater. Unfortunately, because the half-life of the hydroxyl free radicals is short, on the order of microseconds, it is not possible to develop high concentrations. With extremely low concentrations, the required detention times for microorganism disinfection, based on the CT concept (see Chap 12), are prohibitive.

Oxidation of Refractory Organic Compounds. For the reasons cited above hydroxyl radicals are not used for conventional disinfection; instead they are used more commonly for the oxidation of trace amounts of refractory organic compounds found in highly treated effluents. The hydroxyl radicals, once generated, can attack organic molecules by radical addition, hydrogen abstraction, electron transfer, and radical combination (SES, 1994).

1. *By radical addition.* The addition of the hydroxyl radical to an unsaturated aliphatic or aromatic organic compound (e.g., C_6H_6) results in the production of a radical organic compound that can be oxidized further by compounds such as oxygen or ferrous iron to produce stable oxidized end products. In the following reactions the abbreviation R is used to denote the reacting organic compound.

$$R + HO^{\bullet} \rightarrow ROH \tag{11-141}$$

2. *By hydrogen abstraction.* The hydroxyl radical can be used to remove a hydrogen atom from organic compounds. The removal of a hydrogen atom results in the formation of a radical organic compound, initiating a chain reaction where the radical organic compound reacts with oxygen, producing a peroxyl radical, which can react with another organic compound, and so on.

$$R + HO^{\bullet} \rightarrow R^{\bullet} + H_2O \tag{11-142}$$

3. *By electron transfer.* Electron transfer results in the formation of ions of a higher valence. Oxidation of a monovalent negative ion will result in the formation of an atom or a free radical.

$$R^{n} + HO^{\bullet} \rightarrow R^{n-1} + OH \tag{11-143}$$

4. *By radical combination.* Two radicals can combine to form a stable product.

$$HO^{\bullet} + HO^{\bullet} \rightarrow H_2O_2 \tag{11-144}$$

In general, the reaction of hydroxyl radicals with organic compounds, at completion, will produce water, carbon dioxide, and salts; this process is also known as minera.

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Operational Problems

High concentrations of carbonate and bicarbonate in some wastewater can react with $HO^{\bullet}$ and reduce the efficiency of advanced oxidation treatment processes. Other factors can also affect the treatment process, such as suspended material, pH, type and nature of the residual TOC, and other wastewater constituents. *Because the chemistry of the wastewater matrix is different for each wastewater, pilot testing is almost always required to test the technical feasibility, to obtain usable design data and information, and to obtain operating experience with a specific AOP.*

11–11 DISTILLATION

Distillation is a unit operation in which the components of a liquid solution are separated by vaporization and condensation. Along with reverse osmosis, distillation can be used to control the buildup of salts in critical reuse applications. Because distillation is expensive, its application is generally limited to applications such as (1) a high degree of treatment is required, (2) contaminants cannot be removed by other methods, and (3) inexpensive heat is available. The purpose of this section is to introduce the basic concepts involved in distillation. As the use of distillation for wastewater reclamation is a recent development, the current literature must be consulted for the results of ongoing studies and more recent applications.

Distillation Processes

Over the past 20 years, a variety of distillation processes, employing a variety of evaporator types and methods of using and transferring heat energy, have been evaluated or used. The principal distillation processes are (1) boiling with submerged-tube heating surface, (2) boiling with long-tube vertical evaporator, (3) flash evaporation, (4) forced circulation with vapor compression, (5) solar evaporation, (6) rotating-surface evaporation, (7) wiped-surface evaporation, (8) vapor reheating process, (9) direct heat transfer using an immiscible liquid, and (10) condensing-vapor-heat transfer by vapor other than steam. Of these types of distillation processes, multistage flash evaporation, multiple-effect evaporation, and the vapor-compression distillation appear most feasible for the reclamation of municipal wastewater.

Multiple-Effect Evaporation. In multiple-effect evaporation several evaporators (boilers) are arranged in series, each operating at a lower pressure than the preceding one. In a three-stage, vertical-tube evaporator (see Fig. 11–81), after the influent water is pretreated, it enters the heat exchanger in the last stage (No. 2) and progressively warms as it goes through the heat exchangers in the other effects (i.e., stages). As the water moves through the heat exchangers, it condenses the water vapor emanating from the various effects. When the progressively warmed influent water reaches the first stage, it flows down the internal periphery of vertical tubes in a thin film, which is heated by steam. The wastewater feed to the second effect comes from the bottom of the first effect. If entrainment is kept low, almost all of the nonvolatile contaminants can be removed in a single evaporation step. Volatile contaminants, such as ammonia gas and low-molecular-weight organic acids, may be removed in a preliminary evaporation step, but if their concentration is so small that their presence in the final product is not objectionable, this step with its added cost can be eliminated.

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Figure 11–81

Schematic of multiple-effect evaporation distillation process. *(Adapted from Liu and Liptak, 2000.)*

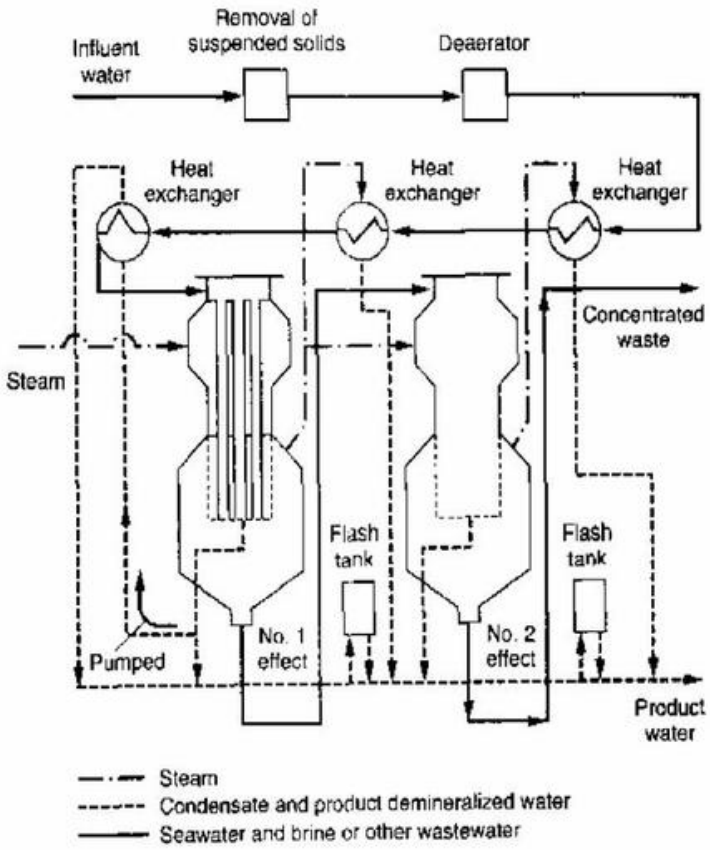
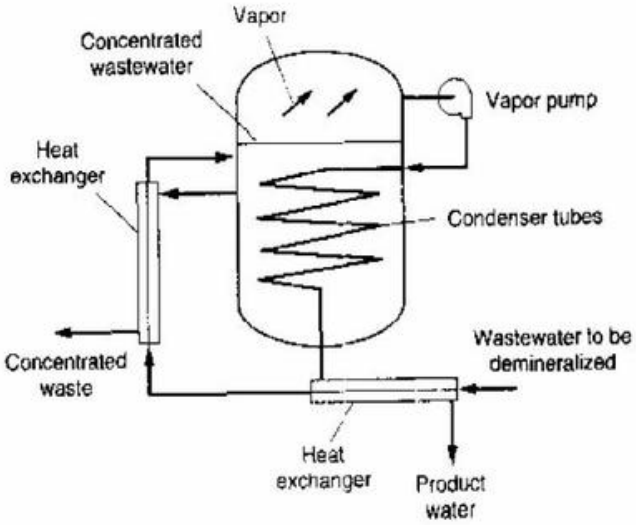


Figure 11–82

Schematic of multistage flash evaporation distillation process.



Multistage Flash Evaporation. Multistage flash evaporation systems have been used commercially in desalination for many years. In the multistage flash process (see Fig. 11–82), the influent wastewater is first treated to remove TSS and deaerated before being pumped through heat transfer units in the several stages of the distillation system, each of which is maintained at a lower pressure. Vapor generation or boiling caused by reduction in pressure is known as *flashing*. As the water enters each stage through a pressure-reducing nozzle, a portion of the water is flashed to form a vapor. In

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turn, the flashed water vapor condenses on the outside of the condenser tubes and is collected in trays (see Fig. 11–82). As the vapor condenses, its latent heat is used to preheat the wastewater that is being returned to the main heater, where it will receive additional heat before being introduced to the first flashing stage. When the concentrated wastewater reaches the lowest-pressure stage, it is pumped out. Thermodynamically, multistage flash evaporation is less efficient than ordinary evaporation. However, by combining a number of stages in a single reactor, external piping is eliminated and construction costs are reduced.

Vapor-Compression Distillation. In the vapor-compression process an increase in pressure of the vapor is used to establish the temperature difference for the transfer of heat. The basic schematic of a vapor-compression distillation unit is shown on Fig. 11–83. After initial heating of the wastewater, the vapor pump is operated so that the vapor under higher pressure can condense in the condenser tubes, at the same time causing the release of an equivalent amount of vapor from the concentrated solution. Heat exchangers can conserve heat from both the condensate and the waste brine. The only energy input required during operation is the mechanical energy for the vapor pump. Hot concentrated wastewater must be discharged at intervals to prevent the buildup of excessive concentrations of salt in the boiler.

Performance Expectations in Reclamation Applications

The theoretical thermodynamic minimum energy required to raise the temperature of wastewater and to provide the latent heat of vaporization is about 2280 kJ/kg. Unfortunately, because of the many irreversibilities in an actual distillation process, the thermodynamic minimum energy requirements are of little practical relevance in the practical evaluation of distillation processes. Typically, about 1.25 to 1.35 times the latent heat of vaporization will be required.

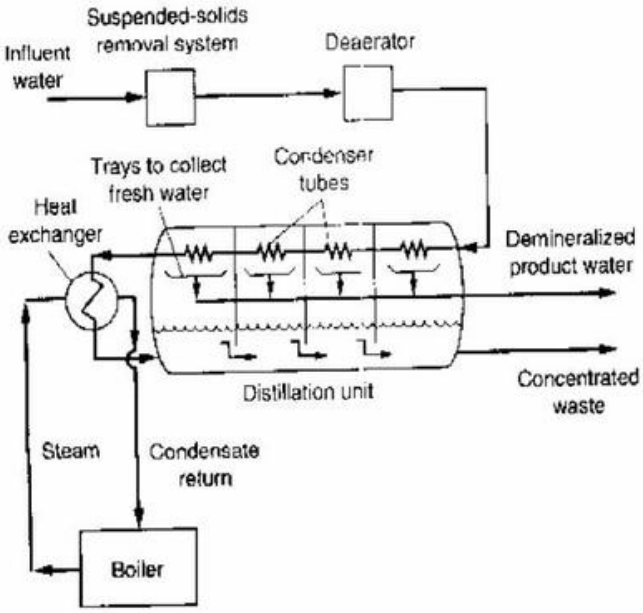

Figure 11–83

Schematic of vapor-compression distillation process. *(Adapted from Liu and Liptak, 2000.)*

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The principal issues with the application of the distillation processes for wastewater reclamation are the carry-over of volatile constituents found in treated wastewater and the degree of subsequent cooling and treatment that may be required to renovate the distilled water. Typical water-quality performance data for a multiple-effect distillation process have been reported for a pilot-scale unit by Rose et al. (1999).

Operating Problems

The most common operating problems encountered include scaling and corrosion. Due to temperature increases, inorganic salts come out of solution and precipitate on the inside walls of pipes and equipment. The control of scaling due to calcium carbonate, calcium sulfate, and magnesium hydroxide is one of the most important design and operational considerations in distillation desalination processes. Controlling the pH minimizes carbonate and hydroxide scales. Most inorganic solutions are corrosive. Cupronickel alloys are used most commonly in seawater desalination. Other metals that are used include aluminum, titanium, and monel.

Disposal of Concentrated Waste

All distillation processes reject part of the influent wastewater. Hence, all of these processes have concentrated wastewater disposal problems. The permissible maximum concentration in the wastewater depends on the solubility, corrosion, and vapor pressure characteristics of the wastewater. Therefore, the waste concentration is an important consideration in process optimization. Disposal of concentrated wastewater brines is essentially the same problem encountered with the membrane process discussed in Section 11–6.

PROBLEMS AND DISCUSSION TOPICS

11–1 The following sieve analysis results were obtained for four different stock sands.

U.S. sieve size designation[a]	Size of opening, mm	Cumulative weight passing, %			
		A	B	C	D
140	0.105	0.4	1.5	0.1	5.0
100	0.149	1.5	4.1	0.8	11.1
70	0.210	4.0	10.0	2.5	20.0
50	0.297	9.5	21.0	8.2	32.0
40	0.420	18.5	40.6	18.5	49.5
30	0.590	31.0	61.0	32.0	62.3
20	0.840	49.0	78.3	58.1	78.3
16	1.190	63.2	90.0	76.3	88.5
12	1.680	82.8	96.0	90.0	94.4
8	2.380	89.0	99.0	96.7	97.8
6	3.360	98.0	99.9	99.0	99.0
4	4.760	100.0	100.0	100.0	100.0

[a]Note: sieve size number 18 has an opening size of 1.0 mm.

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a. For sand sample A, B, C, or D (to be selected by instructor) determine the geometric mean size, the geometric standard deviation, the effective size, and the uniformity coefficient for the stock sand.
b. It is desired to produce from the stock sand a filter sand with an effective size of 0.45 mm and a uniformity coefficient of 1.6. Estimate the amount of stock sand needed to obtain one ton of filter sand.
c. What U.S. standard sieve size should be used to eliminate the excess coarse material?
d. If the sand remaining after sieving in part (c) above is placed in a filter, what backwash rise rate would be needed to eliminate the excess fine material?
e. What depth of sieved material would have to be placed in the filter to produce 600 mm of usable filter sand?
f. On log-probability paper, plot the size distribution of the modified sand. Check against the required distribution and sizes.
g. Determine the headloss through 600 mm of the filter sand specified in part (b) for a filtration rate of 160 $L/m^2 \cdot min$. Assume the sand is stratified and that the maximum and minimum sand sizes are 1.68 (sieve size 12) and 0.297 mm (sieve size 12), respectively. Assume also that $T = 20°C$, α for all layers = 0.4, and $\phi = 0.75$.

An excellent discussion of the procedures involved in developing a usable filter sand from a stock filter sand may be found in Fair et al. (1968).

11-2 Using the equations developed by Fair and Hatch, Kozeny, and Rose, compare the headloss through a 600-mm sand bed. Assume that the sand bed is composed of spherical unisized sand with a diameter of 0.55 mm, the porosity of the sand is 0.40, and the filtration velocity is 240 $L/m^2 \cdot min$ (5 $gal/ft^2 \cdot min$). The temperature is 18°C.

11-3 Using the equation developed by Rose, determine the headloss through a 750-mm sand bed for a filtration rate of 240 $L/m^2 \cdot mm$. Assume that the sand bed is composed of spherical unisized sand with a diameter of 0.40, 0.45, 0.5, and 0.6 mm (size to be selected by instructor) and a porosity of 0.40. The kinematic viscosity is equal to 1.306×10^{-6} m^2/s.

11-4 Solve Problem 11-3 assuming the bed is stratified. Assume that the given sand sizes correspond to the effective size (d_{10}) and that the uniformity coefficient for all of the sand sizes is equal to 1.5.

11-5 If a 0.3-m layer of uniform anthracite is placed on top of the sand bed in Prob. 11-3, determine the ratio of the headloss through the anthracite to that of the sand. Assume that the grain-size diameter of the anthracite is 2.0 mm and porosity is 0.50. Will intermixing occur?

11-6 Given the particle size distribution A, B, C, or D (to be selected by instructor), determine the effective size (d_{10}) and uniformity coefficient UC, and clean water headloss through a stratified bed 600 mm deep. If a layer of anthracite is to be added over 600 mm of sand, determine the effective size required to minimize intermixing.

	Percent of sand retained			
Sieve number	**A**	**B**	**C**	**D**
6–8	2	0	1	0.1
8–10	8	0.1	2	0.7
10–14	10	0.5	4	1.2

(continued)

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(Continued)

Sieve number	Percent of sand retained			
	A	B	C	D
14–20	30	7.4	13	10
20–30	26	32	20	24
30–40	14	30	20	29
40–60	8	25	23	25
Pan	2	5	17	10

11-7 For a given filtration operation it has been found that straining is the operative particulate-matter-removal mechanism and that the change in concentration with distance can be approximated with a first-order equation ($dC/dx = -rC$). If the initial concentration of particulate matter is 10 mg/L, the removal-rate constant is equal to 2.0 cm^{-1}, and the filtration velocity is equal to 100 $L/m^2 \cdot min$, determine the amount of material arrested within the filter in the layer between 3 and 6 cm over a 1.0-h period. Express your answer in $mg/cm \cdot m^2$. Estimate the headloss in the layer at the end of 1.0 h.

11-8 The data in the following table were obtained from a pilot-plant study on the filtration of settled secondary effluent from an activated-sludge treatment process. Using these data, estimate the length of run that is possible with and without the addition of polymer if the maximum allowable headloss is 3 m, the filtration rate is 150 $L/m^2 \cdot min$, and the influent suspended-solids concentration is 15 mg/L. Uniform sand with a diameter of 0.55 mm and a depth of 600 mm was used in the pilot filters.

Depth, cm	Concentration ratio C/C_o	
	With polymer addition	Without polymer addition
0	1.00	1.00
4	0.65	0.70
8	0.26	0.52
12	0.20	0.46
16	0.16	0.42
20	0.15	0.40
24	0.13	0.39
28	0.12	0.38
32	0.12	0.38
36	0.12	0.38
40	0.12	0.38
44	0.12	0.38
48	0.12	0.38
52	0.12	0.38
56	0.12	0.38
60	0.12	0.38

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11-9 Gravity filters are to be used to treat 24,000 m^3/d of settled effluent at a filtration rate of 200 $L/m^2 \cdot min$ (5 $gal/ft^2 \cdot min$). The filtration rate with one filter taken out of service for backwashing is not to exceed 240 $L/m^2 \cdot min$ (6 $gal/ft^2 \cdot min$). Determine the number of units and the area of each unit to satisfy these conditions. If each filter is backwashed for 30 min every 24 h at a wash rate of 960 $L/m^2 \cdot min$ (24 $gal/ft^2 \cdot min$), determine the percentage of filter output used for washing if the filter is out of operation for a total of 30 min/d. What would be the total percentage of filter output used for backwashing if a surface washing system that requires 40 $L/m^2 \cdot min$ (1 $gal/ft^2 \cdot min$) of filtered effluent is to be installed?

11-10 Four different waters are to be desalinized by reverse osmosis using a thin-film composite membrane. For water A, B, C, or D (water to be selected by instructor), determine the required membrane area, the rejection rate, and the concentration of the concentrate stream.

		Water			
Item	**Unit**	**A**	**B**	**C**	**D**
Flowrate	m^3/d	4000	5500	20,000	10,000
Influent TDS	g/m^3	2850	3200	2000	2700
Effluent TDS	g/m^3	200	500	400	225
Flux rate coefficient k_w	m/s	1.5×10^{-6}	1.5×10^{-6}	1.5×10^{-6}	1.5×10^{-6}
Mass transfer rate coefficient, k_i	m/s	1.8×10^{-6}	1.8×10^{-6}	1.8×10^{-6}	1.8×10^{-6}
Net operating pressure	kPa	2750	2500	2800	3000
Recovery	%	88	90	89	86

11-11 Using the data given below, determine the recovery and rejection rates for one of the following reverse osmosis units (unit to be selected by instructor).

		Reverse osmosis unit			
Item	**Unit**	**A**	**B**	**C**	**D**
Flowrate	m^3/d	25,000	50,000	20,000	10,000
Concentrate flowrate	m^3/d	2500	4800	2200	850
Effluent TDS	g/m^3	65	88	125	175
Concentrate	g/m^3	1500	2500	1850	2850

11-12 Using the data given below, determine the flux rate coefficient and the mass transfer rate coefficient.

		Reverse osmosis unit			
Item	**Unit**	**A**	**B**	**C**	**D**
Flowrate	m^3/d	4000	5500	20,000	10,000
Influent TDS	g/m^3	250	3300	5300	2700
Effluent TDS	g/m^3	200	500	400	225
Net operating pressure	kPa	2750	2500	2800	3000
Recovery	%	88	90	89	86

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11-13 Estimate the SDI for the following filtered wastewater samples. If the water is to be treated with reverse osmosis will additional treatment be required?

Test run time, min	Volume filtered, mL			
	A	B	C	D
2		600	400	500
5		300	350	500
10	100	150	0	
20	0	100	0	

11-14 Determine the cost (based on the current price of electricity) to treat a flow of 2500 m^3/d with a TDS concentration of 1300 g/m^3 and a cation-to-anion ratio of 0.13 g-eq/L using an electrodialysis unit. Assume typical values of operation for the electrodialysis unit.

11-15 A wastewater is to be treated with activated carbon to remove residual COD. The following data were obtained from a laboratory adsorption study in which 1 g of activated carbon was added to a beaker containing 1 L of wastewater at selected COD values. Using these data, determine the more suitable isotherm (Langmuir or Fruendlich) to describe the data (sample to be selected by instructor).

Initial COD, mg/L	Equilibrium COD, mg/L			
	A	B	C	D
140	5	10	0.4	5
250	12	30	0.9	18
300	17	50	2	28
340	23	70	4	36
370	29	90	6	42
400	36	110	10	50
450	50	150	35	63

11-16 Using the results from Prob. 11-15, determine the amount of activated carbon that would be required to treat a flow of 5000 m^3/d to a final COD concentration of 20 mg/L if the COD concentration after secondary treatment is equal to 120 mg/L.

11-17 Using the following carbon adsorption data (data set to be selected by instructor), determine the Freundlich capacity factor (mg absorbate/g activated carbon) and Freundlich intensity parameter, 1/n.

Carbon dose, mg/L	Residual concentration, mg/L					
	A	B	C	D	E	F
0	25.9	9.20	9.89	27.5	20.4	9.88
5	17.4	7.36		24.8	19.3	7.95
10	13.2	6.86	8.96	24.2	18.6	7.02

(continued)

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(Continued)

Carbon dose, mg/L	Residual concentration, mg/L					
	A	B	C	D	E	F
25	10.2	3.86	7.83	18.9	16.1	3.66
50	3.6	1.13	5.81	11.8	12.2	0.98
100	2.5	0.22	4.45	2.3	6.7	0.25
150	2.1	0.18	2.98	1.1	3.1	0.09
200	1.4	0.11	2.01	0.9	1.1	0.04

11-18 Determine the theoretical air flowrate required to remove the following compounds in a stripping tower at the indicated concentrations (compound and water to be selected by instructor). Also estimate the height of the stripping tower for a water flowrate of 3000 m^3/d. Values of the Henry's constant may be found in Table 5-36 in Chap. 5.

Compound	Concentration μg			
	Water A		Water B	
	Influent	Effluent	Influent	Effluent
Chlorobenzene	100	5	120	7
Chloroethene	100	5	150	5
TCE	100	5	180	10
Toluene	100	5	200	15

11-19 A quantity of sodium-form ion-exchange resin (5 g) is added to a water containing 2 meq of potassium chloride and 0.5 meq of sodium chloride. Calculate the residual concentration of potassium if the exchange capacity of the resin is 4.0 meq/g of dry weight and the selectivity coefficient is equal to 1.46.

11-20 Determine the exchange capacity for one of the following resins (resin to be selected by instructor). How much resin would be required to treat a flowrate of 4000 m^3/d to reduce the concentration of calcium Ca^{2+} from 125 to 45 mg/L?

Throughput volume, L	Resin A		Resin B	
	Cl^-	Ca^{2+}	Cl^-	Ca^{2+}
0	0	0	0	0
5	2	0	2	0
10	8	0	13	0
15	44	0	29	0
20	65	0	45	0
25	70	0	60	1
30	71	0	69	8
35	71	6	71	17

(continued)

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(Continued)

Throughput volume, L	Resin A		Resin B	
	Cl^-	Ca^{2+}	Cl^-	Ca^{2+}
40	71	20	71	27
45		34	71	354
50		39		39
55		40		40
60		40		40

11-21 Determine the exchange capacity for one of the resins given in Problem 11-20 (resin to be selected by instructor). How much resin would be required to treat a flowrate of 5500 m^3/d to reduce the concentration of calcium Mg^{2+} from 115 to 15 mg/L?

11-22 Four different wastewaters have been reported to have the following ionic composition data. Estimate the selectivity coefficient and determine the amount of wastewater (A, B, C, D, to be selected by instructor) that can be treated by a strong-base ion-exchange resin, per service cycle, for the removal of nitrate. Assume the resin has an ion-exchange capacity of 1.8 eq/L.

Cation	Conc., mg/L	Anion	Concentration, mg/L			
			A	B	C	D
Ca^{2+}	82.2	HCO_3^-	304.8	152	254	348
Mg^{2+}	17.9	SO_4^{2-}	0	0	0	0
Na^+	46.4	Cl^-	58.1	146.3	124	60
K^+	15.5	NO_3^-	82.5	90	21.5	42

11-23 Four different wastewaters have been reported to have the following ionic composition data. Estimate the selectivity coefficient and determine the amount of wastewater (A, B, C, D, to be selected by instructor) that can be treated by a strong-base ion-exchange resin, per service cycle, for the removal of nitrate. Assume the resin has an ion-exchange capacity of 2.5 eq/L.

Cation	Conc., mg/L	Anion	Concentration, mg/L			
			A	B	C	D
Ca^{2+}	82.2	HCO_3^-	321	180	198.5	69
Mg^{2+}	17.9	SO_4^{2-}	65	36.5	124	136
Na^+	46.4	Cl^-	22	95	56	87
K^+	15.5	NO_3^-	46	93	34.5	97

11-24 For the following list of compounds, determine the most suitable method of treatment to reduce the concentration from 100 to 10 $\mu g/L$ using the advanced wastewater-treatment processes discussed in this chapter.

Benzene
Chloroform
Dieldrin
Heptachlor
N-Nitrosodimethylamine
Trichloroethylene (TCE)
Vinyl chloride

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