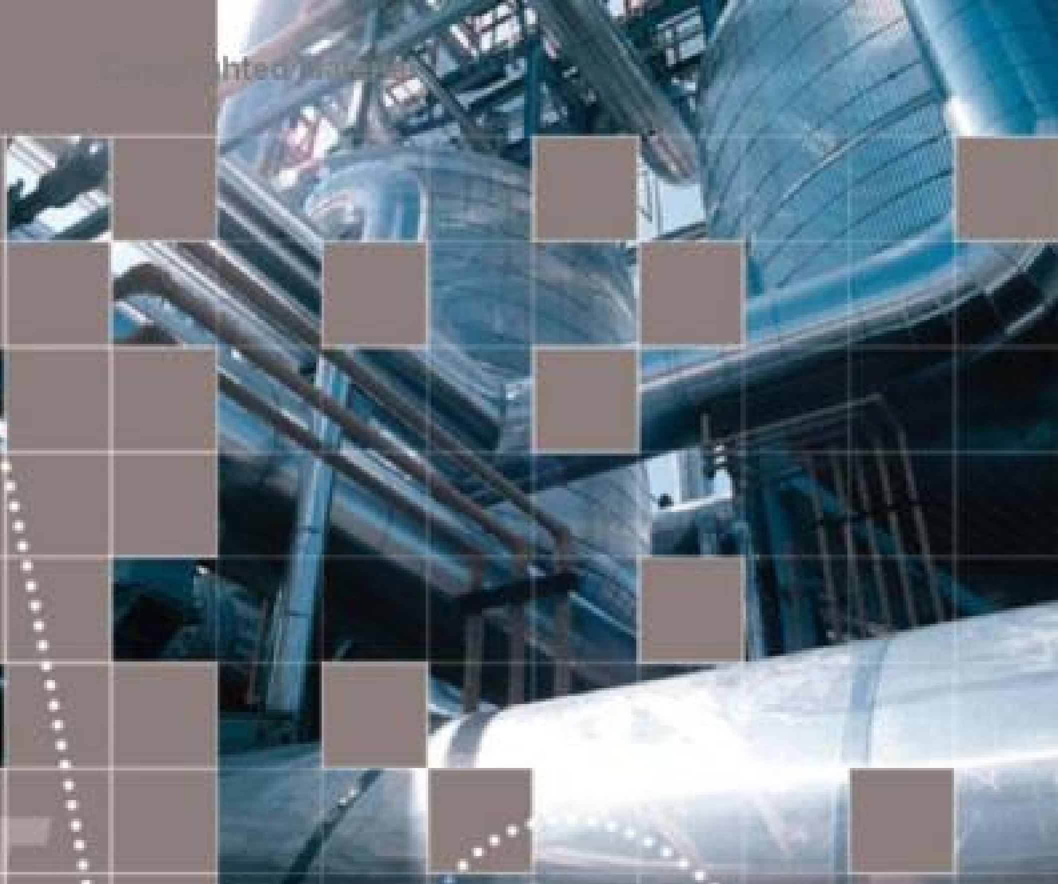

A REAL-TIME APPROACH TO PROCESS CONTROL

SECOND EDITION

WILLIAM Y. SVRCEK

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WILEY

A Real-Time Approach to Process Control

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Second Edition

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Tell me and I forget,
Show me and I may remember,
Involve me and I understand.

Benjamin Franklin
Scientist, Statesman

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Preface

For decades, the subject of control theory has been taught using transfer functions, frequency-domain analysis, and Laplace transform mathematics. For linear systems (like those from the electromechanical areas from which these classical control techniques emerged) this approach is well suited. As an approach to the control of chemical processes, which are often characterized by nonlinearity and large doses of dead time, classical control techniques have some limitations.

In today's simulation-rich environment, the right combination of hardware and software is available to implement a 'hands-on' approach to process control system design. Engineers and students alike are now able to experiment on virtual plants that capture the important non-idealities of the real world, and readily test even the most outlandish of control structures without resorting to non-intuitive mathematics or to placing real plants at risk.

Thus, the basis of this text is to provide a practical, hands-on introduction to the topic of process control by using only time-based representations of the process and the associated instrumentation and control. We believe this book is the first to treat the topic without relying at all upon Laplace transforms and the classical, frequency-domain techniques. For those students wishing to advance their knowledge of process control beyond this first, introductory exposure, we highly recommend understanding, even mastering, the classical techniques. However, as an introductory treatment of the topic, and for those chemical engineers not wishing to specialize in process control, but rather to extract something practical and applicable, we believe our approach hits the mark.

This text is organized into a framework that provides relevant theory, along with a series of hands-on workshops that employ computer simulations that test and allow for exploration of the theory. Chapter 1 provides a historical overview of the field. Chapter 2 introduces the very important and often overlooked topic of instrumentation. In Chapter 3 we ground the reader in some of the basics of single input – single output systems. Feedback control, the elements of control loops, system dynamics including capacitance and dead time, and system modelling are introduced here. Chapter 4 highlights the various PID control modes and provides a framework for understanding control-loop design and tuning. Chapter 5 focuses specifically on tuning. Armed with an understanding of feedback control, control loop structures, and tuning, Chapter 6 introduces some more advanced control configurations including feed-forward, cascade, and override control. Chapter 7 provides some practical rules of thumb for designing and tuning the more common control loops found in industry. In Chapter 8 we tackle a more complex control problem: the control of distillation columns. As with the rest of this text, a combination of theory and applied methodology is used to provide a practical treatment to this complex topic. Chapter 9 introduces the concept of multiple loop controllers. In Chapter 10 we take a look at some of the important issues relating to the plant-wide control problem. Finally, up-to-date information on computer simulation for the workshops can be found on the book website.

Although this text is designed as an introductory course on process control for senior university students in the chemical engineering curriculum, we believe this text will serve as a valuable desk reference for practising chemical engineers and as a text for technical colleges.

We believe the era of real-time, simulation-based instruction of chemical process control has arrived. We hope you'll agree! We wish you every success as you begin to learn more about this exciting and ever changing field. Your comments on and suggestions for improving this textbook are most welcome.

William Y. Svrcek
Donald P. Mahoney
Brent R. Young

Acknowledgements

It would be impossible to mention all of the individuals who contributed to the ideas that form the background of this text. Over the past 5 years, we have interacted with many students, academics, and, perhaps most importantly, practitioners in the field of process control. This, combined with the more than 50 years of cumulative experience among the authors, has led to what we believe is a uniquely practical first encounter with the discipline of chemical process control.

Some who deserve special mention for their influence include Björn Tyréus and Ed Longwell from DuPont, and Paul Fruehauf from Applied Control Engineering. These gentlemen share a passion for the field and a commitment to the practical approach to both teaching and practising process control.

As with any text, many more names were involved in its creation than the three printed on the cover. To those who put in such generous effort to help make this text a reality, we express our sincerest of thanks.

To Dr Barry Cott, Global R&D Leader, Process Control and Optimization, Shell Global Solutions for contributing the section on 'Screening control strategies via steady-state simulation' in Chapter 8.

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To the 1997, 1998, and 1999 fourth-year chemical engineering students at the University of Calgary for their constructive comments on the book and, in particular, the workshops.

Endorsements for the first edition

'As plants are pushed beyond nameplate, it is increasingly obvious that the importance of process control has grown to the point where it is the single biggest leverage point for increasing manufacturing capacity and efficiency. The process engineer, who is best posed to use his process knowledge for getting the most from better control, typically has had just a single course in control. Furthermore, the approach was based on theory rather than on practice, and was immersed in the frequency domain. Real processes are diverse and complex and the view into their behavior is by means of real time trend recordings. This book provides a building block real time approach to understanding and improving process control systems. Practical examples and workshops using models drive home the points and make the principles much more accessible and applicable.'

Gregory K. McMillan, *Senior Fellow, Solutia Inc.*

'At the undergraduate chemical engineering level, the traditional, highly mathematical approach misses the point of what knowledge of control and dynamics the practicing process engineer requires. If BS graduates in chemical engineering simply understood the basics of time based process dynamics and control (capacitance, dead time, PID control action and controller tuning, inventory, throughput, and distillation control), the impact on process design and plant operations throughout the CPI would be immense. Today, these skills are among the least developed in BS chemical engineering graduates, despite having taken the requisite traditional process control course. This text is particularly suitable for any college, university, or technical training program seeking to provide its graduates with a truly practical and applied background in process dynamics and control. With today's widespread commercial availability of high fidelity process simulation software, the understanding gained from this text can be immediately and directly applied.'

Thomas C. Hanson, *Senior Engineering Associate, Praxair, Inc.*

'Several years ago, a recruiter from a major chemical company told me that his company was hesitant to interview students that indicated a first preference in the area of process control because his company "did not have any jobs that made use of Laplace transforms and frequency domain skills". This was an excellent example of the mismatch between what is frequently taught in universities, and what often gets applied in industry. After teaching chemical process control for over 30 years, I feel strongly that good process control is synonymous with good chemical engineering. Industry would be well served if all chemical engineering graduates, regardless of career paths, had a better, more practical working knowledge of process dynamics and control. I think the approach taken in this text is right on target, and is consistent with how we teach at the University of Tennessee. It provides a good hands-on feel for process dynamics and process control, but more importantly, it presents these concepts as fundamentals of chemical engineering. For undergraduate programs looking to transition away from the traditional mathematical-based approach to a more applied, hands-on approach, this text will be an invaluable aid.'

Charles F. Moore, *Professor of Chemical Engineering, University of Tennessee*

'What BS degree chemical engineers need is a base level understanding of differential equations, process dynamics, dynamic modeling of the basic unit operations (in the time domain), basic control algorithms (such as PID), cascade structures and feed forward structures. With these basic tools and an understanding of how to apply them, they can solve most of their control problems themselves. What they do not need is the theory and mathematics that usually surround the teaching of process control such as frequency domain analysis. Graduate education in process control is the place to introduce these concepts.'

James J. Downs, *Senior Engineering Associate, Eastman Chemical Company*

About the authors

William Svrcek is a Professor of Chemical and Petroleum Engineering at the University of Calgary, Alberta, Canada. He received his BSc (1962) and PhD (1967) degrees in Chemical Engineering from the University of Alberta, Edmonton. Prior to joining the University of Calgary he worked for Monsanto Company as a senior systems engineer and as an Associate Professor (1970–1975) in the Department of Biochemical and Chemical Engineering at the University of Western Ontario, London, Ontario. Dr Svrcek's teaching and research interests centre on process simulation control and design. He has authored or co-authored over 150 technical articles/reports and has supervised over 30 graduate students. He has been involved for many years in teaching the continuing education course titled 'Computer Aided Process Design – Oil and Gas Processing' that has been presented world-wide. Most recently this course has been modified to include not only steady-state simulation, but also dynamic simulation and control strategy development and verification. Dr Svrcek was also a senior partner in Hyprotech, now part of Aspen Technology, from its incorporation in 1976. As a Principal, Director, and President (1981–1993) he was instrumental in establishing Hyprotech as a leading international process simulation software company. He is currently providing leadership and vision in process simulation software as the President of Virtual Materials Group Inc. He is a registered Professional Engineer, in both Alberta and Ontario, and a member of professional societies that include The Canadian Society for Chemical Engineering, American Institute for Chemical Engineers, Canadian Gas Processors Association and the Instrument Society of America.

Donald Mahoney is co-founder and Chief Operating Officer with BDMetrics, Inc., a company that develops and markets web-based analytics software. Mr Mahoney earned a Bachelor's Degree in Mechanical Engineering from Penn State, a Master's Degree in Control Theory from Purdue University, and an MBA from the University of Delaware. Mr Mahoney has held research and teaching positions at the US Navy's Applied Research Lab and Purdue University, where he was awarded the staff's 'Outstanding Teaching Award'. He has also lectured extensively on process simulation and control topics, and has published a number of journal articles in the field. Prior to joining BDMetrics, Mr Mahoney was Vice President with AEA Technology Engineering Software/Hyprotech where he led the introduction and launch of more than a half dozen design, modelling and optimization software products. He has held industrial positions at General Motors and DuPont as a control systems engineer and process modelling and control consultant. While at DuPont, Mr Mahoney was involved in the development and support of the chemical industry's first object-oriented dynamic simulation package, TMODS^{TM}.

Brent Young is Senior Lecturer of Chemical and Materials Engineering at the University of Auckland, New Zealand. He received his BE (1986) and PhD (1993) degrees in Chemical and Process Engineering from the University of Canterbury, New Zealand. Prior to his graduate studies, he worked as a Chemical Engineer for Ravensdown Fertilizer Coop's Super Phosphate Plant in Christchurch and developed a process model for the simulation of a rock phosphate grinding circuit. In 1991, he joined the University of Technology in Sydney, Australia, as a lecturer, received tenure in 1994 and was promoted to Senior Lecturer in 1996, continuing his research in the areas of modelling and control of processes, particularly industrial processes. He was an Associate Professor of Chemical and Petroleum Engineering at the University of Calgary from late 1998 to the end of 2005. He joined the University of Auckland in January 2006. He is a registered Professional Engineer and a member of a number of professional societies. His research is centred on the two major areas of process simulation and control, and process design and development – particularly the processing of carbonaceous substances.

1

A brief history of control and simulation

In order to gain an appreciation for process control and simulation it is important to have some understanding of the history and driving force behind the development of both control and simulation. Rudimentary control systems have been used for centuries to help humans use tools and machinery more efficiently and effectively. However, only in the last century has more time and effort been devoted to developing a greater understanding of controls and more sophisticated control systems. The expansion of the controls field has aided the growth of process simulation from relative obscurity to the indispensable and commonplace tool that it is today.

1.1 Control

Feedback control can be traced back as far as the early third century BC [1]. During this period, Ktesibios of Alexandria employed a float valve similar to the one found in today's automobile carburettors to regulate the level in the water clocks of that time [2]. Three centuries later, Heron of Alexandria described another float-valve water-level regulator similar to that used in toilet water-tanks [1]. Arabic water-clock builders used this same control device as late as 1206. The Romans also made use of this first control device in regulating the water levels in their aqueducts. The level regulating device or float valve remained unknown to Europeans and was reinvented in the eighteenth century to regulate the water levels in steam boilers and home water tanks.

The Europeans did, however, invent a number of feedback control devices, namely the thermostat or bimetallic temperature regulator, the safety relief valve, and the windmill fantail. In 1620, Cornlis Drebbel [2], a Dutch engineer, used a bimetallic temperature regulator to control the temperature of a furnace. Denis Papin [2], in 1681, used weights on a pressure cooker to regulate the pressure in the vessel. In 1745, Edmund Lee [1] attached a fantail at right angles to the main sail of a windmill, thus always keeping the main windmill drive facing into the wind. It was not until the Industrial Revolution, particularly in England, that feedback devices became more numerous and varied.

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One-port automata (open loop) evolved as part of the Industrial Revolution and focused on a flow of commands that mechanised the functions of a human operator. In 1801, Joseph Farcot [3] fed punched cards past a row of needles to program patterns on a loom; and in 1796, David Wilkinson [4] developed a copying lathe with a cutting tool positioned by a follower on a model. Oliver Evans [3] built a water-powered flourmill near Philadelphia, in 1784, using bucket and screw conveyors to eliminate manual intervention. Similarly, biscuit making was automated for the Royal Navy in 1833, and meat processing was mechanised in America during the late 1860s. Henry Ford used the same concept for his 1910 automobile assembly plant automation. Unit operations, pioneered by Allen Rogers of the Pratt Institute [5] and Arthur D. Little of MIT [5], led to continuous chemical processing and extensive automation during the 1920s.

The concept of feedback evolved along with the development of steam power and steam-powered ships. The valve operator of Humphrey Potter [6] utilised piston displacement on a Newcomen engine to perform a deterministic control function. However, the flyball governor designed by James Watt [6] in 1769 modulated steam flow to overcome unpredictable disturbances and became the archetype for single-loop regulatory controllers. Feedback was accompanied by a perplexing tendency to overshoot the desired operating level, particularly as controller sensitivity increased. The steam-powered steering systems of the ships of the mid 1800s used a human operator to supply feedback, but high rudder positioning gain caused the ship to zigzag along its course. In 1867, Macfarlane Gray [1] corrected the problem with a linkage that closed the steering valve as the rudder approached the desired point. In 1872, Leon Farcot [1] designed a hydraulic system such that a displacement representing rudder position was subtracted from the steering position displacement, and the difference was used to operate the valve. The helmsman could then indicate a rudder position, which would be achieved and maintained by the servo motor.

Subsequent refinements of the servo principle were largely empirical until Minorsky [7], in 1922, published an analytical study of ship steering which considered the use of proportional, derivative, and second derivative controllers for steering ships and demonstrated how stability could be determined from the differential equations. In 1934, Hazen [8] introduced the term ‘servomechanism’ for position control devices and discussed the design of control systems capable of close tracking of a changing set point. Nyquist [9] developed a general and relatively simple procedure for determining the stability of feedback systems from the open-loop response, based on a study of feedback amplifiers.

Experience with and theory in mechanical and electrical systems were, therefore, available when World War II created a massive impetus for weapon controls. While the eventual social benefit of this and subsequent military efforts is not without merit, the nature of the incentives emphasises the irony seen by Elting Morison [10]. Just as we attain a means of ‘control over our resistant natural environment we find we have produced in the means themselves an artificial environment of such complexity that we cannot control it’.

Although the basic principles of feedback control can be applied to chemical processing plants as well as to amplifiers or mechanical systems, chemical engineers were slow to adapt the wealth of control literature from other disciplines for the design of process control schemes. The unfamiliar terminology was one major reason for the delay, but there was also the basic difference between chemical processes and servomechanisms, which delayed the development of process control theory and its implementation. Chemical plants normally operate with a constant set point, and large-capacity elements help to minimize the effect of disturbances, whereas these would tend to slow the response of servomechanisms. Time delay or transport lag is frequently a major factor in process control, yet it is rarely mentioned in the literature on servomechanisms. In process control systems, interacting first-order elements and distributed resistances are much more common than the second-order elements found in the control of mechanical and electrical systems. These differences made many of the published examples of servomechanism design of little use to those interested in process control.

A few theoretical papers on process control did appear during the 1930s. Notable among these was the paper by Grebe *et al.* [11] that discussed the problem of pH control and showed the advantages of using derivative action to improve controller response. Callender *et al.* [12] showed the effect of time delay on the stability and speed of response of a control system. However, it was not until the mid 1950s that the first texts on process control were published by Young, in 1954 [13], and Ceaglske, in 1956 [14]. These early classical process control texts used techniques that were suitable prior to the availability of computers, namely frequency response, Laplace transforms, transfer function representation and linearization. Between the late 1950s and the 1970s many texts appeared, generally following the pre-computing classical approach, notably those by EcKman [15], Campbell [16], Coughanowr and Koppel [17], Luyben [18], Harriott [19], Murrill [20], and Shinskey [21]. Process control became an integral part of every chemical engineering curriculum.

1.2 Simulation

Prior to the 1950s, calculations had been done manually (using a slide rule) on mechanical or electronic calculators. In 1950, Rose and Williams [22] wrote the first steady-state, multistage binary distillation tower simulation program. The total simulation was written in machine language on an IBM 702, a major feat with the hardware of the day. The general trend through the 1950s was steady-state simulation of individual units. The field was moving so rapidly that by 1953 the American Institute of Chemical Engineers (AIChE) had the first annual review of Computers and Computing in Chemical Engineering. The introduction of FORTRAN by IBM, in 1954, provided the impetus for the chemical process industry to embrace computer calculations. The 1950s can be characterized as a period of discovery [23].

From the early 1960s to the present day, steady-state process simulation has moved from a tool used only by experts to a software tool used daily to perform routine calculations. This was made possible by the advances in computing hardware, the most significant of which has been the proliferation of powerful desktop personal computers

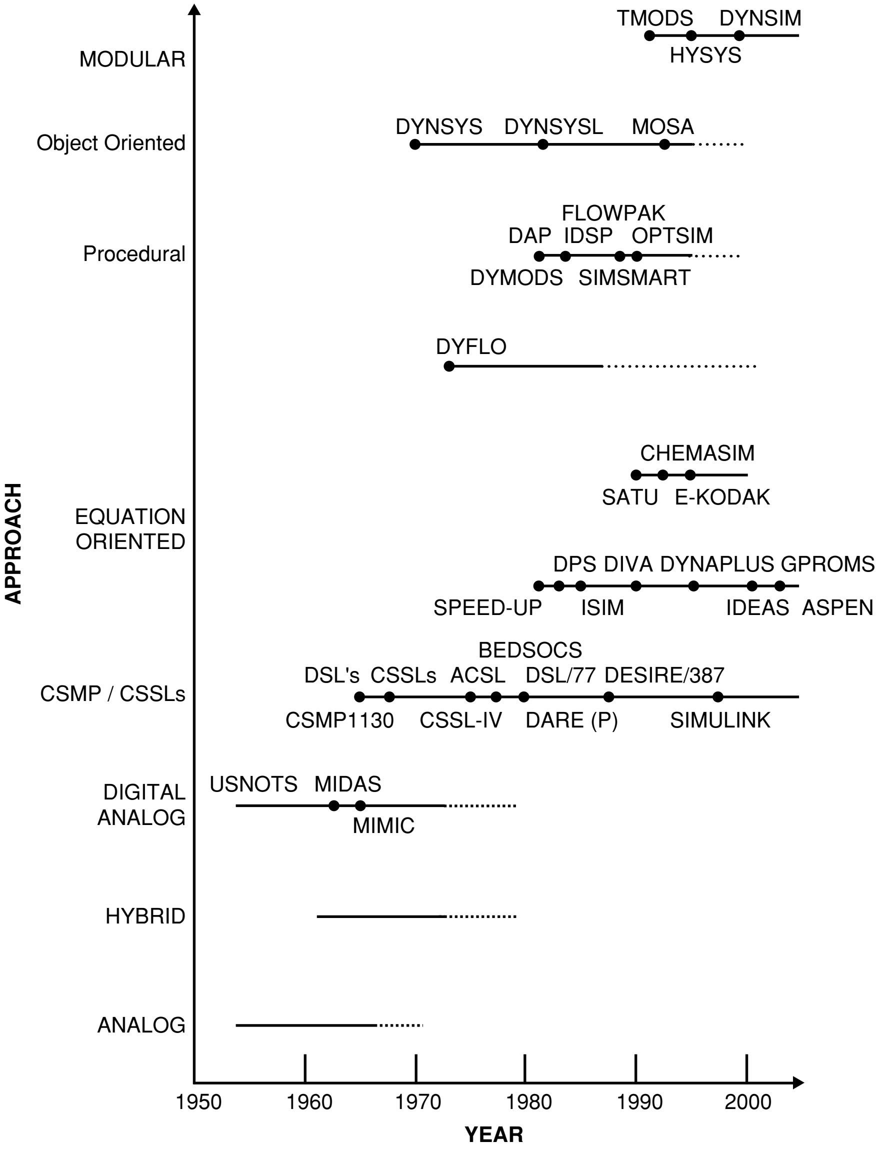

Figure 1.1 Development of dynamic process simulators

(PCs), the development of Windows-based systems software, and the development of object-oriented programming languages. This combination of inexpensive hardware and system tools has led to the proliferation of exceptionally user-friendly and robust software tools for steady-state process simulation and design. Dynamic simulation naturally developed along with the steady-state simulators [24]. Figure 1.1 presents a summary of the growth of dynamic process simulation.

During the 1960s, the size of the analog computer controlled the size of the simulation. These analog computers grew from a few amplifier systems to large systems of a hundred or more amplifiers and finally in the late 1960s to hybrid computers [25]. It was recognized very early that the major disadvantages of analog computers were problem size and dynamic range, both of which were limited by hardware size. Hybrid computers were an attempt to mitigate some of these problems. However, hybrid computers of the late 1960s and early 1970s still had the following problems that limited their general acceptance [25]:

1 Hybrid computers required detailed knowledge of operation for both the analog and digital computers. This translated into long training periods (of a week or more) before an engineer was able to work with the hybrid computer.

2 Hybrid computer simulations were composed of two parts: the analog and digital computer portions. This made debugging complicated, since both parts had to be debugged and then integrated.

3 Documentation was required for both parts of the hybrid simulation, i.e. analog and digital. The analog part was documented by using wiring diagrams. These wiring diagrams quickly became outdated, as changes were made to the analog board that were not always added to the wiring diagram (human nature).

4 Simulations using hybrid computers were extremely time consuming. An engineer had to reserve time in the hybrid simulation laboratory and work in this laboratory in order to solve the problem. This time was devoted entirely to solving one problem and removed the engineer from other effective work.

5 For the majority of simulations, hybrid computers were more expensive to use than digital computers.

Engineers were searching for a dynamic simulator that paralleled steady-state simulators being developed during the late 1960s and early 1970s. Early attempts simply moved the analog to a digital formation (CSMPs, Pactolus, etc.) by providing numerical integration algorithms and a suitable programming syntax. Later versions of these block-oriented dynamic simulators provided more functionality and an improved programming methodology. This approach resulted in various Continuous System Simulation Languages (CSSLs), of which ACSL [26] is the most widely used.

Parallel to the previous approach has been the development of equation-based numerical solvers like SPEEDUP [27]. These tools are aimed at the specialist who has considerable experience in using the tool, knows how to model various processes in terms of their fundamental equations, and is willing to spend considerable time entering code and data into input files, which are compiled, edited and debugged before they yield results of time plots for selected variables over fixed time periods. These equation-based dynamic simulation packages were very much the realm of the expert. Concepts such as ease of use, complex thermodynamic packages, and libraries of reusable unit operations had not migrated to these dynamic simulators.

The first attempts to provide a modular-based dynamic process simulator were DYFLO and DYNSYS [28]. These two early modular simulators differed in their approach. DYFLO provided the simulator with a suite of FORTRAN routines that were linked via a program written by the user. Hence, it was to some extent cumbersome, but useable. DYNSYS [28], on the other hand, provided a key word structure much like the steady-state simulators of the era allowing the user to build a dynamic simulation. Both simulators found limited use owing to the difficulty of producing a simulation, and the actual run times on the computer hardware of the time were often greater than real time.

During the late 1970s and throughout the 1980s only equation-oriented simulators were used. There was a continuing effort to develop and extend dynamic models of plants and use these for control system development. Many companies, from necessity, had groups using this approach to develop specific plant dynamic simulations and subsequently using these simulations for control design and evaluation. Marquardt at CPCIV [29] presented a paper summarizing key developments and future challenges in dynamic process simulation. Since that review, two additional dynamic process simulators have appeared, i.e. Odysseo [30] and Ideas [31].

The key benefits of dynamic simulation [32] are related to the improved process understanding that it provides; plants are, by their nature, dynamic. By understanding the process more fully, several benefits follow naturally. These include improvements in control system design, improvements in the basic operation of the plant, and improvements in training for both operators and engineers.

Control system design is, unfortunately, still often left until the end of the design cycle. This practice frequently requires an elaborate control strategy in order to make the best of a poor design. Dynamic simulation, when involved early in the design phase, can help to identify the important operability and control issues and influence the design accordingly. Clearly, the ideal is not just to develop a working control strategy, but also to design a plant that is inherently easy to control.

Using a rigorous dynamic model, control strategies can be designed, tested, and even tuned prior to start up. With appropriate hardware links, dynamic models may even be used to checkout distributed control system (DCS) or other control system configurations. All of these features make dynamic simulation ideally suited to control applications.

Another benefit involves reconciling trade-offs between steady-state optimizations and dynamic operability. To minimise capital expenditures and operating utility costs, many plant designers have adopted the use of steady-state optimisation techniques. As a result, plant designs have become more complex and much more highly integrated and interactive. Examples include extensive heat exchange networks, process recycles, and minimum holdup designs. While such designs may optimize the steady-state flowsheet, they present particular challenges to plant control and operations engineers, usually requiring advanced control strategies and a well-trained operating staff. This trade-off between steady-state optimisation and dynamic operability is classic and can only be truly reconciled using dynamic simulation.

Once a plant is in operation, manufacturing personnel are continually looking for ways to improve quality, minimise waste, maximise yield, reduce utility costs, and often to increase capacity. It is in this area of process improvements where dynamic simulation has, perhaps, the most value-adding impact. This is also the area where it is most important to minimise the usage barriers for dynamic simulation. Since plant operating personnel are typically busy with the day-to-day operation of the plant, simulation tools that are difficult to understand and use will never see any of the truly practical and value-adding applications. By allowing plant engineers quickly and easily to test theories, illustrate concepts, or compare alternative control strategies or operating schemes, dynamic simulation can have a tremendous cumulative benefit.

Over the past several years, the industry has begun to focus a great deal of attention and interest on dynamic simulation for training purposes. As mentioned earlier, the increased complexity of the plants being designed today requires well-trained operating personnel. In order to be effective, the training simulator should be interactive, realistic, and run in real time. By running a relatively high-fidelity model, operators can test 'what if' scenarios, practise start up and shutdown sequences, and respond to failure and alarm situations.

More recently, training simulators have provided links to a variety of DCS platforms. By using the actual control hardware to run a dynamic model of the plant, operators have the added benefit of training on the same equipment that will be used to operate the real process.

It is important at this point to introduce the notion of breadth of use for a model. We have discussed the use of dynamic simulation for design, control, operations, and now for training. Indeed, it would be beneficial if the same model used to design the plant, develop its controls, and study its operation could be used as the on-line training simulator for the DCS. While this may seem obvious, it is difficult to find examples of such applications. This is primarily due to the absence of commercial simulation tools that provide sufficient breadth of functionality – both engineering functionality and usability.

With all of the benefits to dynamic simulation, why is it that this technology has only begun to see more widespread use recently? To answer this question, it is helpful to continue with the history of simulation and to consider the unique set of skills required to develop a dynamic simulation from first principles.

First, an understanding of and access to the basic data relating to the physical properties of the chemical system is needed. This includes the vapour–liquid equilibrium (VLE) and any reaction equations involved. Second, a detailed understanding of the heat and material balance relationships in the process equipment is required. Third, knowledge of appropriate numerical techniques to solve the sets of differential and algebraic equations is needed. Finally, experience in striking a balance between rigour and performance is needed in order to build a model that is at the same time useful and useable. Thus, there is indeed a unique set of skills required to design a first-principles dynamic simulator.

Because of the computational load, dynamic simulations have been reserved for large mainframe or mini-computers. An unfortunate feature of these large computer systems was their often cumbersome user interfaces. Typically, dynamic simulations were run in a batch mode where the model was built with no feedback from the program, then submitted to the computer to be solved for a predetermined length of simulation time. Only when the solution was reached could the user view the results of the simulation study.

With this approach, 50 to 80 per cent of the time dedicated to a dynamic study was consumed in the model building phase. Roughly 20 per cent was dedicated to running the various case studies, and 10 per cent to documentation and presentation of results. This kind of cycle made it difficult for a casual user to conduct a study or even to run a model that someone else had prepared. While the batch-style approach consumed a disproportionate amount of time setting up the model, the real drawback was the lack of any interaction between the user and the simulation. By preventing any real interaction with the model as it is being solved, batch-style simulation sessions are much less effective. Additionally, since more time and effort are spent building model structures, submitting, and waiting for batch input runs, a smaller fraction of time is available to gain the important process understanding through 'what if' sessions.

Thus, between the sophisticated chemical engineering, thermodynamics, programming and modelling skills, the large and expensive computers, and cumbersome and inefficient user interfaces, it is not surprising that dynamic simulation has not enjoyed widespread use. Normally, only the most complex process studies and designs justified the effort required to develop a dynamic simulation. We believe that the two most significant factors in increased use of dynamic simulation are [24]:

- the growth of computer hardware and software technology;
- the emergence of new ways of packaging simulation.

As indicated previously, there has been a tremendous increase in the performance of PCs accompanied by an equally impressive drop in their prices. For example, it is not uncommon for an engineer to have a PC with memory of upwards of 1 GB, a 100 GB hard drive, and a large flat-screen graphics monitor on his desktop costing less than $1500. Furthermore, a number of powerful and interactive window environments have been developed for the PC and other inexpensive hardware platforms. Windows (2000, NT, XP, etc.), X-Windows, and Mac Systems are just some examples.

The growth in the performance and speed of the PC has made the migration of numerically intense applications to PC platforms a reality. This, combined with the flexibility and ease of use of the window environments, has laid the groundwork for a truly new approach to simulation.

There are literally thousands of person-years of simulation experience in the industry. With the existing computer technology providing the framework, there are very few reasons why most engineers should have to write and compile code in order to use dynamic simulation. Model libraries do not provide the answer, since they do not eliminate the build–compile–link sequence that is often troublesome, prone to errors, and intimidating to many potential users. Given today's window environments and the new programming capabilities that languages such as object-oriented C++ provide, there is no need for batch-type simulation sessions.

It is imperative that a dynamic simulation is 'packaged' in a way that makes it easy to use and learn, yet still be applicable to a broad range of applications and users. The criteria include the following:

- *Easy to use and learn*. It must have an intuitive and interactive graphical environment that involves no writing of code or any compile–link–run cycle.
- *Configurable*. It must provide reuseable modules which can easily be linked together to build the desired model.
- *Accurate*. It must provide meaningful results.
- *Fast*. It must strike a balance between rigour and performance so as not to lose the interactive benefits of simulation.
- *Broadly applicable*. It must provide a broad range of functionality to span different industrial applications, as well as varying levels of detail and rigour.
- *Desktop computer based*. It must reside on a convenient desktop computer environment such as a PC, Mac, or workstation.

With these attributes, dynamic simulation becomes not only available, but also attractive to a much larger audience than ever before. While dynamic simulation is clearly a valuable tool in the hands of seasoned modellers, only when process engineers, control engineers, and plant operating personnel feel comfortable with it will dynamic simulation deliver its most powerful and value-adding benefits.

Even with this emphasis on control system design, chemical plant design used the results of steady-state performance to size the equipment while heuristic methods rather than dynamic systems analysis chose the control schemes. Instruments were field adjusted to give performance as good or better than manual control. When the control schemes, sensing devices, valves and the process itself produced poor results, trial and error was used to find an acceptable level of performance. The lengthy analysis required for an accurate control system design using the equation-based approach could not be justified, or was justified for very few critical loops. Vogel [33] states that, even as late as 1991, only the most challenging and troublesome processes were modelled dynamically with the aim of developing process dynamic behaviour understanding and testing alternative control configurations.

For complex processes that required close control, the weakest link in the control scheme design was usually the dynamic description or model of the process. The response of the sensor, valve, and controller could easily be modelled to within 5 per cent. The modelling error in predicting the dynamic behaviour of the process was generally two to three times greater. The lack of reliable, robust, reusable dynamic process models and suitable software [34] limited the acceptance and use of process control theory.

In summary, the traditional approach to control loop analysis has been through the use of frequency-domain techniques such as Bode diagrams, transfer functions, Nyquist plots, etc. Most of these analysis methods require a working knowledge of Laplace transforms and were developed as pencil-and-paper techniques for solving linear sets of differential equations. Although these frequency-domain techniques are useful for single control loops they are not easily applicable to real multiloop and nonlinear systems which comprise the actual plants that must be controlled in the fluid processing industries.

In the Real-Time (time-domain dynamics) approach, the same set of algebraic and differential equations are encountered as in the frequency domain. However, the major advantage of solving these equations in real time is the ability to observe the interactions of the process, control scheme and load variables much as the operator of a plant observes the behaviour of an actual plant. Dynamic simulation allows for the comparison of several candidate control strategies and assesses the propagation of variation through a process/plant. In other words, dynamic simulation allows for the evaluation of plant-wide versus single-loop control schemes.

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2

Process control hardware fundamentals

In order to analyse a control system, the individual components that make up the system must be understood. Only with this understanding can the workings of a control system be fully comprehended. The rest of this book deals extensively with controller and process characteristics. Therefore, it is appropriate and necessary that hardware fundamentals for the primary elements and final control elements be studied first in this chapter. Discussion of controller hardware is delayed until Chapter 4, where the control equations governing the controllers are covered. Several of the concepts introduced in this chapter are discussed in further detail in later sections of this book.

2.1 Control system components

A control system is comprised of the following components:

1 primary elements (or sensors/transmitters)

2 controllers

3 final control elements (usually control valves)

4 processes.

Figure 2.1 illustrates a level control system and its components. The level in the tank is read by a level sensor device, which transmits the information on to the controller. The controller compares the level reading with the desired level or set point and then computes a corrective action. The controller output adjusts the control valve, referred to as the final control element. The valve opening percentage has been adjusted to correct for any deviations from the set point.

Figure 2.2 is an information flow diagram that corresponds to the physical process flow diagram in Figure 2.1. The information is transmitted between the different control

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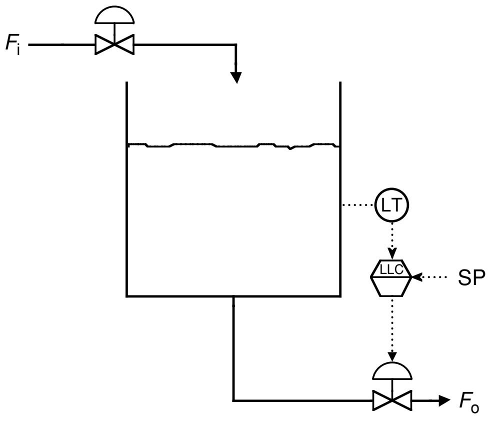


Figure 2.1 Surge tank level controller

system elements as either pneumatic, electronic, or digital signals. These signals often use a live zero. Typical levels are 20–100 kPa (or 3–15 psi) for pneumatic signals, a 4–20 mA current loop that is often converted to 1–5 V for analog electronic signals, and binary digits or bits for digital signals.

2.2 Primary elements

Primary elements, also known as sensors/transmitters, are the instruments used to measure variables in a process such as temperature, pressure, etc. A full listing of the types of primary element available on the market would be very long, but these sensor types

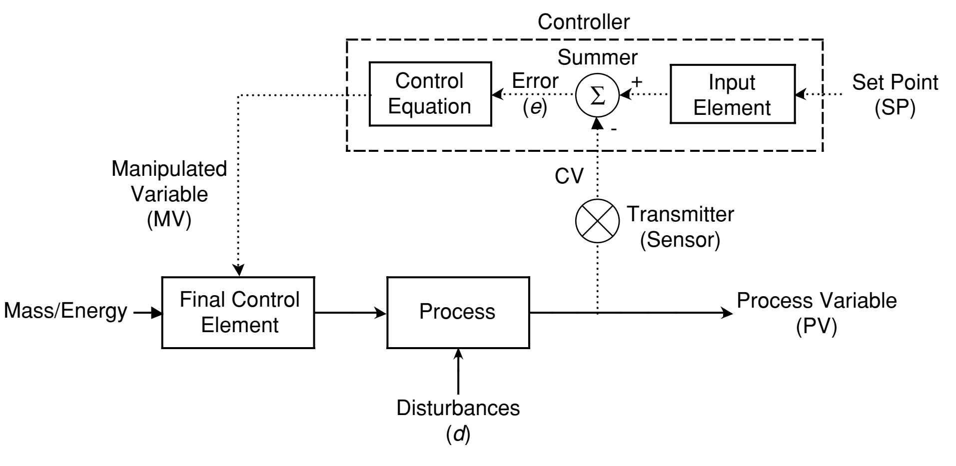


Figure 2.2 Single input–single output block diagram

can be broadly classified into groups including the following:

1 Pressure and level.

2 Temperature.

3 Flow rate and total flow.

4 Quality or analysis instruments, e.g. electrolytic conductivity, pH, pION, moisture, oxidation–reduction potential, gas analysers (O_2, CO_2, H_2) thermal conductivity, gas–liquid chromatography, heat of reaction, calorific value.

5 Transducers (working with the above or as individual units).

Some specific examples of instruments from the more common groups listed above will be examined, including pressure, level, temperature, and flow. It is important to note that this list is not complete or fully representative of the complex developments in this area. Further information and detail can be found in the references.

2.2.1 Pressure measurement

There are numerous types of primary element used for measuring pressure that could be studied; however, this discussion will be limited to some of the most common types encountered. These include manometers, Bourdon tubes, and differential pressure (DP) cells.

Manometers

Manometers are simple, rugged, cheap, and give reliable static measurements. They are, therefore, very popular as calibration devices for pressure measurement. The working concept of a manometer is simple. A fluid with a known density ρ is used to measure the pressure difference between two points, $P_1 - P_2$, based on Equation 2.1, where H is the height difference in the fluid level:

$$P_1 - P_2 = \rho g H \tag{2.1}$$

Figure 2.3 illustrates some of the different manometer types.

The Bourdon tube pressure gauge

The Bourdon tube pressure gauge, named after Eugene Bourdon (circa 1852) and shown in Figure 2.4, is probably the most common pressure gauge used in industry.

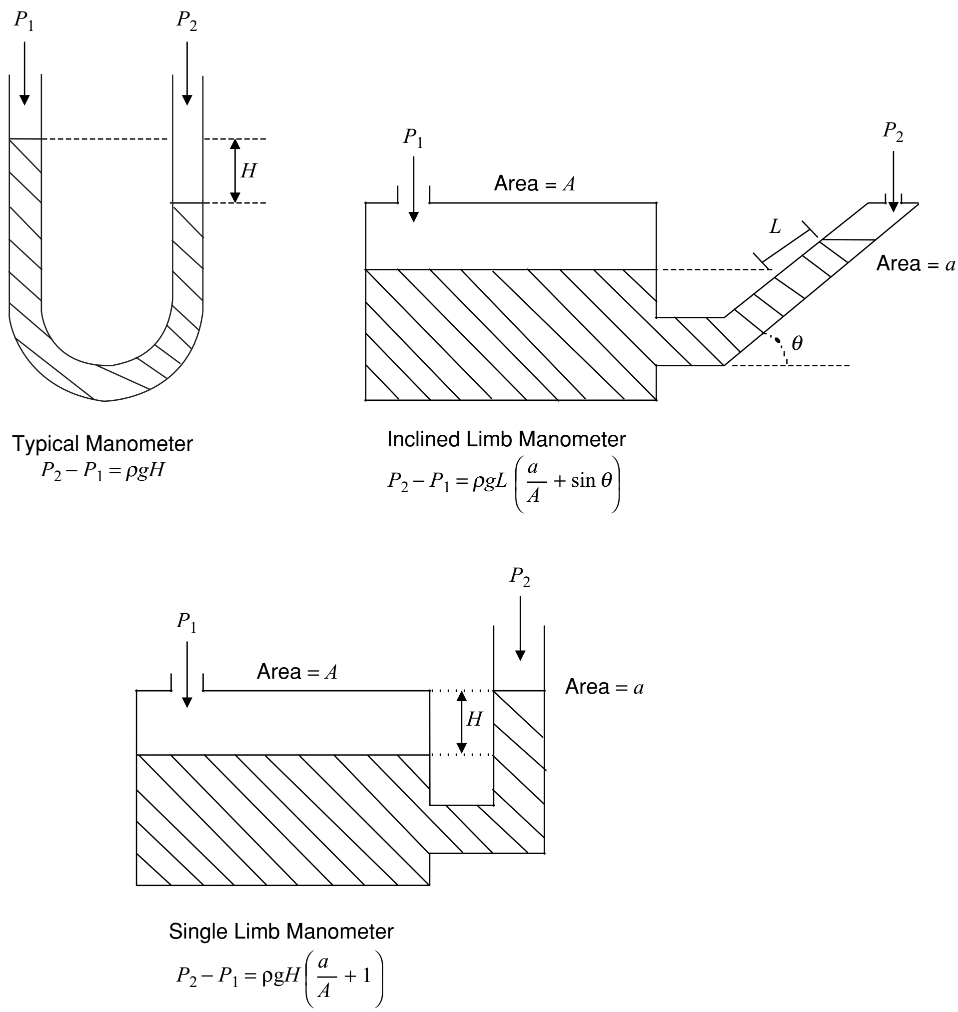


Figure 2.3 Various manometer types

The essential feature of the Bourdon tube is its oval-shaped cross-section. The operating principle behind the gauge is that when pressure is applied to the inside of the tube the tip is moved outward. This pulls up the link and causes the quadrant to move the pinion to which the pointer is attached. The resultant movement is indicated on a dial. A hairspring is also included (not shown) to take up any backlash that exists between quadrant and pinion; this has no effect on calibration.

The accuracy of the gauge is ± 0.5 per cent of full range for commercial models. Generally, the normal working pressure will be specified as 60 per cent of the full scale.

Other types of these gauges include the twist tube, spiral tube, and helical tube. Diaphragm and bellows gauges are two other types of pressure sensor that were developed later. For more details on Bourdon tube materials and design, refer to Giacobbe and Bounds [1], Goitein [2], and Considine [3].

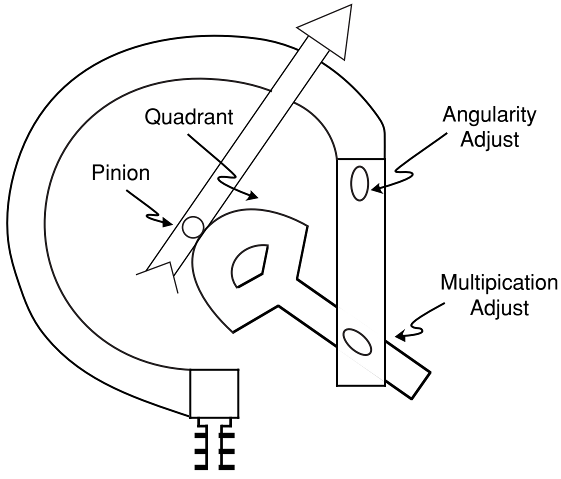


Figure 2.4 Bourdon tube pressure gauge

The differential pressure cell

The DP cell is considered by many as the start of modern-day automation. The DP cell was developed at the outbreak of World War II by Foxboro in Massachusetts, USA, on a government grant provided that it was not patented. The idea was that competition would bring down the price of the instrument.

DP cells allow remote transmission to central control rooms where a small number of operators can control large, complex plants. For example, a typical petroleum refinery processing around 80 000 barrels/day (530 $m^3 h^{-1}$) might have 2000 DP cells throughout the refinery.

Seal systems can be used to enhance the usefulness of the DP cell by facilitating pressure measurement for many temperature ranges (−73°C to 427°C) [4]. They serve to protect the transmitter from the process fluid, using a hydraulic system to conduct the pressure from the process fluid to the transmitter. Only the seal's diaphragm contacts the process fluid and a capillary or tube of fluid transfers the process pressure from the diaphragm to the transmitter. Before a seal is installed, consider ambient conditions, such as temperature, which may introduce errors.

Some of the major benefits of DP cells are that their maintenance is practically zero and no mercury is used in the operation of the transducer.

The pneumatic differential pressure cell Figure 2.5 shows a schematic of a pneumatic DP cell. Pressure is applied to the opposite sides of a silicone-filled twin diaphragm capsule. The pressure difference applies a force at the lower end of the force bar, which is balanced through a simple lever system consisting of the force bar and baffle. This force exerted by the capsule is opposed through the lever system by the feedback bellows. The result is a 3 psi (or 20 kPa if calibrated in SI units) to 15 psi (or 100 kPa if calibrated in SI units) signal proportional to the differential pressure. The

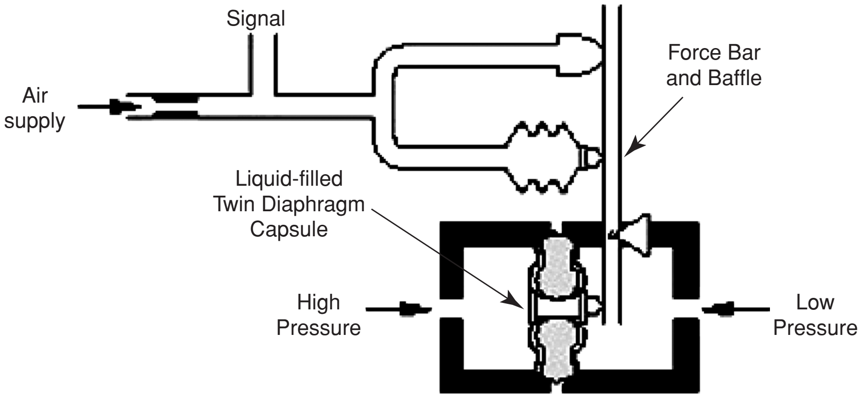


Figure 2.5 Pneumatic DP cell (reproduced by permission of Emerson Process Management)

range of the cell is 1.20 to 210 kPa differential pressure with an accuracy of ±0.5 per cent of the range.

Modern differential pressure cells E-type electronic transducers, strain gauges, capacitive cell transducers, and most recently digital electronics have replaced the pneumatic-type DP cell. Figure 2.6 shows a schematic of an electronic DP cell.

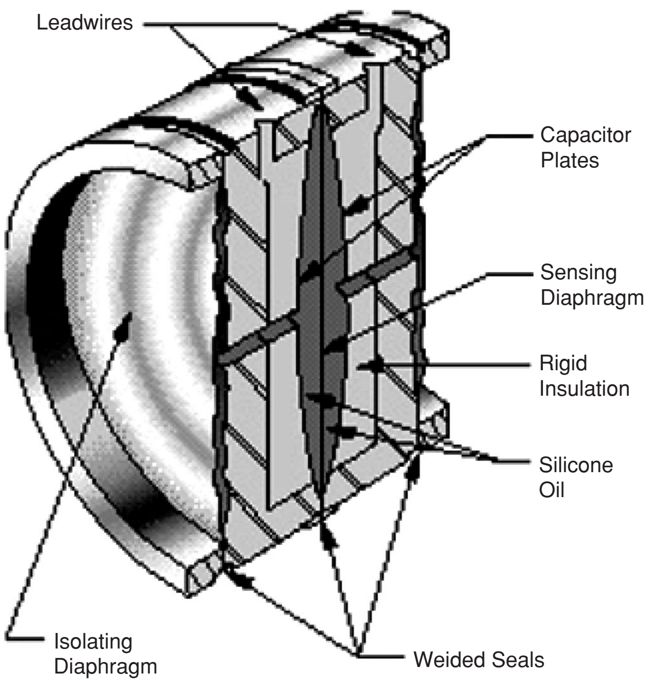


Figure 2.6 Electronic DP cell (reproduced by permission of Emerson Process Management)

Figure 2.7 Model 3051 electronic DP cell (reproduced by permission of Emerson Process Management)

Figure 2.7 is a picture of a modern electronic DP cell, a Model 3051 with Foundation Fieldbus from Rosemount Inc.

The features of the modern electronic DP cell, such as the Rosemount Model 3051 or Honeywell's 'smart' transmitter [5], include remote range change, diagnostics that indicate the location and type of any system faults, easy self-calibration, local digital display, reporting and interrogation functions, and local and remote reporting. The modern DP cell can also be directly connected to a process computer and has the ability to communicate with the computer indicating problem analysis that is then displayed on the computer screen.

2.2.2 Level measurement

Level measurement is the determination of the location of the interface between two fluids which separate by gravity, with respect to a fixed plane. The most common level measurement is between a liquid and a gas. Methods of level measurement include the following [6,7]:

1 Float-actuated devices, such as:
   (a) chain or tape float gauge
   (b) lever and shaft mechanisms
   (c) magnetically coupled devices

2 Pressure/head devices, i.e. DP cells or manometers:
   (a) bubble tube systems
   (b) electrical methods
3 Thermal methods
4 Sonic methods
5 Radar methods
6 Nuclear methods
7 Weight methods.

It is extremely important that vessels are well protected from an overflow condition. An overflowing vessel may have severe safety consequences, impacting nearby employees, the environment and the surrounding community. Some vessels require low-level protection to operate safely. Ideally, each vessel should have a visual indication for the operator, an alarm point and a transmitted level indicator [8].

Factors affecting the choice of level measurement include corrosive process fluids (requiring exotic materials), viscous process fluids which may cause blockages, hazardous atmospheres, sanitary requirements, density changes, dielectric and moisture changes, and the required degree of accuracy and durability.

Pressure/head devices such as the DP cell are the most popular of all level measurement devices. The DP cell can often be used where manometers are impracticable and floats would cause problems. The DP cell requires a constant product density for accurate measurement of level, or a way of compensating for density fluctuations. Figure 2.8 demonstrates a typical set-up for level measurement using a Rosemount Model 3095 level controller, which is essentially a combined DP cell and proportional controller.

Ultrasonic methods

Ultrasonic refers to sound of such high frequency that it is undetectable to the human ear. Frequencies used in level measurement range from 30 kHz to the megahertz range [9].

A transducer sends pulses of ultrasonic sound to the surface of the liquid to be measured. The liquid surface reflects these pulses and the distance from the transducer to the liquid level is calculated. This calculation is based on the speed of the signal and the time elapsed between the sending and receiving of the ultrasonic sound signal (Figure 2.9).

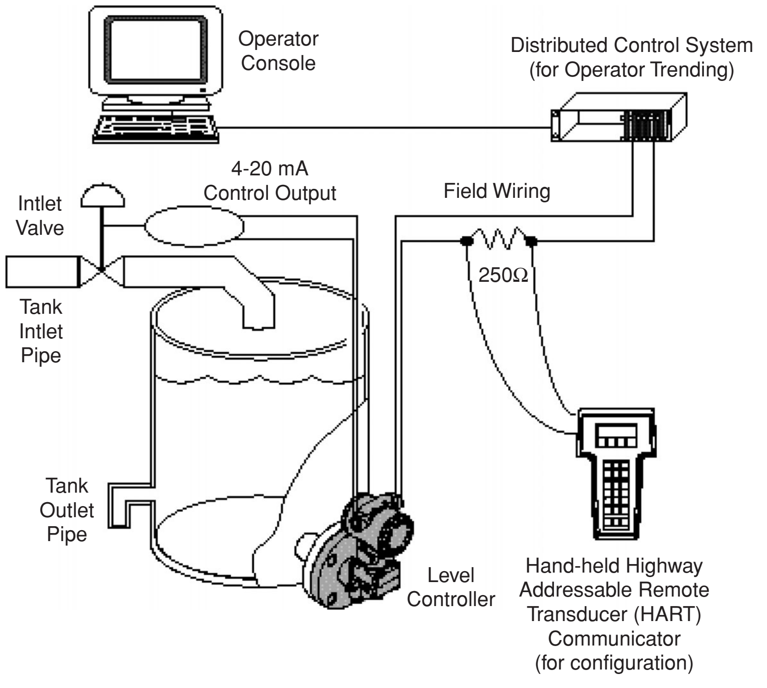


Figure 2.8 Model 3095 multivariable level controller in a liquid level process (reproduced by permission of Emerson Process Management)

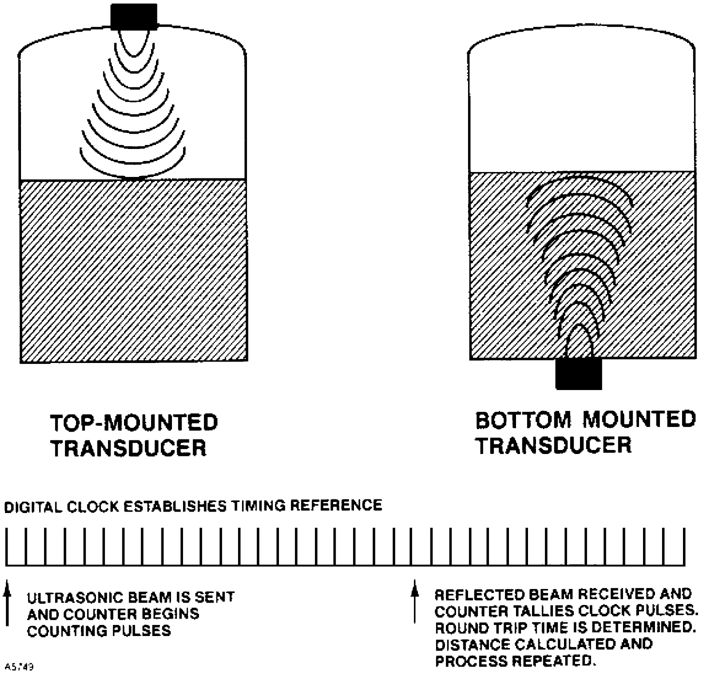


Figure 2.9 Ultrasonic level transmitter [9] (reproduced by permission of Emerson Process Management)

Ultrasonics can be top or bottom mounted. Although a top-mounted device is easier to service, mists, vapours and internal ladders and agitators may cause erroneous readings. Bottom-mounted devices must be calibrated to the density of the fluid measured; however, bubbles and solids in the liquid may skew their reading [9].

2.2.3 Temperature measurement

Methods of measuring temperature include [5]:

1 Change of state

2 Expansion:
   (a) bimetal thermostats
   (b) liquid in glass
   (c) liquid in metal

3 Pressure type:
   (a) gas filled
   (b) vapour pressure filled

4 Electrical:
   (a) resistance
   (b) thermocouple

5 Radiation pyrometers:
   (a) total radiation
   (b) optical

Bimetal thermostats, thermocouples, and resistance thermometers will be discussed in detail.

Bimetal thermostats

The bimetal thermostat works on the concept that different metals expand by different amounts if they are subject to the same temperature rise. If two metals are fixed rigidly together, then a differential expansion takes place when the metals are heated, causing the composite bar to bend. The thermostat employs the bimetal bar to switch on or off a control device depending on the temperature. An illustration of a bimetal thermostat is given in Figure 2.10.

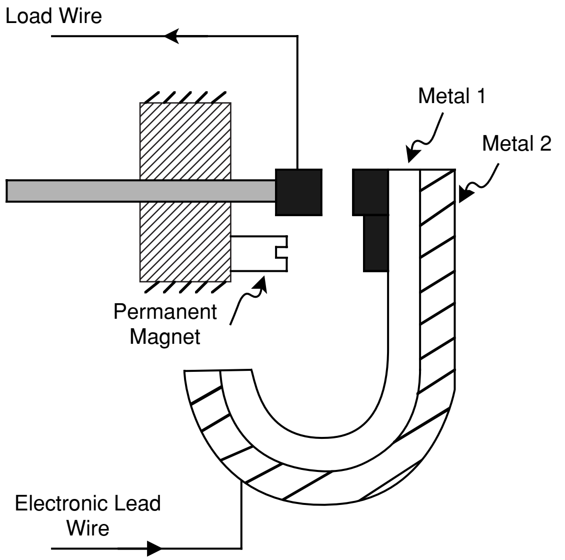


Figure 2.10 Bimetal thermostat

The temperature range for bimetal thermostats is 0 to 400°C with an accuracy of ± 5 per cent, although the accuracy can be increased to ±1 per cent [6]. The deflection–temperature relationship is linear for many metal combinations over a particular temperature range only, and the materials must be chosen with care. These instruments are rugged, cheap, offer direct reading, and can work under conditions of vibration.

Thermocouples

When two dissimilar metal or alloy wires are joined together at both ends to form a loop, and a difference in temperature exists between the ends, a difference in junction potentials exists resulting in a thermoelectric electromagnetic field (e.m.f.). This is known as the Seebeck effect, after Seebeck's 1821 discovery of this phenomenon. The magnitude of the e.m.f. will depend on the types of material used and the temperature difference. This is the concept behind a thermocouple for measuring temperature.

If one junction temperature, the reference or cold junction, is maintained at a constant and known value and the characteristics of the thermocouple are known, then the magnitude of the e.m.f. generated will be a measure of the temperature of the other junction. This other junction is called the hot junction.

The e.m.f. generated for any two particular metals at a given temperature will be the same regardless of the size of the wires, the areas in contact or the method of joining

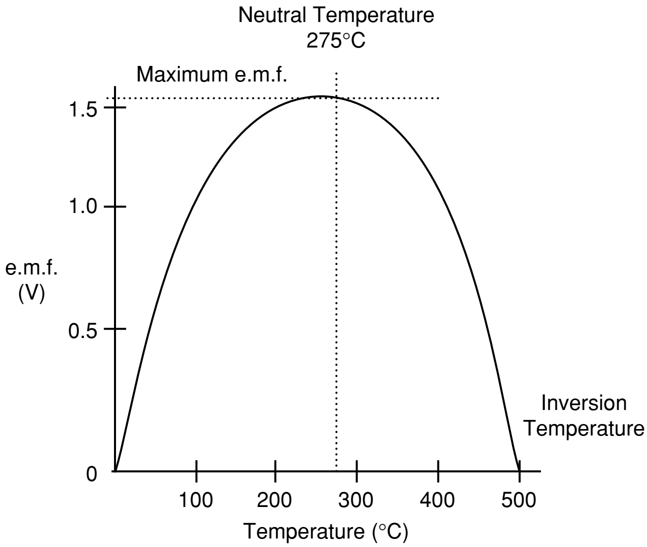


Figure 2.11 Relationship between temperature and e.m.f. for a Cu–Fe system

them together. The relationship between temperature and generated e.m.f. is nonlinear except over limited ranges. On the steep part of the curve, the relationship is

$$e = a(T_1 - T_2) + b\left(T_1^2 - T_2^2\right) \tag{2.2}$$

In Equation 2.2, e is the e.m.f. generated, T_1 and T_2 are the hot and cold junction temperatures in degrees kelvin, and a and b are constants for the given material. An example of the relation is given in Figure 2.11 for a Cu–Fe thermocouple system.

Thermocouple types There are many thermocouple types. Common systems and their ranges are as follows.

Base metal thermocouple types

1 Constantan–copper, type T:−75–93°C (TP) or 93–371°C (TN).

2 Constantan–chromel, type E: 0–316°C (EP) or 316–971°C (EN).

3 Constantan–iron, type J:−73–427°C (JP) or 427–760°C (JN).

4 Alumel–chromel, type K: 0–277°C (KP) or 277–1149°C (KN).

5 Nicrosil–Nisil, type N: 0–277°C (NP) or 277–1149°C (NN).

Noble metal thermocouple types

1 Platinum +10% rhodium–platinum, type S: −18–538°C (SP) or 538–1149°C (SN).

2 Platinum–platinum +13% rhodium, type R: up to 1480°C, depending on sheath materials used.

3 Platinum +5% rhodium–platinum +20% rhodium, type B: up to 1700°C, depending on sheath materials used.

Poisons to thermocouples Selecting the correct thermocouple type is essential:

- Iron deteriorates quickly due to scaling in oxidizing atmospheres at high temperatures.
- Chromel and alumel thermocouples are poisoned by gases that are carbon based, sulphurous, or contain cyanide groups. These thermocouples are better in an oxidizing atmosphere than a reducing atmosphere.
- Platinum must be protected from hydrogen and metallic vapours.

Resistance thermometer detectors

Resistance thermometer detectors (RTDs) are made of either metal or semiconductor materials as resistive elements and may be classed as follows [3]:

1 Wire wound, range −240–260°C, accuracy 0.75%

2 Photo-etched, range −200–300°C, accuracy 0.5%

3 Thermistor beads, range 0–400°C, accuracy 0.5%.

An example is the platinum RTD, which is the most accurate thermometer in the world.

RTDs exhibit a highly linear and stable resistance versus temperature relationship. However, resistance thermometers all suffer from a self-heating effect that must be allowed for, and I^2R must be kept below 20 mW, where I is defined as the electrical current and R is the resistance.

When compared with thermocouples, RTDs have higher accuracy, better linearity, long-term stability, do not require cold-junction compensation or extension lead wires and are less susceptible to noise. However, they have a lower maximum temperature limit and are slower in response time in applications without a thermal well (a protective well filled with conductive material in which the sensor is placed).

Figure 2.12 A selection of thermocouples, RTDs, and accessories (reproduced by permission of Emerson Process Management)

Selecting temperature sensors

Getting the right operating data is crucial in selecting the proper sensor. A good article on selecting the right sensor is by Johnson [10].

Figure 2.12 shows a selection of thermocouples, RTDs, and temperature accessories, such as thermal wells, that are typically available from instrument suppliers (in this case Emerson Process Management). Figure 2.13 shows a picture of a typical temperature sensor and transmitter assembly.

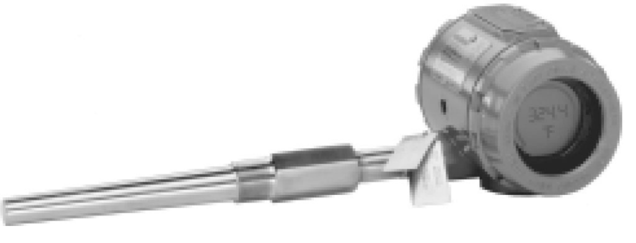


Figure 2.13 Temperature sensor and transmitter assembly (reproduced by permission of Emerson Process Management)

2.2.4 Flow measurement

Flow measurement techniques can be divided up into the following categories [3]:

1 Obstruction-type meters, such as:
   (a) orifice plates
   (b) flow nozzles
   (c) venturi tubes
   (d) pitot tubes
   (e) Dall tubes
   (f) combinations of (a) to (e)
   (g) elbow and target meters
2 Rotational or turbine meters
3 Variable-area meters/rotameters
4 Ultrasonic- and thermal-type meters
5 Square-root extractors for obstruction-type meters
6 Quantity or total flow meters, such as:
   (a) positive displacement
   (b) sliding vane
   (c) bellows type
   (d) nutating disc
   (e) rotating piston
   (f) turbine type
7 Magnetic flowmeters
8 Vortex meters
9 Mass flow meters, such as:
   (a) Coriolis effect flowmeters
   (b) thermal dispersion flowmeters.

Selection of a flowmeter is based on obtaining the optimum measuring accuracy at the minimum price. It should be noted that flowmeters may use up a substantial amount of energy, especially when used in low-pressure vapour service. Therefore, they should only be provided when necessary [8].

There are many factors to consider when selecting a flowmeter, including properties of the fluid being measured (such as viscosity), and performance requirements (such as response time and accuracy). Ambient temperature effects, vibration effects, and ease of maintenance should also be compared when selecting a flowmeter. For a more thorough presentation on the selection of flowmeters, refer to the article by Parker [11].

Orifice plates and magnetic flowmeters will be discussed in detail, since they are two of the most common types found in the fluid processing industry.

Orifice plates

The concentric orifice plate is the least expensive and the simplest of the head meters. The orifice plate is a primary device that constricts the flow of a fluid to produce a differential pressure across the plate. The result is proportional to the square of the flow. Figure 2.14 shows a typical thin-plate orifice meter.

An orifice plate usually produces a larger overall pressure loss than other primary devices. A practical advantage of the orifice plate is that cost does not increase significantly with pipe size. They are used widely in industrial applications where line pressure losses and pumping costs are not critical.

The thin concentric orifice plate can be used with clean homogeneous fluids, which include liquids, vapours, or gases, whose viscosity does not exceed 65 cP (0.065 Pa s)

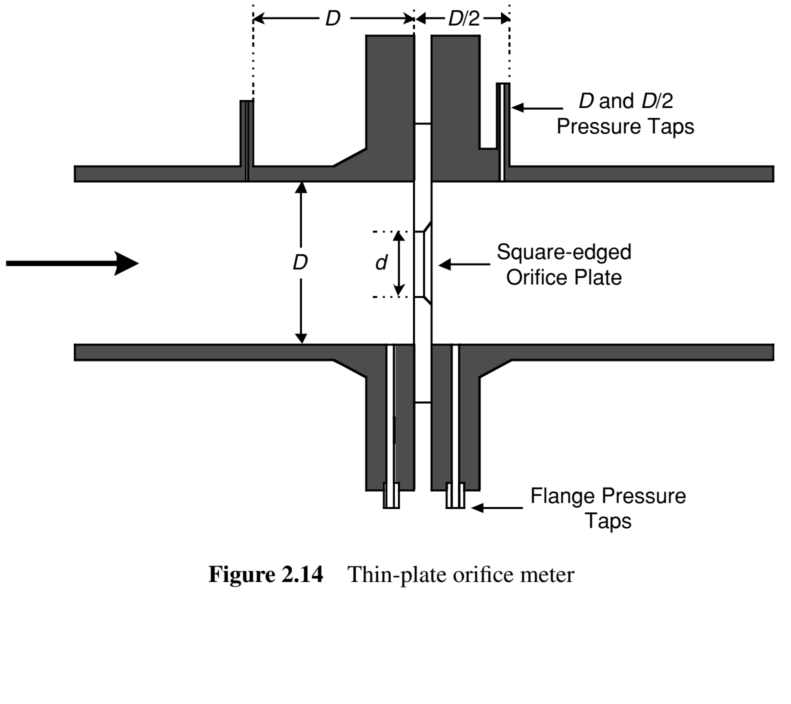


Figure 2.14 Thin-plate orifice meter

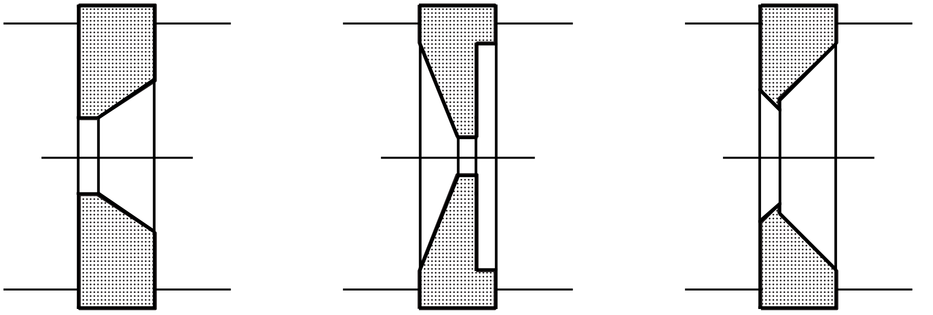


Figure 2.15 Various orifice plate designs

at 15°C. In general, the Reynolds number (Re) should not exceed 10 000. The plate thickness should be 1.5 to 3.0 mm, or, in certain applications, up to 4.5 mm [12].

Many variations for orifice plates have been suggested, especially during the 1950s when oil companies and universities in North America and Europe sponsored numerous PhD studies on orifice plates [3]. Of these only a few have survived, which were the ones that incorporated cheaply some of the features of the more expensive devices. Figure 2.15 shows some of these designs. Other designs that are utilised include eccentric and segmental orifice configurations.

Magnetic flowmeters

The magnetic flowmeter is a device that measures flow using a magnetic field, as implied by the name. The working relationship for magnetic flowmeters is based upon Faraday's law (see Equation 2.3), which states that a voltage will be induced in a conductor moving through a magnetic field.

$$E = kBDV \tag{2.3}$$

Where E is the e.m.f. generated, B is the magnetic field strength, D is the pipe diameter, V is the average velocity of the fluid, and k is a constant of proportionality. As seen in Equation 2.3, when k, B, and D are kept constant, V is proportional to E.

Figure 2.16 illustrates the principle of operation of a magnetic flowmeter.

In the past, magnetic flowmeters have been very expensive compared with orifice plates and DP cells. However, the cost is now very competitive, and in fact magnetic flowmeters are replacing orifice plates where possible.

There are numerous benefits to using a magnetic flowmeter. With polytetrafluoroethylene (PTFE), fibreglass, or rubber liners, the magnetic flowmeter can handle

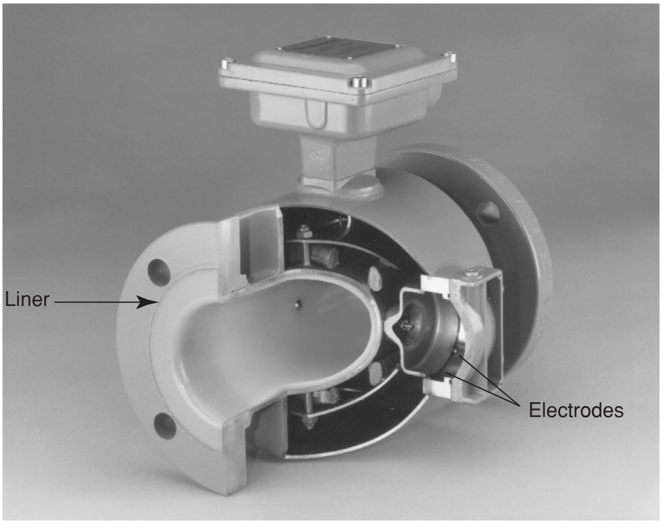


Figure 2.16 Cutaway view of the Model 8705 magnetic flowmeter flowtube (reproduced by permission of Emerson Process Management)

almost any corrosive liquid. The electrodes can be made from very corrosion-resistant metals. Gold, titanium and tantalum have been used in the past. The magnetic flowmeters are virtually maintenance free, and there is no flow obstruction to the stream being measured. Also, they can be readily connected to an electronic controller and they can give out a digital signal that can be fed directly to a computer.

When using a magnetic flowmeter it is necessary that the liquid be conductive, although low conductivities are acceptable. Also, outside capacitance can create a big problem. Calibration should be done carefully initially, with accurate readings done on the liquid conductivity to ensure accurate set up of the magnetic flowmeter.

2.3 Final control elements

Pneumatic, or air-operated, diaphragm control valves are the most common final control element in process control applications. They are used to regulate the flow of material or energy into a system. Variable-speed pumps are also possible, but they are often costly (as motor control is expensive), they are less efficient, break down more often, and maintain maximum pressure if they fail. Electric valves are seen, but only for large applications above 25 cm pipe/valve diameters. Variable electric power control elements, such as rheostats, are used in small applications, such as laboratory water bath temperature control.

Since control valves are the most common final control element we will now devote our discussion to control valves.

2.3.1 Control valves

Since process engineers tend to dedicate their time to tuning control loops, the significance of the performance of the control valve is often overlooked. 'As much as 80 per cent of all process variability can be attributed to poor control valve performance (i.e. how quickly and accurately the control valve responds to the control signal)' [13].

The components that control a valve are [13]:

1 an actuator that serves to open or close the valve;

2 a positioner that works to modulate the flow (limit switches or percentage open);

3 the valve itself, which includes its body (which is a pressure vessel with a passageway through which the process fluid flows), trim and seats:

  (a) the trim is a closure member such as a plug, ball, or gate that modulates the flow of process fluid through the valve;

  (b) the seat is the material (metal or soft polymer) that the closure member contacts to shut off the flow of the process fluid.

The control valve components are illustrated with a ball valve in Figure 2.17.

The sliding-stem control valves are the most common control valve configuration and have at least half the market in control valves. Figure 2.18a and b shows a typical, modern sliding-stem control valve assembly.

The pneumatic diaphragm-operated control valve is the most commonly specified final control element in existence. Pneumatic valves have many advantages over electrically activated valves, but the main ones are initial lower purchase cost, relative ease of maintenance, speed of response, and developed power of the valve plug. This last reason has become less relevant in recent times, since the valve bodies have changed from contoured and ported styles to plug and cage styles in order to avoid unbalanced forces in single-ported designs, especially for high-pressure liquids. Figure 2.19 shows some of the newer styles of valve cage.

Control-valve sizing

The common equation for the flow of a non-compressible fluid through a control valve is given in Equation 2.4 [14], which can be derived from Bernoulli's equation.

$$Q = C_\mathrm{v}\sqrt{\frac{\Delta P}{\mathrm{SG}}} \tag{2.4}$$

In Equation 2.4, Q is the volumetric flow rate, ΔP is the pressure drop across the valve, SG is the relative density compared with water, and C_v is the valve coefficient. C_v is

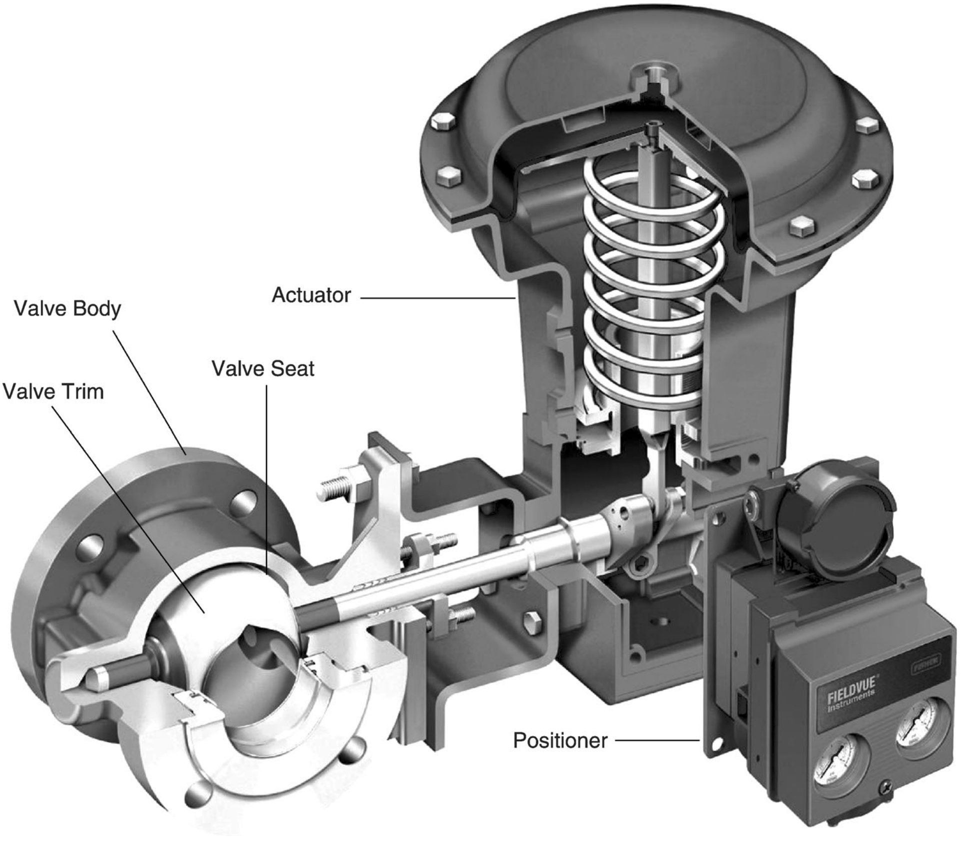
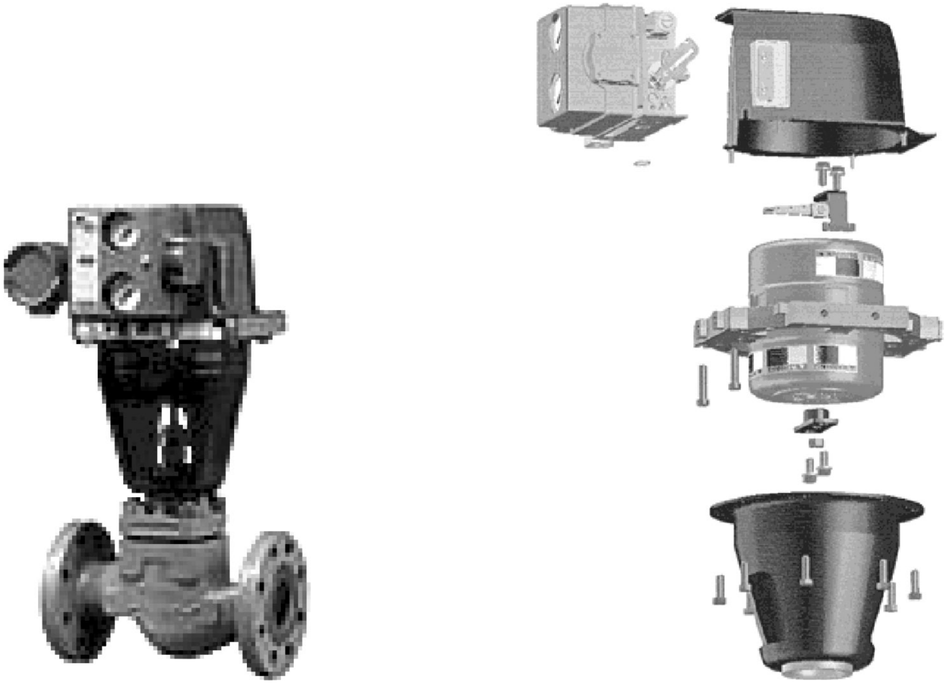


Figure 2.17 Control valve cutaway (reproduced by permission of Emerson Process Management)

defined by convention in field or imperial units as the number of US gallons that will pass through a control valve in 1 min, when the pressure differential across the valve is 1 psi.

C_v varies negligibly with Re for most valve applications. Even in cases where the Reynolds number is low, the Reynolds number at the valve will be high due to the valve restricting flow, and so the valve is normally operating in a region where C_v is independent of the Reynolds number. For valves with a streamlined shape, such as those used for slurries or very viscous liquids, the Reynolds number can be low enough that C_v becomes dependent. In these cases a correction factor for the low C_v is usually supplied in the manufacturers' catalogues under the heading 'Viscosity Correction Factors for C_v' [14–18].

The value of C_v is also a function of A, which is the flow area of the valve. For a given valve, this value of A varies extensively with valve opening. The curve giving the variation of C_v at high Reynolds number with valve opening is called the 'inherent characteristic of the valve'. The maximum value of C_v occurs when the valve is wide open and depends on the design and size of the valve. For geometrically similar valves, C_v is proportional to the valve size.



(a) **(b)**

Figure 2.18 (a) Typical modern sliding-stem control valve assembly; (b) Single-acting spring return actuator and digital valve positioner for a modern sliding-stem control valve (reproduced by permission of Emerson Process Management)

Figure 2.19 Characterised cages for globe-style valve bodies (reproduced by permission of Emerson Process Management)

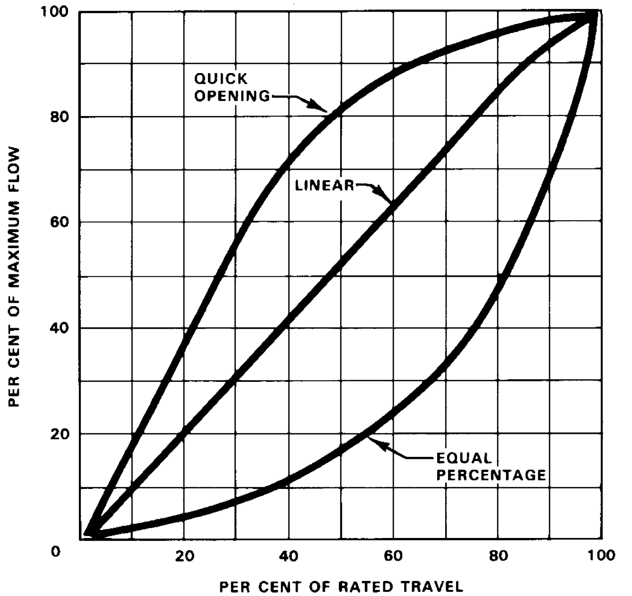


Figure 2.20 Examples of inherent valve characteristic curves (reproduced by permission of Emerson Process Management)

Inherent valve characteristic

The inherent characteristic of a valve is a plot of C_V versus valve opening. This curve is usually plotted as C_V as a percentage of maximum flow (or C_V). Inherent characteristics are usually plotted in this way rather than actual C_V versus actual lift so that the same curve will apply to a set of geometrically similar valves, irrespective of size. If the characteristic curve and the maximum C_V are known, then the C_V at any intermediate lift or opening can be determined.

Three common examples of operating valve characteristics are quick opening, linear, and equal percentage, as illustrated in Figure 2.20.

Operating characteristic

The operating characteristic is a plot of flow versus lift, where lift refers to valve opening, for a particular installation. This is not an inherent property of the valve, and it is usually plotted as flow versus lift in a similar way to the inherent characteristic, with both the flow and the lift being plotted as a percentage of the maximum.

If the pressure drop did not change across the valve with valve opening, then the flow would vary proportionally with C_V, and thus the operating characteristic would be the same as the inherent characteristic curve if both were plotted as a percentage of maximum. However, as the valve closes, the pressure drop across it increases. This increased ΔP is due to the fact that, as the valve resistance increases, the valve's resistance becomes a larger fraction of the total system resistance. This is because, as the valve closes, flow through the system decreases and the system ΔP other than the valve decreases but the valve ΔP increases. This means that, as the valve closes, its C_v falls but its ΔP increases. The result is that the flow does not fall as fast as the C_v, and so the installed or operating flow characteristic differs from the inherent characteristic.

When the valve is shut its resistance is infinite and the whole available pressure drop occurs across it. Thus, the ΔP across the valve varies from a maximum when closed to a minimum when 100 per cent open. The greater this variation, the more the operating characteristic varies from the inherent characteristic. A measure of this deviation is the parameter β, defined by

$$\beta = \frac{\Delta P_v\,(\text{max . flow})}{\Delta P_v\,(\text{min . flow})} \tag{2.5}$$

The smaller the value of β, the greater the deviation of the operating characteristic from the inherent characteristic. For a small β, an operating characteristic like the one shown in Figure 2.21 is obtained.

With such a characteristic as a small β, nearly full flow is obtained when the valve is opened by a small amount. Thus, the effective stem movement range for throttling flow is greatly reduced and erosion of the valve is increased, since the plug is nearly closed at all flows. Therefore, a small β is undesirable and is caused by a valve that is too large. Decreasing the valve size increases the value of ΔP_v (open). This, of course, increases the required pumping costs because of the extra power required to overcome the greater resistance of the valve.

Valve selection based on control performance

From the point of view of control performance, there are two aspects that need to be considered when selecting a valve. These are the valve size and valve inherent characteristic.

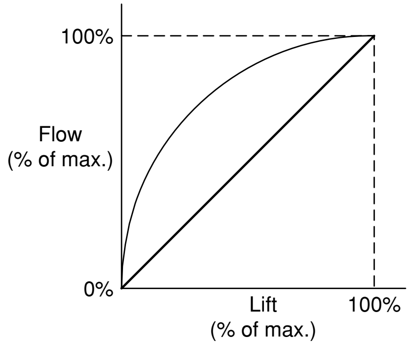


Figure 2.21 Operating characteristic for a small value of β

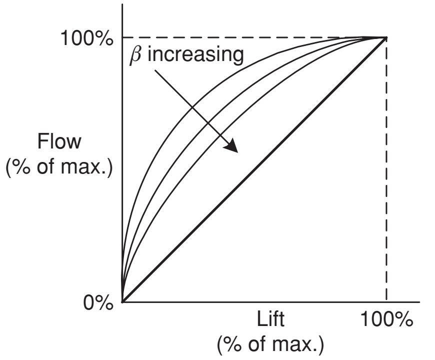


Figure 2.22 Effect of increasing the value of β

As stated previously, these two aspects are not independent since the flow characteristics obtained depend on both. Ideally, the valve size should be decided during the design phase of the plant in conjunction with the choice of pipe and pump size. In this way, it can be ensured that the valve pressure drop is a satisfactory proportion of the total pressure drop, and thus will produce a satisfactory operating characteristic.

It is usually recommended that β be at least 30 per cent. In satisfying this relationship the valve is seldom the same size as the pipe, and usually the valve will be one size smaller, with the minimum recommended size being 50 per cent of the line size. When the control valve is added to an existing system, it is often sized to handle the maximum required flow with the available pressure drop. However, this generally results in an oversised valve, since the pump was not originally designed for losses in the valve. This, in turn, then leads to a poor valve characteristic and unsatisfactory control. For this reason, an equal percentage characteristic should be selected so that the operating characteristic tends towards linear (Figure 2.22). In addition, the equal percentage characteristic is more forgiving when sized incorrectly or if there is insufficient pressure drop across the valve. The equal percentage valve plug has become the standard type of plug when valves are put into an existing system.

If the control system is included as part of the original design, then an equal percentage may not be the best choice, and a linear flow characteristic may also not be suitable. If the process is not linear, then the required open-loop gain to obtain a given degree of stability will vary with the operating point. With a nonlinear operating characteristic the valve gain will also vary with the operating point, and so it may be possible to match the valve's characteristic to that of the rest of the control system to produce a roughly constant system gain at all operating points.

This matching of the valve characteristic to that of the process is only relevant if the operating point of the process does not vary over the whole range, i.e. it is not subject to large disturbances. In this case, there may be no way of matching the valve to the process if there is more than one variable that produces appreciable changes to the operating point. This is because the best characteristic for the compensation of the effects of one load variable may be different from that required for another. However, if there is only one load variable, then it is often possible to determine the best shape of the valve's inherent characteristic by the use of process dynamics.

Valve selection based on process dynamics

Gain is defined as the change in output divided by the change in input. Each component in the control loop has a gain term associated with it. The control valve has a very clearly defined gain term that depends on valve type, size, pressure drop, etc. The process gain term depends upon the process response to a change in input and the various load disturbances imposed upon it. A control system should be designed such that a controller produces an effect equal to the disturbance but 180° out of phase, to bring about cancellation. Thus, for good control the loop gain should be unity, as shown in Equation 2.6. If the gain is less than unity then the disturbance is not fully cancelled, and if the gain is greater than unity then the corrective action is excessive. This concept of gain is explained in greater detail in Chapter 3.

$$K_{\text{controller}} K_{\text{transmitter}} K_{\text{process}} K_{\text{valve}} = 1 \tag{2.6}$$

For a given set of controller settings, the controller gain and the sensor/transmitter gains may be considered as constants, resulting in the relationship of Equation 2.7:

$$K_{\text{process}} K_{\text{valve}} = \text{Constant} \tag{2.7}$$

The gain of the control valve can be computed from the C_{v} versus lift curves as the slope of the tangent to the curve, as shown in Figure 2.23.

If this is done for the common types of control valve over their whole range, then the gain curves shown in Figure 2.24 are obtained.

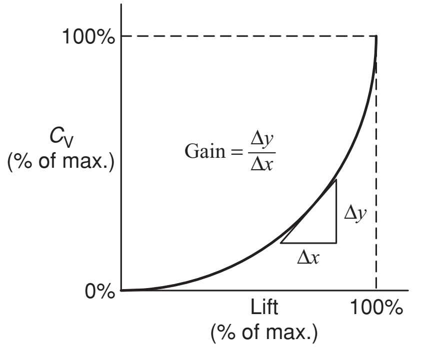


Figure 2.23 Calculation of the control valve gain

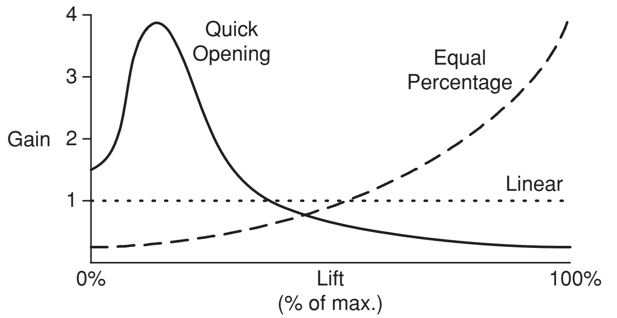


Figure 2.24 Gain curves for common types of control valve

With a quick opening control valve the gain increases to a maximum at about 20 per cent of the valve opening; it then decreases exponentially, resulting in decreasing effectiveness as the valve approaches the fully open position. This indicates that the valve would be good for a process whose gain increases with the variable used to control it.

With a truly linear control valve the gain is a constant, not depending on lift at all. The flow is proportional to the valve position. For example, at 50 per cent open the flow is 50 per cent of the maximum. This characteristic would suit a plant whose gain was independent of the operating point.

With an equal percentage control valve the gain increases exponentially for equal percentage lifts between 10 and 100 per cent. This inherent characteristic is clearly best used for a process whose gain decreases as the load increases.

There are, however, rules of thumb for selecting control valves and matching inherent valve characteristics to common process control loops or processes where the valve pressure drop is fairly constant. The equal percentage characteristic is the most common and is used where variations in pressure drop are expected or in systems where a small percentage of the total system pressure drop is taken across the valve, such as in pressure and flow control. More detailed recommendations are available from control valve vendors, e.g. [15,16].

Quantitatively, β is used to recommend a valve characteristic type. If $\beta > 0.5$, implying relatively less variation in pressure drop as the valve operates, then a linear characteristic is recommended. If $\beta < 0.5$, implying relatively more variation in pressure drop as the valve operates, then an equal percentage is recommended. Ultimately, however, valve gain is used to check the valve selection. For a selected valve, the procedure is as follows.

1 Calculate ΔP across the valve for several flows from the minimum to the maximum flow.

2 Calculate the corresponding C_v values (C_g for gases/vapour, or C_s for steam).

3 Obtain the percentage open data corresponding to the C_v (or C_g, etc.) values calculated in step 2 (vendor data C_v versus percentage open).

4 Calculate the control valve gain as $K_{cv} = \%\Delta\text{Flow}/\%\Delta\text{Opening}$

5 Calculate the average gain.

6 Are the minimum and maximum control valve gains within ± 50 per cent of the average value calculated in step 5? If the answer is yes, then there will be good stable control. If the answer is no, then check other valves' characteristics.

Control valve rangeability

The required range of C_v should fall between 20 and 80 per cent of valve opening. This tends to provide a relatively constant gain and the most stable range of control. This can be checked by doing an analysis of the valve gain as discussed previously and in Purcell [19]. At less than 10 per cent open it can be difficult for the control valve to stabilise and it tends to oscillate, whereas at greater than 80 per cent opening the gain begins to vary too much.

Control valve pressure drop

Control valves control flow by absorbing a pressure drop, which must be specified. This pressure drop is an economic loss to the system, since, typically, it must be supplied by a pump or compressor. As such, economics might dictate sizing a control valve with a low pressure drop, but this results in a larger valve that may have a decreased range of control. Often, the pressure drop to be taken across the control valves is specified when detailed plant hydraulics are not complete and so it needs to be estimated. As such, there are several rules of thumb. Typically, the control valve pressure drop is estimated as 50 per cent of the friction pressure drop taken across the equipment plus piping, or 33 per cent of the total system pressure drop (excluding the valve). Minimum pressure drops have been stated as 10 per cent of the system pressure drop for equal percentage valves, as 25 per cent of the system pressure drop for linear valves [20] or 35 kPa for rotary valves, and 69 kPa for globe valves [19].

As stated previously, the key to sizing control valves properly is to specify the range across which they have to function. Specifying a pressure drop with the above rule of thumb at one condition means that the pressure drop required at other conditions must be checked. The required pressure drops as other conditions are governed by system hydraulics. For example, assume that we have a system pressure drop of 50 kPa. Based on taking 33 per cent of the total system drop at design flow, excluding the valve, a pressure balance results in the control valve taking 25 per cent of 50 kPa, or 12.5 kPa, while the system takes 37.5 kPa. For simplicity, assume there is no elevation component. What happens if the flow increases by 20 per cent above the design value? Hydraulics states that the system pressure drop will increase to 72 kPa. Where is this pressure drop going to come from? The answer is the control valve, but in this case we do not have sufficient pressure drop to supply it and so this system could not have an

Table 2.1 Base (minimum) pressure drops for control valve types

Control valve type	ΔP_b (kPa)
Single plug (globe)	75.8
Double plug (globe)	48.3
Cage	27.6
Butterfly	1.4
Ball	6.9

increase of 20 per cent flow above design. The rule of thumb pressure drop should be at maximum flow and then you should check what happens at minimum flow. A more engineered approach is derived by Connell [21] with the results given in Equation 2.8. In this approach, the pressure drop across a control valve can be estimated:

$$\Delta P_{cv} = 0.05 P_s + 1.1 \left[\left(\frac{Q_m}{Q_d} \right)^2 - 1 \right] \Delta P_f + \Delta P_b \tag{2.8}$$

where ΔP_{cv} (kPa) is the pressure drop across a control valve, P_s (kPa) is the upstream or supply pressure, Q_m is the maximum anticipated flow rate, Q_d is the design flow rate, ΔP_f (kPa) is the friction pressure drop at the design flow rate, and ΔP_b (kPa) is the base (minimum) control valve pressure drop.

The first term accounts for a fall off in overall system pressure drop by using 5 per cent of the system start pressure. The second term accounts for an increase in system flow from design to maximum and the corresponding friction pressure drop. The last term is the base (minimum) control valve pressure drop from [21], given in Table 2.1.

It should be noted that the values for butterfly and ball valves in Table 2.1 appear to be somewhat lower than what others recommend. As such, a minimum pressure drop of 27.6 kPa should be used unless experience indicates a specific low-pressure drop application, e.g. a sulphur plant air control valve.

Practical control-valve sizing

Note that, to calculate the range of size required, the following sizing procedures require the pressure drop at minimum flow and at maximum flow, not just at design flow or an arbitrary multiple thereof. However, conditions at design flow can also be incorporated and be helpful. Note that the procedures use the equations developed by Fisher [15,16], but the sizing procedures are generic. In order to size a control valve properly, the following process information must be known:

- fluid type and viscosity;
- range of controlled flows (minimum and maximum);

- range of inlet and outlet pressures (minimum and maximum pressure drops corresponding to flows);
- specific gravity (SG);
- temperature.

In addition, the following data need to be specified before a valve is purchased:

- shut-off pressure
- leakage rate (ANSI/IEC Leakage Class [14,17])
- noise tolerance (ANSI/IEC Standard [14,17]).

Currently, manufacturers worldwide are implementing the IEC valve sizing code [17]. This is a procedure that allows tighter noise prediction, particularly in gas service.

Liquid control-valve sizing

The basic procedure for sizing, for a given flow rate and pressure drop, is to calculate the required C_v as per a rearrangement of Equation 2.4:

$$C_\mathrm{v} = \frac{Q}{\sqrt{\Delta P/\mathrm{SG}}} \tag{2.9}$$

This calculation should be performed at minimum and maximum conditions to obtain $C_{\mathrm{v,min}}$ and $C_{\mathrm{v,max}}$. Subsequently, these required values are compared with a C_v range for a particular valve. Typically, the required C_v values should fall in the range of about 20 to 80 per cent of the valve opening.

A plot of Equation 2.4 (Figure 2.25) implies that the relationship of flow is linear with respect to the square root of pressure drop, with the slope equal to C_v and that the flow can be continually increased with pressure drop.

Actually, a limit is reached where this is no longer true, known as choked flow. As liquid passes through a reduced cross-sectional area, the velocity increases and the pressure decreases. The point of minimum pressure and maximum velocity is the *vena contracta*. As the fluid exits, the velocity is restored but the pressure is only partially restored, creating a pressure drop as shown in Figure 2.26. Note that pressure recovery is how much pressure is restored from the vena contracta. As such, all else being equal, a high-recovery valve has a low pressure drop and vice versa.

If the minimum pressure falls below the vapour pressure of the liquid, it partially vaporises. This is what causes cavitation and flashing, discussed in more detail later. At the vena contracta, as the pressure decreases, the density of the vapour phase, and mixture, decreases. Eventually, this decrease in density offsets any increase in the velocity of the flow so that no additional mass flow is realised, according to the continuity equation.

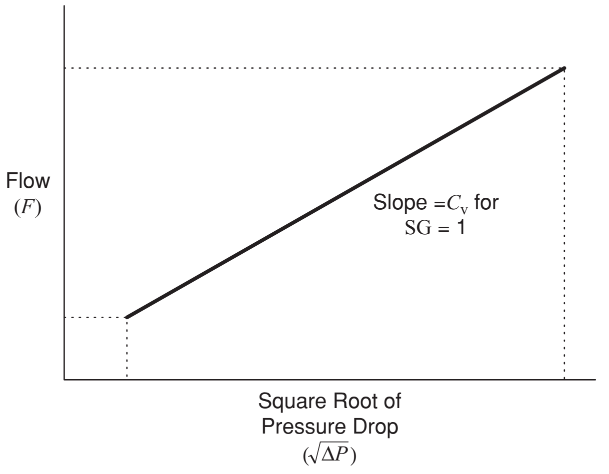


Figure 2.25 Flow versus square root of pressure drop across a control valve

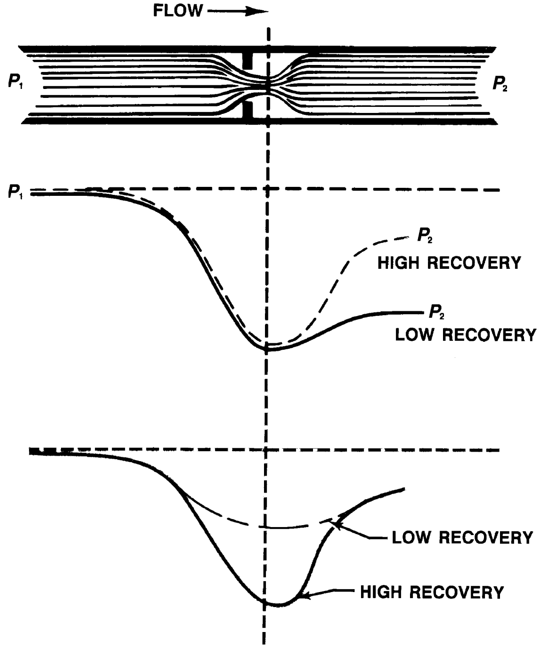


Figure 2.26 Pressure profiles across a control valve (reproduced by permission of Emerson Process Management)

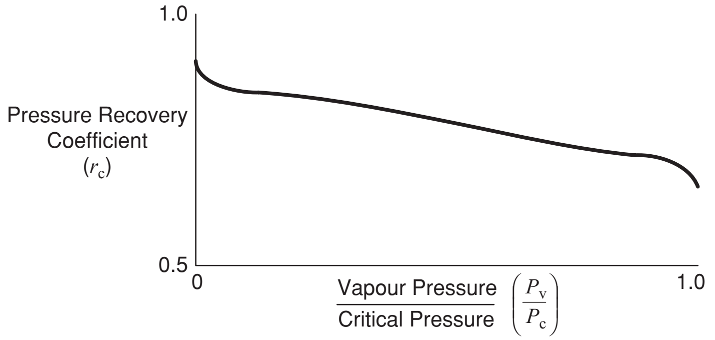


Figure 2.27 r_c versus P_v/P_c

The issue becomes how choked flow is integrated into liquid valve sizing. Fisher Controls [15,16] defines a pressure recovery coefficient as follows:

$$K_m = \frac{P_1 - P_2}{P_1 - P_{vc}} \tag{2.10}$$

where P_{vc} is the pressure at the vena contracta.

Experimentally, it has been found that $P_{vc} = r_c P_v$, where P_v is the vapour pressure. Typically, r_c is obtained from a graph, like Figure 2.27.

Note that water has a similar but different curve for obtaining r_c. Equation 2.10 can be rearranged as

$$\Delta P_{allow} = (P_1 - P_2)_{allow} = K_m(P_1 - r_c P_v) \tag{2.11}$$

In Equation 2.11, K_m is constant for a particular valve. Other control valve vendors also have expressions for ΔP_{allow} that are functions of vapour and critical pressures. Overall, this results in the following liquid sizing procedure:

1 Obtain conditions at minimum and maximum (flow, pressure drop) and properties SG, P_c and P_v.

2 Initially assume a valve size (one size smaller than piping) and valve type to obtain the pressure recovery coefficient K_m, or equivalent.

3 Calculate $C_{v,min}$ and $C_{v,max}$ using the smaller of ΔP_{actual} and ΔP_{allow} in Equation 2.9 at both conditions.

4 If C_v falls in the range of 20–80 per cent open, then the valve is adequate, otherwise a larger valve is required. Note that the design C_v should be at about 50–60 per cent open.

The minimum flow often corresponds to maximum pressure drop and vice versa and these actual pressure drops are determined by system hydraulics (as discussed earlier). Also note that there are C_v corrections for viscous flow for cases where the valve Reynolds numbers are less than 5000. Finally, note that the valve characteristic type and rangeability should be checked, as detailed earlier. Although sizing equations in this sections use Fisher nomenclature, they are equivalent to ANSI/ISA versions if F_L^2 (where L signifies liquid) is substituted for K_m.

Gas, vapour and steam control-valve sizing

The major differences between liquid service and gas, vapour or steam service are:

- that the fluid is compressible;
- the phenomenon of critical flow.

When the ratio of $\Delta P/P_1$ exceeds 0.02, the gas is undergoing compression. Critical flow occurs when the flow is not a function of the square root of the pressure drop across the valve, but only of the upstream pressure. This phenomenon occurs when the fluid reaches sonic velocity at the vena contracta. Since gas cannot travel faster than sonic velocity, critical flow is a flow-limiting condition for gas. It has been found that critical flow occurs at different $\Delta P/P_1$ ratios, depending on whether the valve is high or low recovery.

Fisher, as well as other vendors, has equations for gas, vapour and steam flow which have two parameters. One parameter represents flow capacity and the other parameter represents the type of valve and its effect on critical flow. The Fisher equations are as follows:

$$C_g = \frac{Q_{\mathrm{scfh}}}{\sqrt{\frac{520}{\mathrm{SG}T}}\, P_1 \, \sin\left(\frac{C_a}{C_1}\sqrt{\frac{\Delta P}{P_1}}\right)} \qquad \text{(ideal gas)} \tag{2.12}$$

$$C_g = \frac{W_{\mathrm{lb/hr}}}{1.06\sqrt{\rho_1 P_1}\, \sin\left(\frac{C_a}{C_1}\sqrt{\frac{\Delta P}{P_1}}\right)} \qquad \text{(non-ideal gas)} \tag{2.13}$$

$$C_s = \frac{W_{\mathrm{lb/hr}}(1 + 0.000\,65 T_{\mathrm{SH}})}{P_1 \, \sin\left(\frac{C_a}{C_1}\sqrt{\frac{\Delta P}{P_1}}\right)} \qquad \text{(steam)} \tag{2.14}$$

Where Q_{scfh} is the gas flow rate, $W_{\mathrm{lb/hr}}$ is the mass flow rate, P_1 is the upsteam pressure and T and T_{SH} are temperature.

The constant C_a, in the denominator is 59.64 if the sine evaluation is in radians and is 3417 if the sine evaluation is in degrees and $C_1 = C_g/C_v$.

For these equations, when $\Delta P/P_1 \leq 0.02$ and $\sin(x) \approx x$, the C_g equation reduces to a 'gas version' of the basic equation for liquids because the pressure drop is far below the critical value and the compressibility is negligible.

$$C_v = \frac{Q_{\text{scfh}}}{\sqrt{\frac{520}{\text{SG}T}} P_1 C_a \sqrt{\frac{\Delta P}{P_1}}} \tag{2.15}$$

At the critical pressure drop, $\sin(x) \approx 1$ and C_g is only a function of P_1:

$$C_g = \frac{Q_{\text{scfh}}}{\sqrt{\frac{520}{\text{SG}T}} P_1} \tag{2.16}$$

Overall, this results in the following gas, vapour, or steam sizing procedure:

1 Obtain conditions at minimum and maximum (flow, pressure drop) and ρ_1, if necessary.

2 Initially assume a valve size (one size smaller than piping) and valve type to obtain C_1.

3 Calculate $C_{g,\min}$ and $C_{g,\max}$ or $C_{s,\min}$ and $C_{s,\max}$.

4 If C_g or C_s fall in the range of 20–80 per cent open, then the valve is adequate, otherwise a larger valve is required. Note that the design C_g or C_s should be at about 50 per cent to 60 per cent open.

The ANSI/ISA valve sizing equation for gas is

$$C_v = \frac{Q}{NP_1 Y \sqrt{\frac{\Delta P/P_1}{GTZ}}} \tag{2.17}$$

In Equation 2.17, Y is an expansion factor (ratio of flow coefficient for a gas to that for a liquid) that plays a similar role to C_1, G is specific gravity and Z is a compressibility factor. Although the form of this equation seems much different than the Fisher one, the results are equivalent. Again, the valve characteristic type and rangeability should be checked.

Cavitation and flashing

As stated previously, if the pressure in the vena contracta falls below the vapour pressure of a liquid, then it will partially vaporise. If the pressure recovers above the vapour pressure, then the gas bubbles collapse on the metal and tend to break it away in small

Figure 2.28 Typical appearance of cavitation damage (reproduced by permission of Emerson Process Management)

pieces. This is known as cavitation (Figure 2.28). Because the pressure drop across the valve varies as the opening varies, cavitation may not occur across the entire range of valve opening. If the pressure does not recover above the vapor pressure, then flashing occurs, which can erode the valve plug and seat (Figure 2.29).

Fisher uses a similar equation to the Equation 2.18 to describe cavitation pressure drop:

$$(P_1 - P_2)_{cav} = \Delta P_{cav} = K_c (P_1 - r_c P_v) \tag{2.18}$$

The values for K_c are constant for a particular type of valve. Fisher has related K_c to K_m, with a few examples given in Table 2.2. Other control valve vendors use a similar equation.

Figure 2.29 Typical appearance of flashing damage (reproduced by permission of Emerson Process Management)

Table 2.2 K_c/K_m values for some valve types

Valve type	K_c/K_m
Globe valve (cavitation control trim)	1.00
Globe valve (standard trim)	0.85
Ball valve	0.67
Butterfly valve	0.50

Control valves can be designed to prevent cavitation, or at least to minimise the damage from flashing. For example, cavitation control-type valve trims use the concept of reducing the pressure in several small increments through several stages, instead of one larger pressure drop in a single stage. This avoids the pressure in the vena contracta dropping below the liquid vapour pressure. Flashing is determined by the system, not the control valve, because the outlet pressure is below the vapour pressure of the liquid. However, flashing damage can be minimised by using specially designed valve trims.

Valve positioners

Valve positioners are used to assist positioning control valves under difficult service applications where the control valve may otherwise be out of balance. Their operation employs the negative feedback principle. The position of the valve stem is balanced via cam and beam with the signal from the controller. The out-of-balance motion is detected by a nozzle, which increases the air to the top of the valve via a relay until equilibrium is obtained. Figure 2.30 illustrates the function of a positioner for a diaphragm actuator, and Figure 2.31 shows a modern control valve utilising a digital valve positioner.

Valve positioners should be used when any of the following conditions apply:

- single-ported valves with high pressure drops that require large stem thrusts;
- viscous liquids, sludges, and slurries;
- large distances between the controller and control valve;
- three-way control valves;
- unusually tight packing required because of corrosive fluids, low emissions, or high temperatures;
- large valves that use high volumes of control air;
- split range operation, which is when two or more valves are operated by one controller.

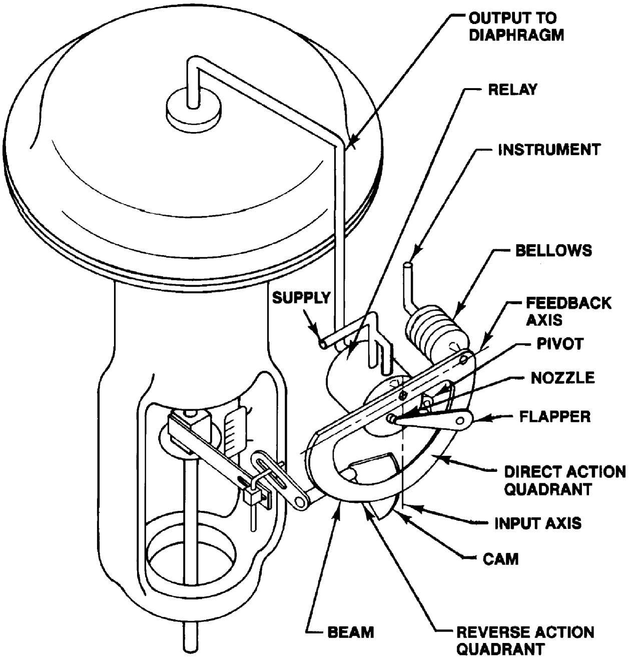


Figure 2.30 Pneumatic positioner schematic for diaphragm actuator (reproduced by permission of Emerson Process Management)

Figure 2.31 Modern control valve and digital valve positioner (reproduced by permission of Emerson Process Management)

In addition, with increasing emphasis on economic performance, valve manufacturers currently recommend that positioners be considered for valve applications where process variability performance is important [22].

Fail-safe design

Fail-safe design means that if a plant has to close down because of instrument power supply failure, of either air or electricity, then the process is designed to shut down safely. This ensures safety for the environment, people, product, and equipment. Fuel-gas valves to fired heaters would fail closed; cooling water valves generally fail open. Several control-valve designs are available that allow this purpose. The way a valve is classified is by the manner it closes under the action of the spring. There is fail safe open and also fail safe close type designs. Figure 2.32 illustrates these fail-safe designs.

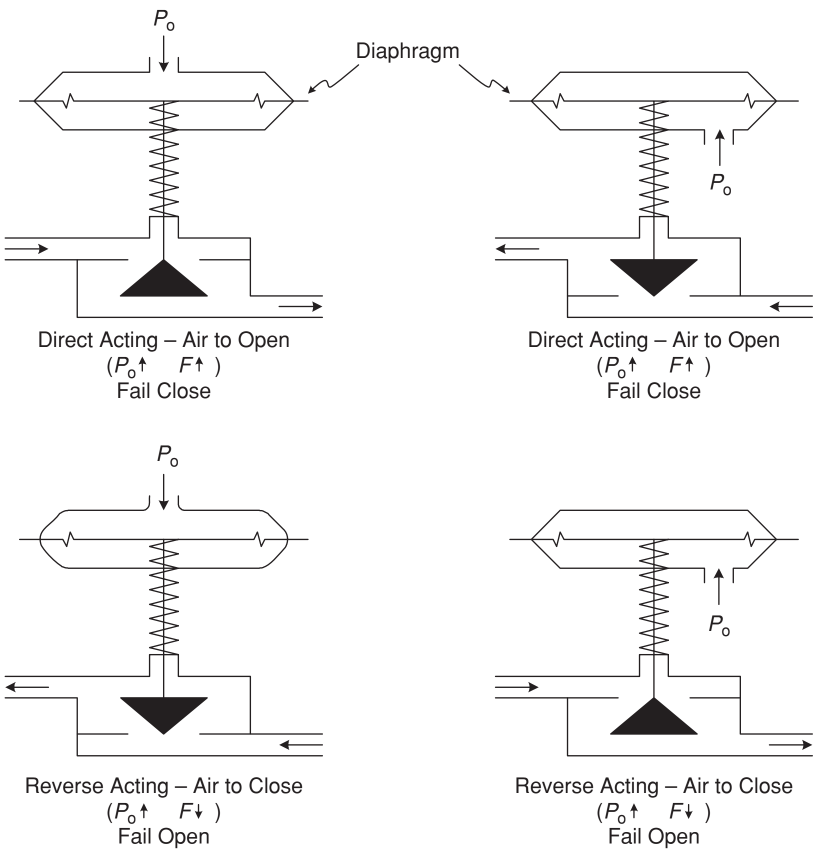


Figure 2.32 Fail-safe design of valves

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3

Fundamentals of single input–single output systems

In this chapter we describe the basic components and concepts of single input–single output (SISO) control systems, along with some of the physical attributes commonly found in these systems. We will also explain the characterisation of system responses and provide an introduction to modelling various processes. After studying this chapter, the reader should understand:

- the basic components of an SISO process control loop;
- the difference between open-loop and closed-loop control;
- the concept of direct-acting and reverse-acting controls;
- what process capacitance is and what it contributes to process controllability;
- what process dead time is and what challenges it poses to process controllability;
- how to develop some of the basic equations that govern first-order system response with feedback control.

We recommend that the student review the fundamentals of differential equations and some of the more common numerical methods to aid in understanding the mathematical development and solution of the various process models presented in this chapter. Some excellent sources for such a review are included in the references [1–3].

3.1 Open-loop control

Most readers will be familiar with how the speed of an automobile is controlled. The basic 'process' set-up is quite standard, as illustrated in Figure 3.1. There is an air–fuel mixture feed, a throttle that regulates how much feed is introduced, and there is the

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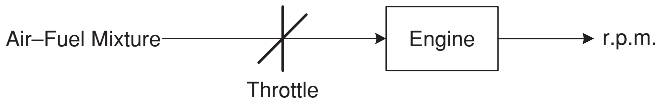

Figure 3.1 Illustration of car metaphor

engine itself that converts combustion energy into rotating mechanical energy that turns the wheels at a certain *r.p.m*. Consider a car on a straight, flat road on a still day. Move the throttle to just the right position and you will achieve the desired speed. Once set, there is no need to adjust it. After all, if nothing changes in the environment, the r.p.m of the engine should stay right where it is. This is a familiar example of *open-loop control.*

Figure 3.2 illustrates a more generic process than our automobile example, but the basic elements are the same. Here, instead of an air–fuel mixture feed, we use the term *mass and energy* feed. Instead of a throttle, we call it a *final control element* (FCE). Instead of an engine, we have a *process.* And instead of an output r.p.m., we call the measured output of the process the *controlled variable/process variable.*

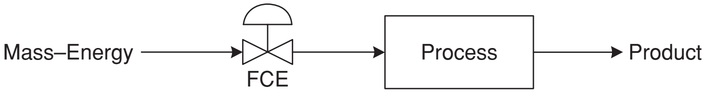

Figure 3.2 A simplified process view

By definition, *open-loop control* places the FCE in a fixed position, or a prescribed series of positions, with the expectation that nothing will change (i.e. there will be no disturbances) to cause the desired state of the system (set-point value) to drift. Other examples of open-loop control are traffic lights and a washing-machine cycle. Once the control action is initiated, it will proceed through the prescribed steps, or remain fixed without any knowledge of the actual status of the process. Sometimes this actually works. Much of the time, however, it does not. Consider our automobile example if the road suddenly rises steeply or a strong headwind is encountered. A more realistic view of a process or plant is shown in Figure 3.3.

3.2 Disturbances

The process shown in Figure 3.3 adds the more realistic dimension of upsets or disturbances d. Upsets and disturbances typically come in three types: input disturbances, load disturbances, and set-point disturbances. An *input disturbance* is a change in the mass or energy of the supply, or input, to the process that may cause the condition of the process variable to drift from its set-point value SP. A *load disturbance* is any other upset, except for an input mass or energy change, which may alter the quality of the

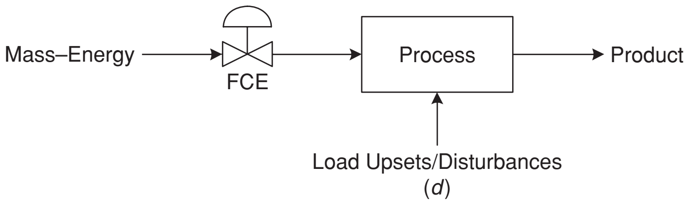

Figure 3.3 A realistic process view

process variable from the desired set-point value. A *set-point disturbance* occurs when the desired state of the process variable (PV) changes, and the process must adjust to a new state. The biggest difference between input disturbances and load disturbances (and the reason we distinguish between them) is that load disturbances cannot typically be anticipated, and they are often not measured directly. The only way we find out about them is by observing the effect they have on the product conditions or quality. While input disturbances may also be difficult to anticipate, they are often measured, and corrective action can more readily be taken. For this reason, we will focus primarily on load disturbances for the remainder of this chapter.

Returning to our automobile example and adding the more realistic dimension of disturbances, we see that, in order to ensure we keep a steady r.p.m., we need to be able to adjust the throttle position constantly in order to keep a constant speed. This is essentially the function that cruise control carries out, and is an example of automatic feedback control. Simply put, *automatic feedback control* provides an automatic adjustment to the FCE in an attempt to maintain the conditions of the process variable at the desired set-point value SP in the presence of disturbances d.

3.3 Feedback control overview

The simplest and most widely used method of process control is the feedback control loop shown in Figure 3.4. Note that we take a measurement of the process variable (indicated by CV, or controlled variable, in Figure 3.4) we care about ('Transmitter' in Figure 3.4 means 'measuring device') and this value is compared with a set-point (SP) to create an error, or departure from aim. In Figure 3.4, OP indicates the operating point around which this calculation takes place. This error is used to drive the corrective action of the FCE via the controller. Note that the output of the controller 'manipulates' the mass and energy into the system via the FCE. Thus, the property that the controller manipulates is referred to as the *manipulated value* (MV). The action of the controller may be aggressive or sluggish; it depends on the internal equations of the controller (sometimes called the control algorithm or control law) and the tuning that is used. We will discuss controller types in Chapter 4 and tuning in Chapter 5. In order to control a process successfully, it is important to select both the right process variable and the

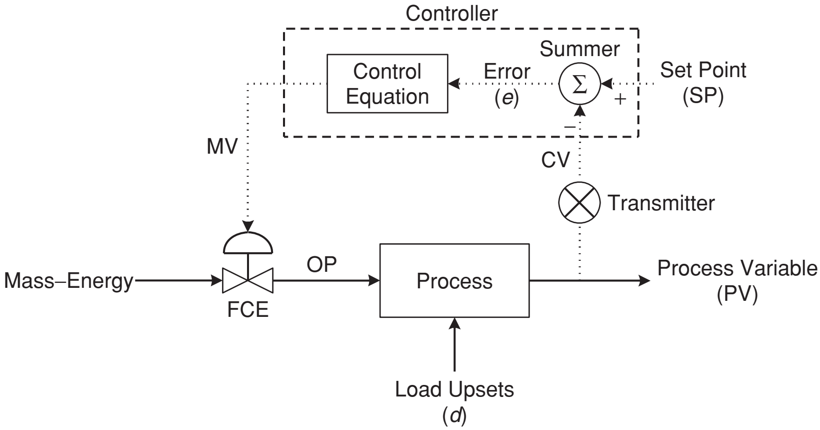

Figure 3.4 SISO feedback control loop

right manipulated value. The process variable is typically a temperature, pressure, flow, composition or level, and is typically a variable that (a) is important to the product quality or the process operation and (b) is responsive to changes in the selected manipulated value.

It is interesting to note that if an automatic feedback controller succeeds in keeping the PV at the desired SP in the presence of load disturbances then, by necessity, there will be changes in the MV dictated by the controller. So in effect, *process control takes variability from one place, and moves it to another*. Thus, the trick to process control is understanding where variability can be tolerated and where it cannot, and designing schemes that manage variability to acceptable levels.

The heat exchanger, shown in Figure 3.5 [4], illustrates this transference of variability. The temperature of the 'Hot Feed to Downstream Unit' stream is important to control (its temperature is the PV for this controller). The 'Hot Utility' stream flow is manipulated (it is the MV for this controller) in order to keep the PV at its set-point in the presence of load disturbances introduced in the 'Feed from Upstream Unit' stream. With no control, these disturbances would make their way into the 'Hot Feed to Downstream Unit' stream. With control, the MV absorbs these disturbances while keeping the PV at or near the set-point. Thus, in effect, this controller transfers disturbances to the MV that would otherwise pass to the PV. As the control algorithm and/or tuning changes, so too does the amount of variability transferred.

Let us summarize our discussion so far:

- Open-loop control suffices when no disturbances are present to threaten the desired state of the product.

- 'Real' processes must operate in the presence of disturbances; therefore, they require some sort of control. Automatic feedback control is the most common form of control.

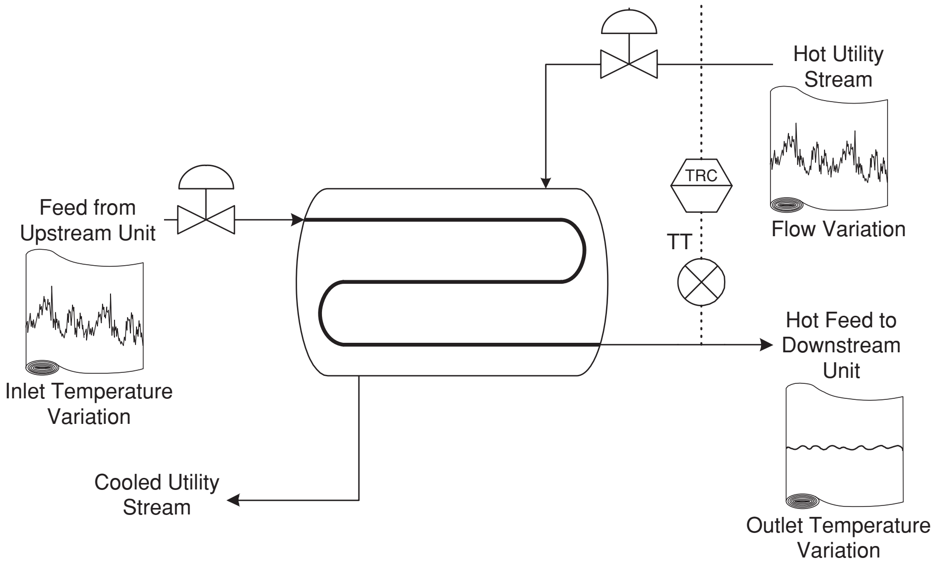

Figure 3.5 Transformation of variation from temperature to flow (courtesy of CACHE Corporation) [4]

- The basic elements of a feedback controller are:

  1 the process variable PV, which represents the variable that is important to maintain under control;

  2 the set-point SP, which represents the desired value of the PV;

  3 the error e, which is a measure of the difference between the PV and the SP;

  4 the controller, whose 'control law' and tuning drive the corrective action and influence the response of the SISO system;

  5 the FCE (typically a valve) to which the controller output is attached and through which the controller exercises its influence on the PV;

  6 the manipulated variable MV, which represents the variable in the process to which the PV is sensitive, and to which the FCE is attached.

- A feedback controller works by measuring the PV and comparing it with the SP to generate an error. The error, conditioned by the controller type and tuning, drives appropriate changes in the FCE (and thus the MV) such that the PV is driven back in the direction of the SP.

Having provided an overview for the need for and basic operation of feedback control, we will now take a closer look at how such control loops are configured.

3.4 Feedback control: a closer look

Mathematically, the error drives the action of the controller. The sign of this error is an important consideration, and requires more development than one might expect. Let us begin with the notion of positive and negative feedback.

3.4.1 Positive and negative feedback

Positive feedback represents a controller contribution that reinforces the error; therefore, it precludes stability. Consider the audio feedback that occurs when a microphone is placed too close to the speaker that amplifies the microphone's output. Sound from the microphone is amplified through the speaker. If this sound re-enters the microphone, it adds to itself, and so on until the speaker saturates with a deafening tone. This is an example of positive feedback. Since positive feedback has no useful purpose for automatic control, we will consider it no further.

Negative feedback represents a controller contribution that diminishes the error; therefore, it tends to add to stability. The cruise control in our automobile example works with negative feedback. If the speed is too high, then the controller cuts back on the flow of the air–fuel mixture, thereby reducing the error. The opposite happens when the speed is too low.

As you can see, only negative feedback presents a viable control loop capable of maintaining stability. However, there are many different elements in a typical control loop, each one with a potentially reinforcing or subtracting contribution. Thus, we need to understand the 'action' of each component in the loop in order to determine whether, in the aggregate, the control loop will provide negative feedback. By action, we typically mean the sign relationship between an element's input and output. The next section will explain this.

3.4.2 Control elements

Let us first look at the action of the process element of the controller. Consider a furnace that heats your home in the winter. When the energy that drives the furnace increases, the temperature in the surrounding rooms increases as well. This is known as an increase/increase (I/I) relationship, or a direct-acting element [5]. *Direct action* refers to a control-loop element that, for an increase in its input, also experiences an increase in its output.

Now consider an air conditioner that cools your home in the summer. When the energy that drives the air conditioner increases, the temperature in the surrounding

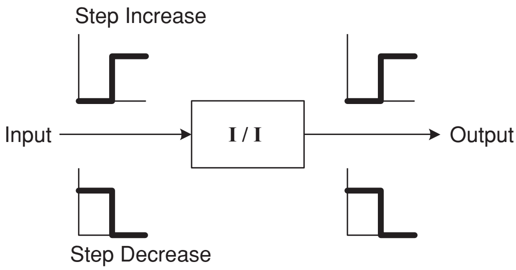

Figure 3.6 Increase/increase component action

rooms decreases. This is known as an increase/decrease (I/D) relationship, or a reverse-acting element. *Reverse action* refers to a control-loop element that, for an increase in its input, experiences a decrease in its output [5].

Consider a general component with I/I action as shown in Figure 3.6. Ignoring the relative amplitudes between input and output, if there is an increasing or decreasing input then there will be a corresponding increasing or decreasing output.

Now consider a general I/D component, as shown in Figure 3.7. For this element, if there is an increasing or decreasing signal, then a resulting decreasing or increasing output signal will result.

Connecting several I/I components in series, as shown in Figure 3.8, will result in an overall I/I action.

As seen in Figure 3.9, if a single I/D component is placed anywhere in the sequence then the overall action is I/D.

Figure 3.10 shows that when two I/D blocks are in series then there is an overall I/I action. It can further be shown that whenever there are an *even* number of I/D blocks in a series the overall effect is I/I, and whenever there is an *odd* number of I/D blocks in a series the overall action is I/D.

Every component in the typical loop (shown in Figure 3.11) including the sensor/transmitter, the controller, the FCE, and the process is either direct or reverse acting. Recall that only negative feedback presents a viable control loop capable of maintaining

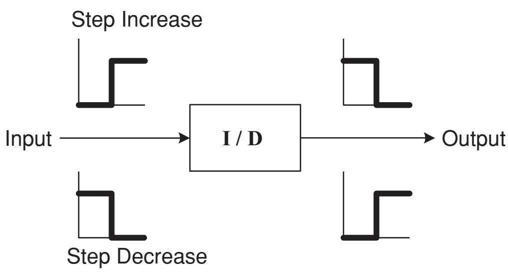

Figure 3.7 Increase/decrease component action

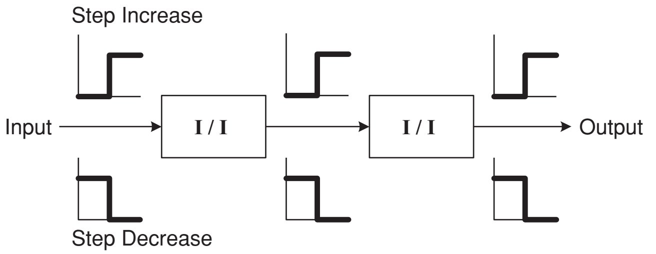

Figure 3.8 Increase/increase components in series

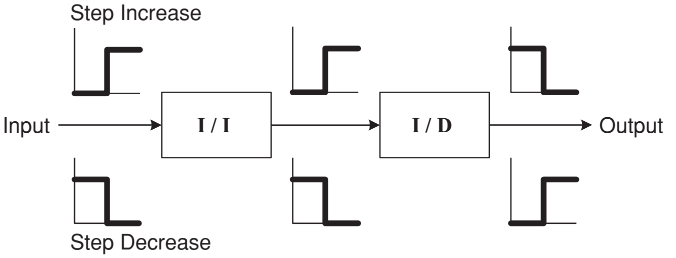

Figure 3.9 Combined components in series

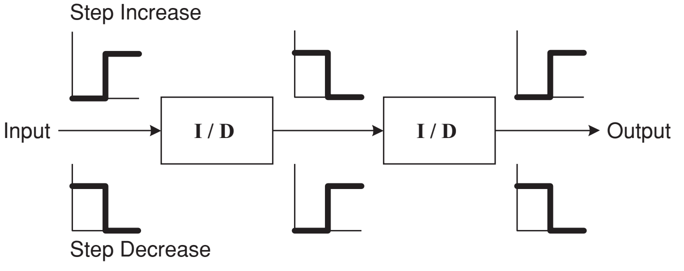

Figure 3.10 Increase/decrease components in series

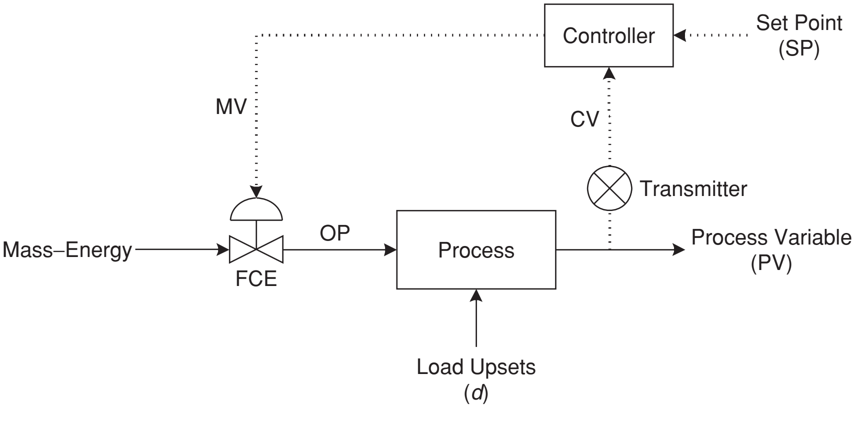

Figure 3.11 Typical SISO loop

stability. Thus, in the aggregate, the overall action of the control loop must be I/D or reverse acting. D/D or direct action generates, by definition, positive feedback.

Since the overall action of the control loop is determined by the action of each of the individual components, let us take a look at each of the typical control-loop elements in turn.

3.4.3 Sensor/transmitter

For the majority of applications, the sensor/transmitter produces an increasing output for an increasing input; therefore, the sensor/transmitter is typically direct acting. There are some special cases where a sensor/transmitter may be reverse acting. However, this is generally not the case, and even if it were, as will be shown later, this poses no problem.

3.4.4 Processes

Most processes are direct acting; however, they can also be reverse acting. Let us examine several major types of process and determine the relationships between the signs of their inputs and outputs.

The first process is a single tank shown in Figure 3.12. For this process an increase in the input F_i causes an increase in the level h for a fixed valve position. Hence, this is a direct-acting process.

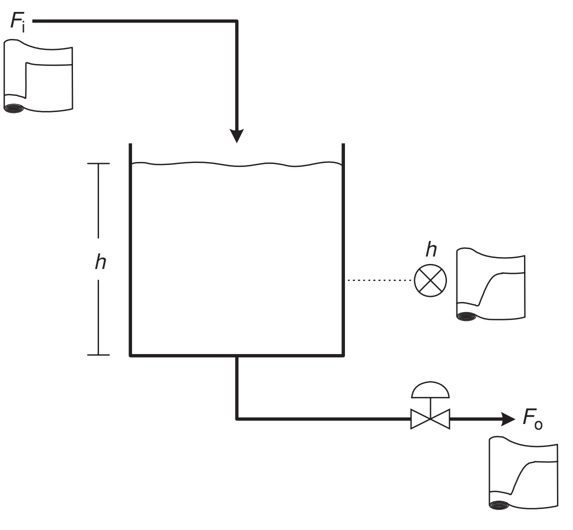

Figure 3.12 Mass flow process

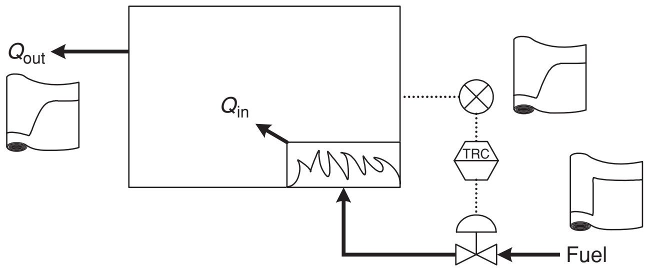

Figure 3.13 Energy flow process

Now consider the energy flow or heating process illustrated in Figure 3.13. In this case, increasing the fuel flow results in an increase in temperature. Hence, this is also an I/I process, or direct acting.

Finally, consider the case of the reactor shown in Figure 3.14. We assume that the feed and catalyst are mixed and the resulting chemical reaction generates heat; in other words, the reaction is exothermic. The rising temperature from this generated heat is the process output, and the cold water flow to the reactor jacket is the process input. The result is a reverse-acting process, since an increase in cold water flow will result in a decrease in reaction temperature.

3.4.5 Final control element

An FCE can be almost anything that controls the flow of mass or energy into or out of a process. It may be a motor speed control on the fan blades of an air-cooled heat exchanger, a star valve on a bin containing solids, etc. However, in the fluid processing industries about 90 per cent of all final control elements are valves. Hence, it is necessary to understand the action of a control valve. For a manual control valve, as the stem

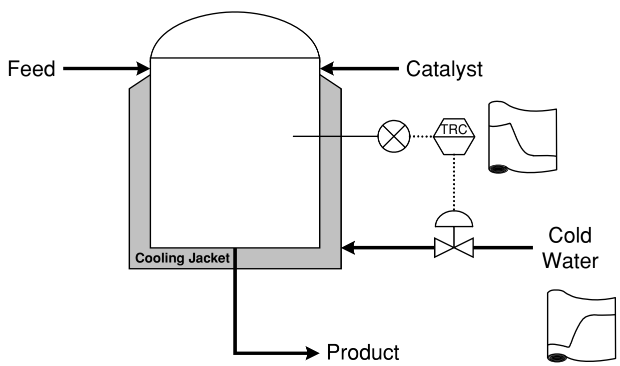

Figure 3.14 Exothermic reactor

position of the valve is moved upwards or open, the flow through the valve also increases, resulting in direct action. Most valve actuators in process control applications are pneumatically activated. In the case of an air-to-open actuator (fail closed), an increasing air signal causes the actuator to stroke open and, therefore, the flow through the valve increases. This is direct acting. For an air-to-close actuator (fail open), an increasing air signal to the actuator closes the valve and flow decreases, resulting in a reverse-acting valve.

Therefore, in the case of the FCE being a valve, it may have either direct or reverse action depending on the actuator chosen for the valve. The desired action is chosen so that fail-safe operation is achieved. For fail-safe operation, the engineer must consider whether a 'fail open' valve or 'fail closed' valve would provide the best safety in the event of a failure. A 'fail closed' valve simply means that if the energy supply to the valve was to fail then the valve would close, allowing no flow. Conversely, a 'fail open' valve opens when the energy supply fails. Cooling water to a reactor is best by an air-to-close valve. Loss of instrument air would fail the valve in the open position (because it takes air to close it), ensuring that there is sufficient cooling and preventing damage to the reactor. Conversely, the valve controlling the steam flow to a reboiler should be an air-to-open valve. Loss of instrument air here would fail the valve closed (since it takes air to open it), ensuring that the column will not overheat during the failure. See the 'Fail-safe design' section in Chapter 2 for an illustration of air-to-open and air-to-close conventions.

3.4.6 Controller

All controllers, whether implemented as stand alone or as part of a distributed control system (DCS) application, have a switch which will allow either direct or reverse action. In general, the action of the controller is the last to be specified, since there is typically little choice in the action of the other elements in the loop. Once the other elements' actions are known, the controller action may be set such that the overall loop action is reverse acting, or I/D.

For the components shown in Figure 3.15, assume the action shown in Figure 3.16, and also assume an air to open actuator (I/I) for the valve. Note that the valve, process, and sensor/transmitter are all direct acting. Therefore, in order to get the desired negative feedback action (I/D overall loop action) the controller must be set to reverse action.

Next, consider the situation for an air to close (I/D) actuator, shown in Figure 3.17. In this situation, the controller must be set to direct action in order to achieve the overall negative feedback required for the loop.

3.5 Process attributes: capacitance and dead time

As we will see later in this chapter and elsewhere, the equations that govern the dynamics of some of the unit operations in typical process plants can be quite complex. Despite

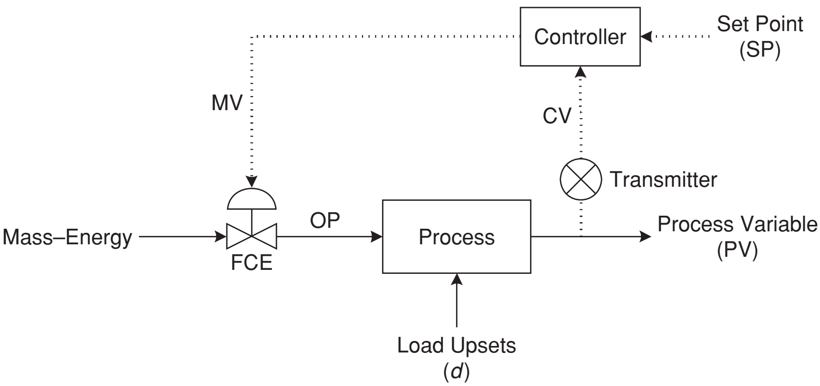

Figure 3.15 SISO feedback control loop

this complexity, many processes behave as if they were first-order systems, many exhibiting transport delay or dead time. Because of this, it is important to understand two fundamental process dynamic characteristics: capacitance and dead time.

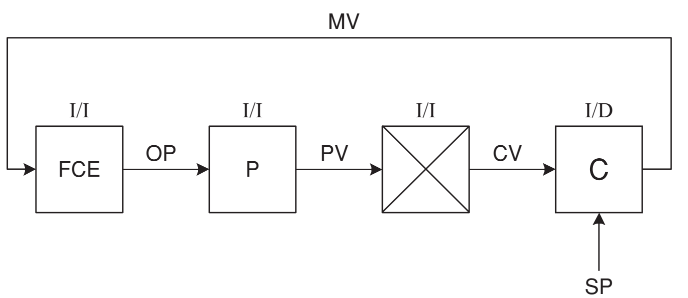

Figure 3.16 Component input/output for air-to-open actuator

3.5.1 Capacitance

Simply stated, *capacitance* represents a system's ability to absorb or store mass or energy. Capacitance may also be defined as the resistance of a system to the change of mass or energy stored in it, i.e. inertia. A common example of a capacity-dominant process is one that stores energy (Figure 3.18).

In this example the process consists of an oven which is storing heat to maintain a particular temperature T. The gas flow creates a flow of energy in Q_{in}. Q_{out} is the flow of energy to the ambient or, in other words, the heat loss to the ambient. For an increase or decrease in the valve position changing the gas flow in, the temperature would

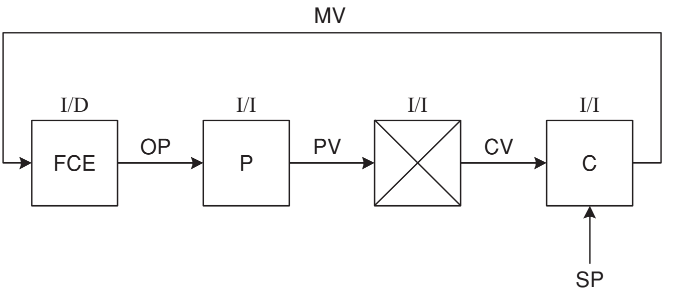

Figure 3.17 Component input/output for air-to-close actuator

correspondingly increase or decrease. It would not, however, change instantaneously with a change in valve position. This behaviour is due to the system's capacitance.

Consider the classical capacity-dominant system shown in Figure 3.19: the surge tank. In this example the tank has a volume in which a mass of liquid is stored. Consider what would happen to the level H in the tank, if the inflow F_{i} were increased. One would certainly expect the level to rise. However, if F_{i} was increased by 10 per cent, the level would not increase to a steady-state value instantaneously. It would eventually reach a higher level, but the capacitance of the tank limits the rate of change in level; thus, it takes some time to reach a new steady-state level. In other words, the tank has inertia and self-regulation. *Self-regulation* occurs when a process, in this case tank level, eventually lines out to a steady-state value for each input step change, rather than ramping off indefinitely.

The rate of change of volume in the tank can be written as a lumped parameter model, where all the resistance to flow is assumed to be associated with the valve, and all the capacitance of the process is assumed to be associated with the tank. This model is shown in Equations 3.1 and 3.2. The basis of Equation 3.1 is the principle of conservation, mass balance in this case (i.e. what goes in must come out or get accumulated in the system).

$$\text{In} - \text{Out} = \text{Accumulation} \tag{3.1}$$

$$\rho_{\text{in}}\, Q_{\text{in}} - \rho_{\text{out}} Q_{\text{out}} = \frac{\mathrm{d}\,(\rho_{\text{out}} V)}{\mathrm{d}t} = \frac{\mathrm{d}\,(\rho_{\text{out}} A H)}{\mathrm{d}t} \tag{3.2}$$

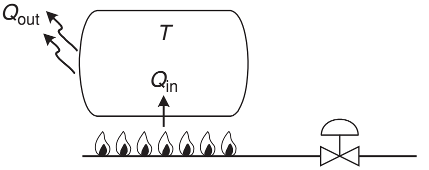

Figure 3.18 Capacity-dominated process: energy storage

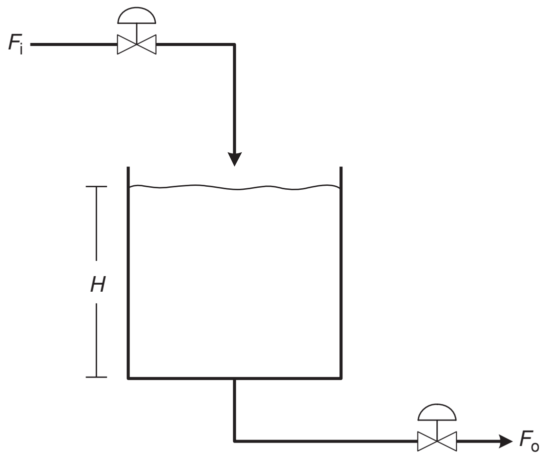

Figure 3.19 Capacity-dominated process: surge tank

In Equation 3.2, Q is the volumetric flow rate of water, V is the volume of the tank, A is the cross-sectional area of the tank, ρ is the density, and H is the water level. Assuming the density and cross-sectional area are constant results in

$$Q_{\text{in}} - Q_{\text{out}} = \frac{A\,\mathrm{d}H}{\mathrm{d}t} \tag{3.3}$$

The flow Q_{out} is determined by the valve characteristic $V(X_{\text{p}})$, with X_{p} being the valve plug-stem expressed in per cent opening, the valve flow constant C_{v}, and the square root of the pressure drop across the valve as given in

$$Q_{\text{out}} = V\left(X_{\text{p}}\right) C_{\text{v}}\sqrt{\Delta p} = V\left(X_{\text{p}}\right) C_{\text{v}}\sqrt{\rho g H} \tag{3.4}$$

In Equation 3.4, V is a function of X_{p}, where V is the fraction of the total volumetric flow rate. Refer to Figure 2.21 for an illustration of how V varies with X_{p} for different types of valve. The symbol g is the acceleration due to gravity. Substituting for Q_{out} into Equation 3.3, which is a first-order differential equation, would result in a nonlinear first-order differential which, unfortunately, has no analytical solution. The response of head (level) H to changes in Q_{in} or valve position could only be determined by numerical methods. However, if Q_{out} is linearized using a Taylor series expansion about a desired operating level then the first-order differential equation can be solved analytically for a disturbance in flow into the tank. For a single variable the Taylor series can be written as

$$F(H) = F(H_{\text{o}}) + \left(\frac{\partial F}{\partial H}\right)_{H_{\text{o}}} (H - H_{\text{o}}) + \text{HOT} \tag{3.5}$$

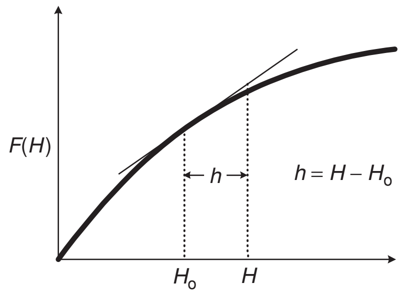

Figure 3.20 First-order linearisation

In performing a first-order linearisation, shown in Figure 3.20, the higher order terms (HOT) are neglected since $h = (H - H_o)$ is small. Setting $F(H)$ (Equation 3.6) as a function of head H and substituting into Equation 3.5 results in the linear form shown in Equation 3.10.

Let

$$F(H) = V(X_p)C_v\sqrt{\rho g}\sqrt{H} = K_v\sqrt{H} \tag{3.6}$$

$$F(H) = F(H_o) + \left(\frac{\partial F}{\partial H}\right)_{H_o}(H - H_o) \tag{3.7}$$

$$\left(\frac{\partial F}{\partial H}\right)_{H_o} = \frac{1}{2}\left(\frac{K_v}{\sqrt{H_o}}\right) \tag{3.8}$$

Or:

$$F(H) = F(H_o) + \left(\frac{K_v}{2\sqrt{H_o}}\right)(H - H_o) \tag{3.9}$$

Or:

$$F(H) - F(H_o) = \left(\frac{K_v}{2\sqrt{H_o}}\right)(H - H_o) \tag{3.10}$$

Let us now complete the derivation of the linear differential equation that describes the response (also called time behaviour or personality) of the level to feed flow disturbances, starting with Equation 3.11.

$$A\frac{\mathrm{d}H}{\mathrm{d}t} = Q_{in} - F(H) \tag{3.11}$$

At the initial or steady-state condition, Equation 3.11 can be written as

$$A\frac{\mathrm{d}H_o}{\mathrm{d}t} = Q_{\mathrm{in}_o} - F(H_o) \tag{3.12}$$

Subtracting the above two equations results in

$$A\frac{\mathrm{d}(H_o - H)}{\mathrm{d}t} = Q_{\mathrm{in}_o} - Q_{\mathrm{in}} - [F(H_o) - F(H)] \tag{3.13}$$

Equation 3.13 can be rewritten in terms of deviation or variation variable, $h = (H_o - H)$ and $q_{\mathrm{in}} = (Q_{\mathrm{in}_o} - Q_{\mathrm{in}})$, as is shown in Equation 3.14 and in a slightly more simplified form as Equation 3.15:

$$A\frac{\mathrm{d}h}{\mathrm{d}t} = q_{\mathrm{in}} - \frac{K_v}{2\sqrt{H_o}}h \tag{3.14}$$

$$A\frac{\mathrm{d}h}{\mathrm{d}t} = q_{\mathrm{in}} - \frac{h}{R} \tag{3.15}$$

In Equation 3.15, R (min m^{-2}) is the resistance to flow. Equation 3.15 can be written in the standard form for a first-order linear differential equation (LDE):

$$RA\frac{\mathrm{d}h}{\mathrm{d}t} + h = Rq_{\mathrm{in}} \tag{3.16}$$

where $RA = \tau$, the time constant (units of time).

Using the classical mathematical approach to solutions of a first-order LDE, one can proceed as follows:

$$\frac{\mathrm{d}C}{\mathrm{d}t} + P(t)\,C \;=\; Q(t) \tag{3.17}$$

Equation 3.17 has a solution of the form shown in Equation 3.18 [6]:

$$C = \mathrm{e}^{-\int P(t)\,\mathrm{d}t} \int Q(t)\,\mathrm{e}^{\int P(t)\,\mathrm{d}t}\;\mathrm{d}t \;+\; C_1\;\mathrm{e}^{-\int P(t)\,\mathrm{d}t} \tag{3.18}$$

C_1 is a constant of integration evaluated from the initial conditions. Writing the LDE for the tank level in general form gives

$$\frac{\mathrm{d}h}{\mathrm{d}t} + \frac{h}{\tau} = Kq_{\mathrm{in}} \tag{3.19}$$

where $K = 1/A$.

When there is a step input of size q_{in}, a solution only exists for times greater than zero, and this is shown as Equation 3.20. C_1 is evaluated at initial conditions yielding $C_1 = -Rq_{\text{in}}$.

$$h(t) = Rq_{\text{in}}\left(1 - \mathrm{e}^{-t/\tau}\right) \tag{3.20}$$

The time constant τ characterizes the response of the first-order system and is discussed in greater detail in the next section. All higher-order systems can be broken down into sets of first-order systems, and the time constants of these LDEs can be used to ascertain the relative importance of each from a dynamic response perspective. That is, the dominant, or largest, time constant will determine the speed of the response. The commonly used rule of thumb is that any subsystem with a time constant an order of magnitude (10 times) less than the dominant time constant can be described by steady-state or algebraic equations.

Some practical perspectives on capacitance

While the workshop associated with this chapter will illustrate capacitance with simulation, it is worthwhile examining the practical characteristics of capacitance because, as you will find out, capacitance can be a control engineer's best friend. A capacity-dominated system is described by

$$h(t) = Kq_{\text{in}}\left(1 - \mathrm{e}^{-t/\tau}\right) \tag{3.21}$$

Consider a step change in q_{in}. Mathematically, the change in $h(t)$ begins immediately, even though the full impact of the change in q_{in} will take some time. Now, consider a controller whose aim is to keep $h(t)$ at some set-point. Although we have not yet introduced controller algorithms and tuning, consider the most simple of control functions:

$$\Delta\text{MV} = K_{\text{c}}(\text{SP} - \text{PV}) = K_{\text{c}}e \tag{3.22}$$

where MV is the manipulated variable, K_{c} is the controller gain, PV is the process variable, SP is the set-point and e is the error. In short, corrective action carried out by the MV is simply a constant multiplier of the error.

Returning to our capacity-dominated process, the instant the PV (in this case $h(t)$) deviates from the set-point, an error e will be generated, and the MV will make some adjustment. The larger K_{c} is, the greater the corrective action. The fact that changes in the input (MV or q_{in}) have an immediate effect on the output (PV or $h(t)$) helps immensely, since the corrective action required to drive the error to zero is a straight algebraic function of the error. In the limit, as K_{c} approaches infinity, the error will be driven to zero and perfect control is achieved. Unfortunately, real life is not perfect, and controller gains never function at infinity. In practical terms, there is nothing that shows an absolutely 'immediate' response either, i.e. there is no 100 per cent pure capacitance system. However, this line of argument illustrates an important point: *for capacity-dominated systems, effective control can often be achieved using simple control and large gains*. As we will see in the next chapter, 'simple control' can be as simple as proportional-only control.

With such a simple approach to controlling capacity-dominated systems, it is easy to see why capacitance is often regarded as the control system engineer's ally. As with most things, too much is not good either. Recognise that large capacities typically increase capital cost. In addition, although capacitance acts as a buffer to upsets, if too large a volume of 'upset' material is allowed to accumulate, then it can take a long time to work out of the system. Thus, the wise process designer will typically use dynamic simulations to balance the trade-offs between the capital cost optimum and the dynamic operability and control optimum.

3.5.2 Dead time

We sometimes say that capacitance is our friend because it has the tendency to dampen out disturbances and lends itself to simple controls and straightforward tuning. Dead time, on the other hand, is typically regarded as the arch enemy of the process control engineer. Let us find out why.

Everyone has likely had the frustrating experience of showering in a very old building. The distance between the hot and cold water taps and the showerhead must be quite substantial, for it takes several seconds for you to feel the effect when you turn the tap. Assuming you have been lucky enough to get the water to a comfortable temperature, there is always the inconsiderate or unaware house guest who flushes the toilet mid shower. Quickly, you race to cut back on the scalding hot water source. There is no immediate effect and seconds seem like an eternity, so you crank the valve some more. The effect of your manipulation starts to make itself known, but within a few seconds you realise that you overdid it, and the water is suddenly freezing cold! Now your manual adjustments to the tap begin all over again, but may not settle down until a few more freeze – scald cycles play out. This example illustrates the menace of dead time. Although it may not be as entertaining, let us look at dead time mathematically and try to understand a little more about this dynamic process characteristic.

Dead time is a characteristic of a physical system that causes an input disturbance to be delayed in time, but unaffected in form. Whereas capacitance changes the form of the input disturbance (i.e. a step is filtered into a typical first-order curve), dead time is a *pure* delay of the input disturbance. Dead time is also referred to as transport lag, or distance velocity lag. A typical example process is the continuous weighing system shown in Figure 3.21.

For instance, assume there is a conveyor belt, L metres long, moving at some velocity v. The dead time t_{DT}, is calculated as shown

$$t_{\mathrm{DT}} = \frac{L}{v}[=]\left[\frac{\mathrm{m}}{\mathrm{m\ min}^{-1}}\right][=][\mathrm{min}] \tag{3.23}$$

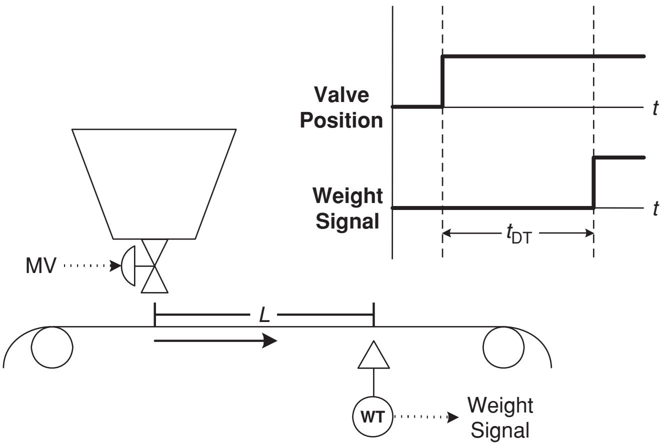

Figure 3.21 Continuous weighing system

If the valve is opened by some amount, increasing the material on the belt, there will be a delay equal to t_{DT} minutes before the increased weight is sensed at the weight sensor/transmitter.

Another classic example is a liquid flowing in a pipe. If the liquid is flowing at a velocity v through a pipe length l then an analogous situation exists to the weighing system. If a slug of liquid were followed through the pipe at the instant the valve is opened, it would take an amount time t_{DT} for the slug to go from one end of the pipe to the other. The delay times in these two cases would not necessarily be the same, but the delay effect is similar.

For a pure dead time element, assume that a step input of magnitude A occurs. The magnitude of the output step would also be A, except displaced in time by the dead time amount. The static gain K_{ss} would by definition be dimensionless and equal to one, as in Equation 3.24:

$$K_{ss} = \frac{A}{A} = 1 \tag{3.24}$$

Using the weighing system example of dead time, a 10 kg increase in material on the conveyor belt would result in a 10 kg increase at the weight sensor or a static gain of one. Similarly, it can also be shown that the above analysis holds for the pipe flow situation. In each of these cases a pure dead time exists since $K_{ss} = 1$. However, consider the scenario shown in Figure 3.22.

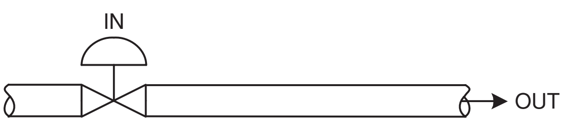

Figure 3.22 Valve/pipe flow system

In this scenario, the input to the valve is the input to the system and the output from the system is the flow through the pipe. If the opening of the valve is increased by some percentage of valve span A per cent, then an increase in flow of B $\mathrm{m^3\ s^{-1}}$ is delayed by an amount t_{DT}, where t_{DT} is the time it takes to see an increase in flow at the exit or measurement point in the pipe.

$$K_{\mathrm{ss}} = \frac{\Delta \mathrm{out}}{\Delta \mathrm{in}} = \frac{B\left(\mathrm{m^3\ s^{-1}}\right)}{A\,(\%)} \tag{3.25}$$

For the case shown in Equation 3.25, not only are there units but also the ratio of B to A is not necessarily one. It should be kept in mind that for pure dead time, i.e. the pipe alone, $K_{\mathrm{ss}} = 1$. However, for the case where another component is involved in the dead time, the component serves to supply units to the overall gain.

Some practical perspectives on dead time

The workshop associated with this chapter will illustrate dead time with simulation, and will show just how its presence makes it difficult to control a process. Let us look at why.

A system with capacitance and dead time (actually, quite a common combination) is described by

$$h\,(t) = K q_{\mathrm{in}}\left[1 - \mathrm{e}^{-(t-\mathrm{DT})/\tau}\right] \tag{3.26}$$

Consider a step change in q_{in}. Mathematically, the change in $h(t)$ will not be seen until DT time elapses. From that point on, the response in $h(t)$ will be exactly as that illustrated in the capacity-dominated system. Consider again, a controller whose aim is to keep $h(t)$ at some set-point. Also consider again the most simple of control algorithms:

$$\Delta\mathrm{MV} = K_{\mathrm{c}}(\mathrm{SP} - \mathrm{PV}) = K_{\mathrm{c}}e \tag{3.27}$$

where MV is the manipulated variable, K_{c} is the controller gain, PV is the process variable, SP is the set-point and e is the error. Note again that the corrective action carried out by the MV is simply a constant multiplier of the error. Unlike in the capacity-dominated system, the PV (in this case $h(t)$) will not react immediately to the change in q_{in}. For DT time units, $h(t)$ will go unaffected. Only after DT time will $h(t)$ begin to change. At that time MV will, as before, act to drive the error to zero. However, because there is no longer an instantaneous response of $h(t)$ to the MV, the error can no longer be driven to zero by a large gain. In fact, the larger K_{c} becomes, the more the controller is apt to overreact. Recall our shower example! Thus, dead time, by 'hiding' the disturbances that lurk in the system, makes the job of rejecting disturbances extremely difficult. *The larger the dead time, in proportion to the amount of capacitance, the more difficult* *control becomes*. It is largely the presence of dead time (along with process interactions and nonlinear process response) that keeps control engineers earning a decent living. In general, the more dead time can be designed or engineered out of a system the better. Also, for any given amount of dead time, the more capacitance the better. Can you think through why? *Hint*: 2 min of dead time in a chemical process with long response times (large capacitance) will not cause too much trouble. What about 2 min of dead time in the control loops of a jet airplane?

Let us summarize what we have just covered:

- Two types of feedback exist: *positive* and *negative* feedback. Only negative feedback produces stable control.
- Each element in a control loop has a particular action, or sign relationship between its input and output response. This action is important to understand, so that the action of the overall control loop produces negative feedback.
- Elements whose output increases with an increase in their input are said to be *direct-acting* (I/I) elements. Elements whose output decreases with an increase in their input are said to be *reverse-acting* (I/D) elements.
- Two important dynamic process response characteristics are *capacitance* and *dead time*.
- Capacitance acts to absorb and store mass and energy and, as such, tends to be a natural buffer to disturbances. Thus, this aids in process control . . . to an extent. Too much can create other problems, such as high capital cost and overly sluggish recovery from upsets. In general, capacity-dominated systems can be controlled with simple controls and large controller gains.
- Dead time imposes a pure delay on disturbances, effectively hiding the disturbance from the process, the measurements and the controls until it is well into the system. Dead time deteriorates controllability, especially if it is large relative to the amount of capacitance in the system with which it is associated. Dead time should generally be minimized as far as possible.

3.6 Process dynamic response

By this point, the reader should have an understanding of the need for and function of feedback control, should understand the elements of the feedback loop, and understand some of the qualitative features of the process dynamic response. While standard feedback control does not require extensive understanding of the process being controlled, some process understanding is important. In fact, the more we understand about the process, the easier our overall control system design may become. We will touch on

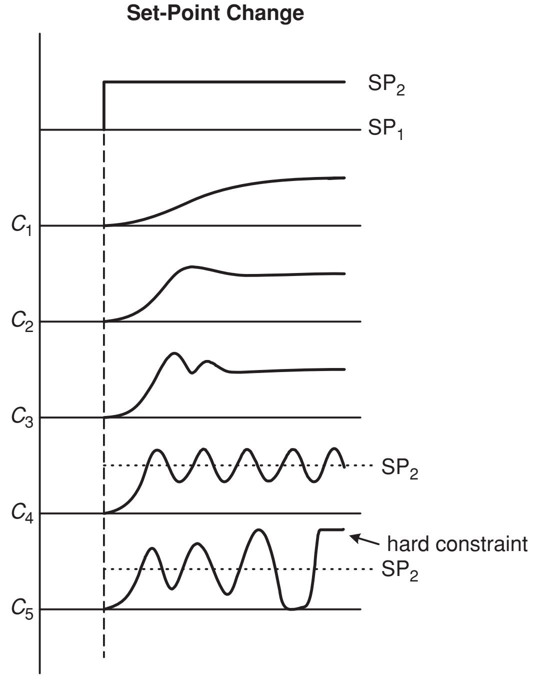

Figure 3.23 Typical PV response to set-point and load disturbance upsets

this more in Chapter 10 on plant-wide control. For now, let us simply take a look at the typical process dynamic responses seen in process plants. Process response determines how easily a process can be controlled, and also impacts the tuning required to achieve acceptable performance.

Up to now, we have looked primarily at control as a means of maintaining the PV at a fixed set-point in the presence of load disturbances. Set-point changes are also types of disturbances that a control loop must be able to handle. Production rate changes, for example, require that a flow rate set-point be changed. We will use the set-point change disturbance and the load disturbance as a means of illustrating process dynamic response.

As illustrated qualitatively in Figure 3.23, numerous real-time or dynamic responses are possible in returning the PV to the set-point. The response labelled C_1 in both cases shown would be classed as overdamped, i.e. a slow, sluggish return to set-point. C_2 presents the case for critical damping, i.e. the fastest return to set-point without oscillation. C_3 is a case where there is oscillation, and C_5 shows the case where instability is occurring, i.e. showing a hard constraint.

It is possible to adjust the feedback control loop to give any of the above responses. The form of the response desired depends on the process being controlled. For the most part, responses C_1 to C_3 would give desired behaviour, since each results in a return of the process variable to the desired set-point. C_4 is useful in some cases for adjusting the controller, which is also referred to as tuning. Several methods developed for controller tuning depend on information gained from the uniformly oscillating loop shown in C_4. C_5 results in instability and is not desirable for control.

3.7 Process modelling and simulation

Let's examine the response of SISO control systems in further detail. In order to examine a system and its response to disturbances, an understanding of the system equations is essential and a means by which to solve these model equations. The basic steps to examining a system dynamically are to determine the equations that describe the system, solve these equations for the desired solution, and then characterize the system response. The first two steps have already been done for the single tank scenario described by Figure 3.19. Now we will take this process one step further and examine the system response.

All process systems respond to various disturbances in different ways. Certain types of response are characteristic of specific types of process. Two of the most common personalities are those for first- and second-order systems. The single tank that was mathematically modelled in the previous section is an example of a first-order type of system.

3.7.1 First-order systems

If a step input is applied to a capacity-dominated process such as a single tank, the output begins to change instantaneously but does not reach its steady-state value for a period of time. This is true of any process that is capacitive in nature. It takes approximately 5τ for the output of the capacity process to reach its final, steady-state value. The time constant τ is defined as the amount of time it takes the output of the system to reach 63.2 per cent of its steady-state value. τ is a basic characteristic of capacity-dominated physical systems. The time constant τ can be defined in electrical terms as the product of the resistance times the capacitance:

$$\tau = RC \tag{3.28}$$

where R is the resistance in the system and C is its capacitance, with the units of each being appropriate for the system in question to make the time constant units be units of time, i.e. seconds, minutes, etc.

Figure 3.24 demonstrates the step response behaviour of the single tank example discussed previously. The equations describing the tank system were developed in the previous section and are clearly first-order differential equations. Any single-capacity system is typically a first-order system and will respond in the same manner illustrated in Figure 3.24.

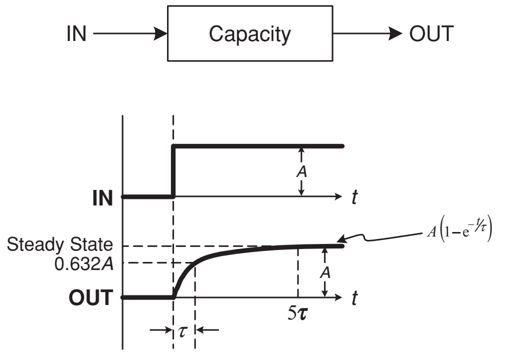

Figure 3.24 First-order system response to a step input

3.7.2 Second-order and higher-order systems

Higher-order responses are the result of multi-capacitance processes that contain vessels in series, fluid or mechanical components of a process that are subjected to accelerations causing inertial effects to become important, or the addition of controllers to a system. In a chemical plant, higher-order systems that result from a combination of capacities and controllers are very common. Typical examples are reactors in series, heat exchangers and distillation columns. In the case of distillation columns, when controllers are attached to the column, very high-order, nonlinear differential equations result when the system is mathematically modelled. Mechanical component time constant and natural frequencies are very small relative to the process time constants and frequencies, and, as such, the resultant effects are typically minor.

In order to get a feel for what a second-order system looks like, we will first examine a familiar component from the SISO system. An integrated part of the SISO system that results in a second-order differential equation is the diaphragm-operated control valve shown in Figure 3.25.

In order to derive the system equation we first apply Newton's second law, which states:

$$\sum_{i=1}^{n} F_i = Ma \tag{3.29}$$

The spring force, viscous friction and acceleration terms are described as follows:

$$\text{spring force} = KX_\mathrm{v} \tag{3.30}$$

$$\text{viscous force} = b\frac{\mathrm{d}X_\mathrm{v}}{\mathrm{d}t} \tag{3.31}$$

$$\text{acceleration term} = W\frac{\mathrm{d}^2X_\mathrm{v}}{\mathrm{d}t^2} \tag{3.32}$$

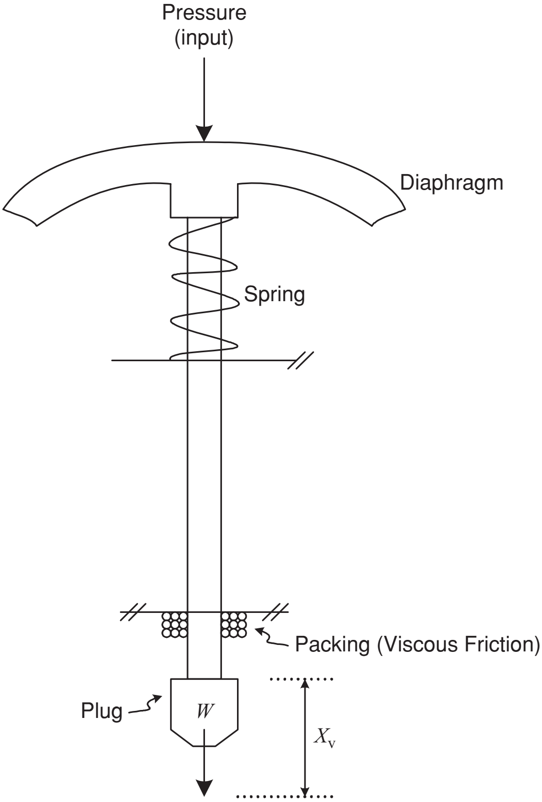

Figure 3.25 Diaphragm-operated control valve

The force is a function of time and is equal to the pressure applied to the top of the valve $P(t)$ times the area A. This term is referred to as the forcing function.

$$W\frac{d^2X_v}{dt^2} + b\frac{dX_v}{dt} + KX_v = AP(t) \tag{3.33}$$

where X_v is the position of the plug (output), P is the pressure at the input, K is the spring constant, b is the coefficient of viscous friction, and W is the weight of the plug and stem.

Equation 3.33 is obviously of the second-order differential form and, hence, when simulated will give a typical second-order response. To understand better what type of response these second-order systems will display, we will examine another common system and generalise its equations. The closed-form solution is best illustrated by an example that is familiar, namely the spring, mass, and damper system presented in Figure 3.26. Note that this is really just a simplification of the control valve example just cited.

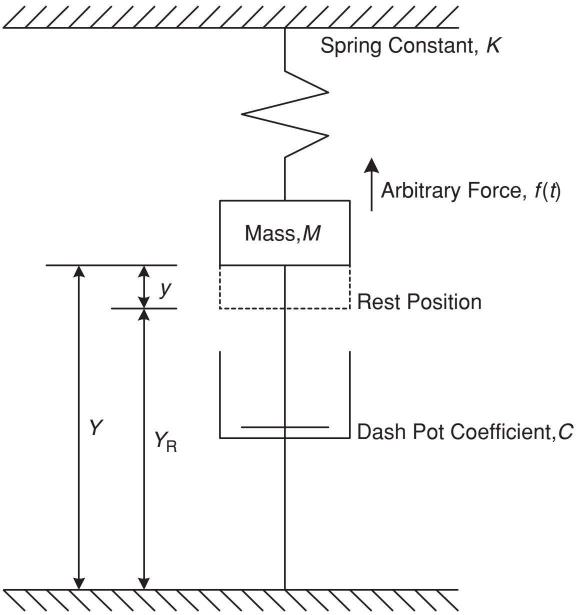

Figure 3.26 Spring, mass, and dash-pot (damper) system

If we perform the same analysis on this system as on the previous one, then the following equation is obtained describing the system where both the force and displacement are positive in the upward direction:

$$M\frac{\mathrm{d}^2Y}{\mathrm{d}t^2} + C\frac{\mathrm{d}Y}{\mathrm{d}t} + K\left(Y - Y_\mathrm{R}\right) = f\left(t\right) \tag{3.34}$$

Since the rest, or equilibrium, position Y_R is constant, Equation 3.34 can be rewritten in terms of the displacement y from the rest position. In this manner, we will be looking at variations about the equilibrium position, i.e. steady state. This is a common approach in system analysis, because analysis of even nonlinear systems about a steady state results in a linear system, i.e. ordinary differential equations (ODEs).

$$y = Y - Y_\mathrm{R} \tag{3.35}$$

$$\frac{\mathrm{d}y}{\mathrm{d}t} = \frac{\mathrm{d}Y}{\mathrm{d}t} \tag{3.36}$$

$$\frac{\mathrm{d}^2y}{\mathrm{d}t^2} = \frac{\mathrm{d}^2Y}{\mathrm{d}t^2} \tag{3.37}$$

Substituting Equations 3.35, 3.36, and 3.37 into Equation 3.34 results in

$$M\frac{d^2y}{dt^2} + C\frac{dy}{dt} + Ky = f(t) \tag{3.38}$$

Another point to be made about the analysis of our system is that we used the lumped parameter simplification. All the mass, friction (dash pot), and self-regulation (spring) are considered to be lumped at one point. The use of a lumped parameter model instead of a distributed parameter simplifies the mathematics of the model by producing ordinary differential instead of partial differential equations.

In order to solve the second-order equation which will give the position of the mass as a function of time, we need the specific set of initial conditions. For a second-order differential equation we use the following conditions:

$$Y(0) = \text{a negative constant} \tag{3.39}$$

$$\frac{d[Y(0)]}{dt} = 0 \tag{3.40}$$

In the simplest case the forcing function $f(t)$ can be set to zero. The resulting homogeneous differential equation can be found by finding the roots of the characteristic equation, given by

$$Mr^2 + Cr + K = 0 \tag{3.41}$$

For a second-order algebraic, these roots are given by Equation 3.42 and are called the eigenvalues.

$$r_{1,2} = \frac{-C \pm \sqrt{C^2 - 4KM}}{2M} \tag{3.42}$$

Provided that the roots are all real and unique, the solution is

$$y = C_1\,e^{r_1 t} + C_2\,e^{r_2 t} \tag{3.43}$$

C_1 and C_2 are constants evaluated using the two initial conditions. The resulting plot of y versus t will have one of the general responses shown in Figure 3.27. The system descriptive parameters, in this case C, K, and M, govern the particular response or behaviour of the system.

Second-order systems are very common in the chemical industry and have recently received much attention. The equation describing a second-order system such as the spring, mass, and damper example can be further generalized by dividing Equation

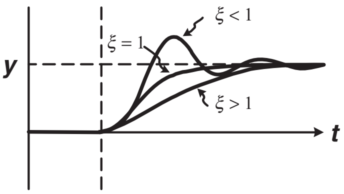

Figure 3.27 Typical second-order response

3.38 by M:

$$\frac{d^2y}{dt^2} + \left(\frac{C}{M}\right)\frac{dy}{dt} + \left(\frac{K}{M}\right)y = \frac{f(t)}{M} \tag{3.44}$$

Equation 3.44 can be further generalized by defining the following terms:

$$\omega_n = \sqrt{\frac{K}{M}} \tag{3.45}$$

$$\xi = \frac{C}{2\sqrt{MK}} \tag{3.46}$$

These generalised terms then characterize the response of the system. The term ω_n is called the undamped natural frequency, and ξ is known as the damping coefficient. Note that the natural frequency of the system is proportional to K (the tendency to self-regulate) and inversely proportional to M (the capacitance or inertia of the system). Note also that damping is directly proportional to C (system friction), but inversely proportional to M (capacitance or inertia) and K (self-regulation). Understanding how the frequency and damping in a system are affected by these fundamental process characteristics can be useful when the control scheme for a real chemical process is undertaken. Remember, sometimes some simple changes in the process itself can make the job of designing and tuning a regulatory control system much simpler!

The solution to Equation 3.44 can again be found by finding the roots of the following characteristic equation:

$$r^2 + 2\xi\omega_n r + \omega_n^2 = 0 \tag{3.47}$$

$$r = \frac{-2\xi\omega_n \pm \sqrt{4\xi^2\omega_n^2 - 4\omega_n^2}}{2} \tag{3.48}$$

which simplifies to

$$r = -\xi\omega_n \pm \omega_n\sqrt{\xi^2 - 1} \tag{3.49}$$

or

$$y(t) = e^{-\xi\omega_n t}\left[C_1\, e^{\left(-\omega_n\sqrt{\xi^2-1}\right)t} + C_2\, e^{\left(\omega_n\sqrt{\xi^2-1}\right)t}\right] \tag{3.50}$$

The response of the system will depend mainly on the damping coefficient ξ. When $\xi < 1$, the system is underdamped and has an oscillatory response. The smaller the value of ξ, the greater the overshoot. If $\xi = 1$, the system is termed critically damped and has no oscillation. A critically damped system provides the fastest approach to the final value without the overshoot of an underdamped system. Finally, if $\xi > 1$, the system is overdamped. An overdamped system is similar to a critically damped system, in that the response never overshoots the final value. However, the approach for an overdamped system is much slower and varies depending upon the value of ξ. These typical responses are illustrated in Figure 3.27.

Now let us examine the case of multiple capacities in series. Consider the two non-interacting tanks in series shown in Figure 3.28.

The mass balances for Tank 1 and Tank 2 are given by Equations 3.51 and 3.52 respectively:

$$A_1 \frac{dh_1}{dt} = F_i - F_1 \tag{3.51}$$

$$A_2 \frac{dh_2}{dt} = F_1 - F_2 \tag{3.52}$$

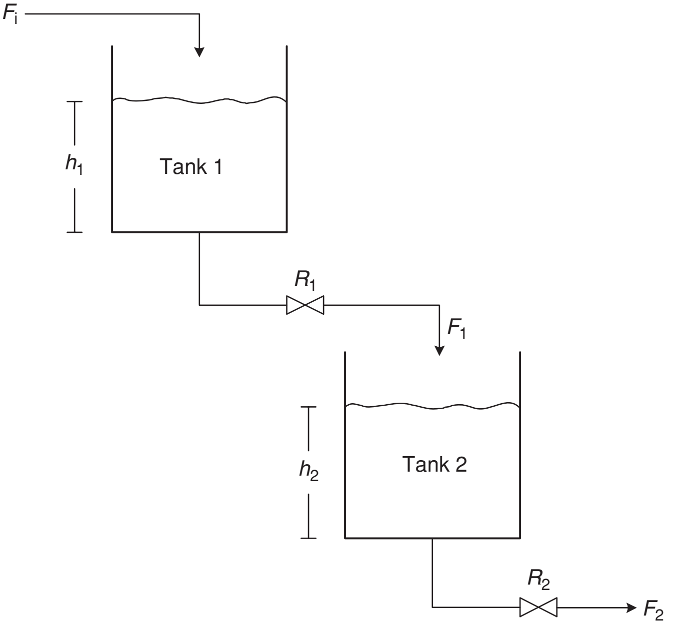

Figure 3.28 Two non-interacting tanks in series

If linear resistance to flow is assumed for the valves in the system, then the following equations are obtained:

$$F_1 = \frac{h_1}{R_1} \tag{3.53}$$

$$F_2 = \frac{h_2}{R_2} \tag{3.54}$$

Now substitute Equations 3.53 and 3.54 into Equations 3.51 and 3.52 to give

$$A_1 R_1 \frac{dh_1}{dt} + h_1 = F_i R_1 \tag{3.55}$$

$$A_2 R_2 \frac{dh_2}{dt} + h_2 = \frac{R_2}{R_1} h_1 \tag{3.56}$$

Differentiating Equation 3.56 with respect to time and rewriting h_1 in terms of h_2 gives a second-order ODE.

$$\frac{d^2 h_2}{dt^2} + \left(\frac{1}{A_2 R_2} + \frac{1}{A_1 R_1}\right)\frac{dh_2}{dt} + \left(\frac{1}{A_1 A_2 R_1 R_2}\right) h_2 = \frac{F_i}{A_1 A_2 R_1} \tag{3.57}$$

We can then apply the same generalisation to Equation 3.57 as we did for Equation 3.44. This generalisation gives the following:

$$2\xi\omega_n = \frac{1}{A_2 R_2} + \frac{1}{A_1 R_1} \tag{3.58}$$

$$\omega_n = \sqrt{\frac{1}{A_1 A_2 R_1 R_2}} \tag{3.59}$$

Now that the equations describing the system have been developed, the system can be simulated and its response to disturbances examined. Based on the equations developed for the single tank and the non-interacting tanks in a series, what type of response would the level in the first and second tank display?

3.7.3 Simple system analysis

Questions often arise in the design of a process concerning the controllability of the system, alternative control schemes, and variations in the process design to achieve quality and/or throughput. In order to answer such questions, without building the plant,

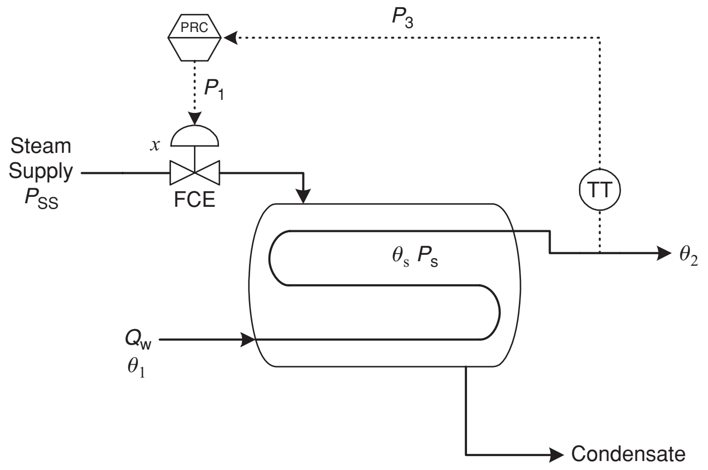

Figure 3.29 Typical process schematic of shell and tube heat exchanger

it is necessary to have available a rigorous mathematical model or modelling system. Once the system, which includes plant unit operations and controllers, is modelled and simulated, the effect of various parameters and control schemes can be predicted and evaluated.

The determination of the process mathematical model is often the most difficult and time-consuming step in control system analysis. This is a result of the dynamic nature of the process; in other words, how the system reacts during upsets or disturbances. The problem is further complicated by process nonlinearities and time-varying parameters. To illustrate the modelling procedure we will look at developing a model for a shell and tube heat exchanger with temperature control [7], shown in Figure 3.29.

Constructing a word block diagram

Before starting to analyse the process, it is helpful to construct a system block diagram (Figure 3.30). The purpose of the block diagram is to identify the components of the SISO and all load disturbances. Recall that a simple SISO feedback loop is comprised of four basic units:

1 process
2 sensing element/transmitter
3 controller
4 FCE (typically valve and positioner).

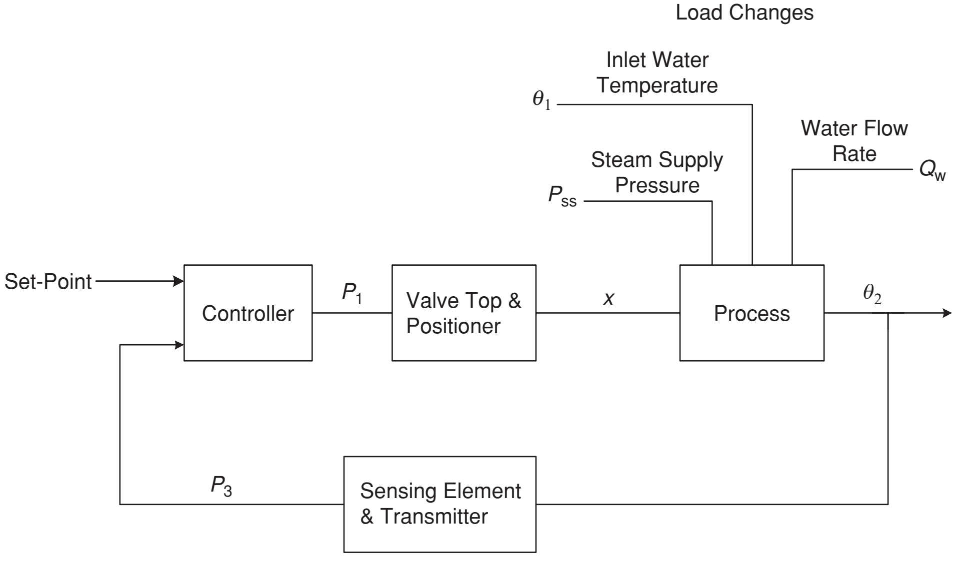

Figure 3.30 Word block diagram of shell and tube heat exchanger

In order to limit the complexity of the analysis, all parameters will be assumed to be constant. The procedure will be to develop the ODEs for each of the components and combine these into a system model.

Valve top and positioner

One approach to modelling SISO loop components is through step testing. For the valve and valve positioner, a first-order ODE such as the following may be determined:

$$\tau \frac{\mathrm{d}\Delta X}{\mathrm{d}t} + \Delta X = K \Delta P_1 \tag{3.60}$$

where X is the valve stem position, P_1 is the valve top pressure (that opens and closes the pneumatic valve, after the controller compares the set-point to P_3), τ is the first-order time constant, K is the steady state gain, and t is the time. For this valve and valve position, the steady-state gain is $K = 0.179$ mm kPa^{-1} and the time constant is $\tau = 0.033$ min. Therefore, we can write Equation 3.60 as

$$0.033 \frac{\mathrm{d}\Delta X}{\mathrm{d}t} + \Delta X = 0.179 \Delta P_1 \tag{3.61}$$

Equations 3.60 and 3.61 are written in terms of variation variables. Variation variables represent a change from or about a steady-state level of the variable. The gain was determined by dividing the valve stem travel by the pneumatic signal range, which is 14.3 mm divided by 80 kPa (the range in this case is from 20 to 100 kPa). The time constant of 0.033 is determined experimentally using a step input signal (discussed previously in Section 3.7.1).

Sensing element and transmitter

These components can also be represented by a first-order ODE, as in Equation 3.62. In most cases this is an adequate representation for SISO loop analysis.

$$\tau \frac{\mathrm{d}\Delta P_3}{\mathrm{d}t} + \Delta P_3 = K\,\Delta\theta_2 \tag{3.62}$$

where P_3 is the pressure signal from the temperature sensor/transmitter to the controller and θ_2 is the exit water temperature. The steady state gain K is determined from the sensor/transmitter ranges, which are 0–100°C input and 20–100 kPa output. This results in a gain of 0.80 kPa °C^{-1}.

The time constant of the measuring element equals the thermal resistance times the capacitance. The thermal resistance can be modelled as a function of area and the heat transfer film coefficient:

$$R_{\mathrm{t}} = \frac{1}{A\,h} \tag{3.63}$$

where R_{t} (°C kW^{-1}) is the thermal resistance, A (m^2) is the surface area, and h (kW °C^{-1} m^{-2}) is the film coefficient.

The thermal capacitance is a function of mass and specific heat:

$$C_{\mathrm{t}} = WC \tag{3.64}$$

where C_{t} (kJ°C^{-1}) is the thermal capacitance, W (kg) is the mass of the sensing element, and C (kJ kg^{-1} °C^{-1}) is the specific heat.

The time constant, which has the units of hours as a result of the units of the film coefficient, can then be found using Equations 3.63 and 3.64:

$$\tau = R_{\mathrm{t}}C_{\mathrm{t}} = \frac{WC}{Ah} \tag{3.65}$$

Once the sensing element has been selected, its area, weight, and specific heat are fixed. Therefore, τ is only a function of h. If the manufacturer provides a time constant τ for a given set of conditions, other τ can be estimated based on the new conditions. If the system properties are about the same as those the manufacturer used during the step tests, then h will primarily be a function of fluid velocity. Also, the bulb has a time constant over a limited velocity range as follows:

$$\tau = 1.15\frac{1}{v^{0.58}} \tag{3.66}$$

where τ(s) is the time constant and v (m s^{-1}) is the fluid velocity.

Thus, for a fluid velocity of 0.686 m s^{-1}, the time constant of the sensing element is 0.024 min (1.43 s).

Process model

The process model can be determined either from first principles (the mechanistic approach) or by 'black boxing'. Mechanistic approaches attempt to model transient energy and mass balance. Black-box models are often even simpler and describe the input-output behaviour with no recourse to conservation principles. Using a linear system analysis approach, the heat exchanger is modelled as one lump (lumped parameter approach) in which a small change is made in the valve stem position ΔX and its effect on the outlet water temperature θ_2 is predicted. A change in X results in more or less steam entering the shell, which changes the energy input to the heat exchanger. This change in energy input is accounted for by a change within the exchanger and a change in the energy leaving the exchanger. If the inlet water flow rate Q_{w}, and inlet temperature θ_1 are constant, then any change in energy will show up as a change in water outlet temperature θ_2.

The steam flow Q_{s} through the valve can be modelled as follows:

$$Q_{\mathrm{s}} = f(X, P_{\mathrm{s}}) = \frac{3}{2}C_{\mathrm{v}}P_{\mathrm{s}} = 0.000\,86XP_{\mathrm{s}} \tag{3.67}$$

where Q_{s} (kg s^{-1}) is the steam flow, X (mm) is the valve stem position or travel, and P_{s} (kPa) is the steam pressure.

In terms of variation variables, the change in steam flow can be modelled as follows:

$$\Delta Q_{\mathrm{s}} = \alpha\Delta P_{\mathrm{s}} + \beta\Delta X \tag{3.68}$$

where

$$\alpha = \left(\frac{\partial Q_{\mathrm{s}}}{\partial P_{\mathrm{s}}}\right)_X = 0.000\,86X_{\mathrm{op}}$$

$$\beta = \left(\frac{\partial Q_{\mathrm{s}}}{\partial X}\right)_{P_{\mathrm{s}}} = 0.000\,86P_{\mathrm{s_{op}}}$$

and, 'op', is the operating point around which X and P_{s} are defined.

Table 3.1 Saturated steam temperature

P_s (kPa)	θ_s (°C)
300	133.5
350	138.9
400	143.6
450	147.9
500	151.8

If one assumes saturated steam in the shell, then there is a unique relationship between changes in steam pressure P_s and changes in steam temperature θ_s:

$$\Delta P_s = \gamma \Delta\theta_s = \left(\frac{\partial P_s}{\partial \theta_s}\right)\theta_s \tag{3.69}$$

The coefficient γ can be evaluated from steam tables at the shell nominal operating pressure of 432 kPa; refer to Table 3.1. As can be seen from Figure 3.31, at an operating pressure of 432 kPa, γ can be linearised to give a value of 11.0 kPa °C^{-1}.

An energy balance on the shell using the principle of energy conservation can be written as follows:

$$\Delta Q_s m = (W_t C_t + W_s C_s)\,\Delta\theta_s + h_t A_t\left(\Delta\theta_s - \frac{\Delta\theta_2}{2}\right) \tag{3.70}$$

where m (kJ kg^{-1}) is the latent heat of condensing steam (2129 kJ kg^{-1} for this example), W_t (kg) is the tube weight, C_t (kJ kg^{-1} °C^{-1}) is the specific heat of tube material (0.5 kJ

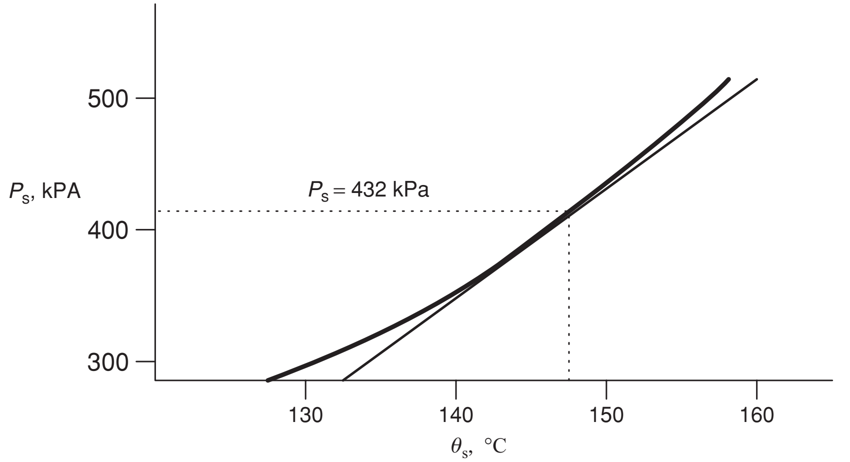

Figure 3.31 Steam pressure versus steam temperature

$\mathrm{kg^{-1}\ {}^\circ C^{-1}}$ for this example), W_s (kg) is the shell weight (19.6 kg for this example), C_s ($\mathrm{kJ\ kg^{-1}\ {}^\circ C^{-1}}$) is the specific heat of shell material, h_t ($\mathrm{kW\ {}^\circ C^{-1}\ m^{-2}}$) is the tube-side film coefficient, A_t ($\mathrm{m^2}$) is the tube area ($h_t A_t = 0.376\ \mathrm{kJ\ s^{-1}\ {}^\circ C^{-1}}$), θ_s (°C) is the steam temperature and θ_2 (°C) is the outlet temperature.

The shell-side energy balance contains a number of simplifying assumptions. The shell-side heat transfer film coefficient is assumed to be negligible; thus, the temperatures of the tube and shell walls are equal to the condensing steam temperature. Shell steam capacity is also assumed negligible due to the small volume. Losses to the atmosphere are neglected, i.e. the shell is well insulated.

In the development of a mathematical model the validity of assumptions is always debatable and depends on the use of the model, the required accuracy, the equipment size and the configuration. These need to be considered in light of where and how the model is to be used.

The water temperature was taken as the average between the inlet and outlet temperatures. This assumption is valid since the inlet temperature is assumed to be constant; hence, the change in the average water temperature is half the change in the outlet water temperature. An energy balance for the water flowing in the tube side results in the following equation:

$$W_w C_w \frac{\mathrm{d}\left(\Delta\theta_{2/2}\right)}{\mathrm{d}t} + Q_w C_w \Delta\theta_2 = h_t A_t \left(\Delta\theta_s - \frac{\Delta\theta_2}{2}\right) \tag{3.71}$$

where Q_w ($\mathrm{kg\ s^{-1}}$) is the water flow into the tube, C_w ($\mathrm{kJ\ kg^{-1}\ {}^\circ C^{-1}}$) is the specific heat of water (4.2 $\mathrm{kJ\ kg^{-1}\ {}^\circ C^{-1}}$ for this example) W_w (kg) is the water weight in the tube, h_t ($\mathrm{kW\ {}^\circ C^{-1}\ m^{-2}}$) is the tube-side film coefficient, A_t ($\mathrm{m^2}$) is the tube area, θ_s (°C) is the steam temperature, and θ_2 (°C) is the outlet temperature.

Controller

If we use a standard PI controller, it can be modelled using Equation 3.72. A PI controller takes remedial action proportional to the magnitude of both the error and the integral of the error, and is rigorously defined in Chapter 4.

$$\Delta P_1 = K_c \left(\Delta e + \frac{1}{T_i}\int_0^t \Delta e\ \mathrm{d}t\right) \tag{3.72}$$

where K_c is the controller gain (an adjustable tuning parameter of the controller), T_i is the integral time (another adjustable tuning parameter of the controller), and Δe is the error and is defined as the difference between the measured variable and the set-point, which is 65°C for this example.

$$\Delta e = \Delta P_3^{\mathrm{SP}} - \Delta P_3 = 65 - \Delta P_3 \tag{3.73}$$

Response

The time response of the outlet temperature to various load disturbances can be determined by integrating the set of ODEs, as developed previously. This can be accomplished by using one of the standard mathematical software packages, such as MATLAB™ [8]. For clarity, these equations are repeated below with their original numbering and are in the order that they appear in the control loop, with the values of the constant parameters shown.

$$0.033\frac{\mathrm{d}\Delta X}{\mathrm{d}t} + \Delta X = 0.179\Delta P_1 \tag{3.61}$$

$$\Delta Q_\mathrm{s} = 0.000\,86X_\mathrm{op}\Delta P_\mathrm{s} + 0.000\,86P_{\mathrm{s_{op}}}\Delta X \tag{3.68}$$

$$\Delta P_\mathrm{s} = 11.0\Delta\theta_\mathrm{s} \tag{3.69}$$

$$(2129)\,\Delta Q_\mathrm{s} = (0.5\,W_t + 0.5 \times 19.6)\,\Delta\theta_\mathrm{s} + \tag{3.70}$$

$$0.376\left(\theta_\mathrm{s} - \frac{\Delta\theta_2}{2}\right) \tag{3.71}$$

$$3.96\frac{\mathrm{d}\left(\Delta\theta_{2/2}\right)}{\mathrm{d}t} + Q_\mathrm{w}4.2\Delta\theta_2 = (0.376)\left(\Delta\theta_\mathrm{s} - \frac{\Delta\theta_2}{2}\right) \tag{3.72}$$

$$0.024\frac{\mathrm{d}\Delta P_3}{\mathrm{d}t} + \Delta P_3 = 0.80\Delta\theta_2 \tag{3.62}$$

$$\Delta P_1 = K_\mathrm{c}\left[(65 - \Delta P_3) + \frac{1}{T_\mathrm{i}}\int_0^t (65 - \Delta P_3)\ \mathrm{d}t\right] \tag{3.72 and 3.73}$$

The resulting response for this system to a PI controller set-point change is shown in Figure 3.32.

The response in Figure 3.32 is based on a simple linear lumped parameter model; thus, for large disturbances or set-point changes that exceed the limits of the linear assumptions and other operating point, quite different responses will be obtained.

3.7.4 Classical modelling for control approaches

The previous simple analysis example follows a pre-computing classical approach where a simple linearized lumped parameter model of the system was developed. In the pre-computing or classical approach this simple model was solved by the application of analytical methods such as Laplace transforms and frequency-response analysis. For completeness, these methods will be briefly introduced here. The interested reader should refer to the texts that take this pre-computing classical approach, such as

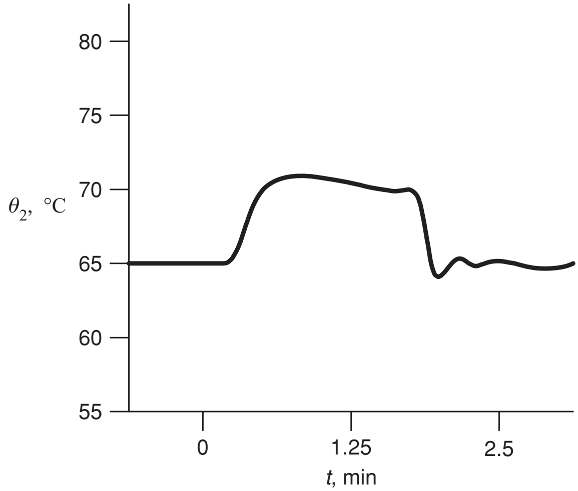

Figure 3.32 Typical response for a PI controller set-point change

Coughanowr and Koppel [9], Luyben [10], Harriott [11], Murrill [12], and Shinskey [5].

Laplace transforms

Solving Laplace transforms is a process of leaving the time domain, where a differential equation may be too difficult to solve without a computer, and entering the Laplace domain, where the transform of that differential equation is readily solved, and then coming back into the time domain with the inverse transform of the Laplace domain solution.

Differential equations maybe classified as per the taxonomy in Figure 3.33.

HCCLODEs and NHCCLODEs are of the form given in Equation 3.74 .

$$a_0\frac{\mathrm{d}^n y}{\mathrm{d}t^n} + a_{n-1}\frac{\mathrm{d}^{n-1} y}{\mathrm{d}t^{n-1}} + \cdots\cdots + a_1\frac{\mathrm{d}y}{\mathrm{d}t} + a_0 y = f(t) \tag{3.74}$$

Mathematically, equations of this type comprise a very narrow class of equations, but they represent quite a number of simple physical systems. For homogeneous systems, $f(t) = 0$. The simplest HCCLODE is given by

$$a_1\frac{\mathrm{d}y}{\mathrm{d}t} + a_0 y = 0 \tag{3.75}$$

An example of a physical system represented by Equation 3.75 is the linearised equation for change of level h, in a cylindrical tank of cross-sectional area A, with no

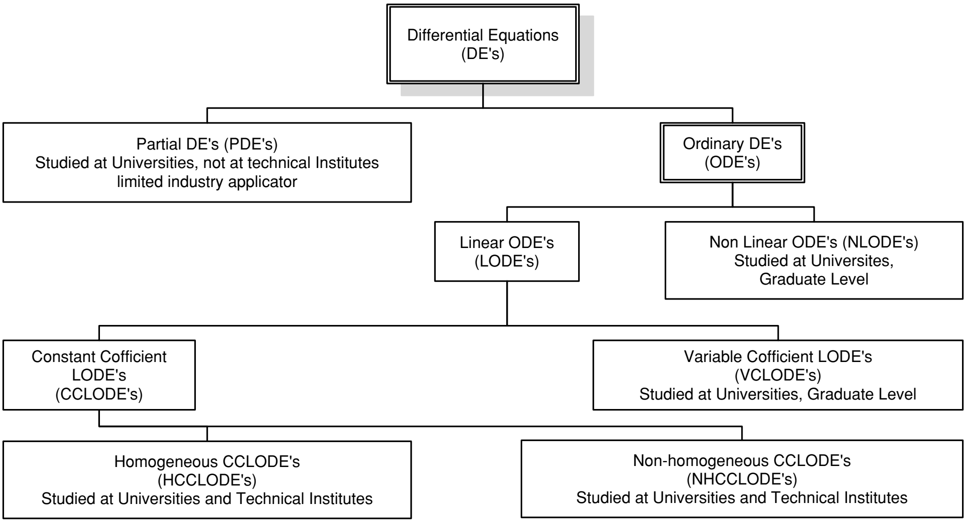

Figure 3.33 Differential equations

input flow, and output flow regulated by a valve of resistance R.

$$RA\frac{\mathrm{d}h}{\mathrm{d}t} + h = 0 \tag{3.76}$$

Other examples are the voltage V in a resistor capacitor RC circuit without an applied input voltage, and the change of current i in a resistor–inductor RL circuit without an applied input current, as described by Equations 3.77 and 3.78 respectively.

$$RC\frac{\mathrm{d}V}{\mathrm{d}t} + V = 0 \tag{3.77}$$

$$Ri\frac{\mathrm{d}L}{\mathrm{d}t} + i = 0 \tag{3.78}$$

In the above examples, the equations describe physical systems with one energy storage or capacitance element, and one energy-dissipating or resistive element and are not subject to external forces.

Solving a first-order equation such as the above example can be done using analytical integration:

$$\frac{\mathrm{d}y}{\mathrm{d}t} = \frac{-a_0}{a_1}y$$

from which

$$y = y(0)\,\mathrm{e}^{(-a_0/a_1)/t}$$

where $y(0)$ is the initial energy. However, this is the only type of equation that can be solved thus.

To solve this equation using Laplace transforms we use the following definition of the Laplace transform (Equation 3.79) and standard transforms of functions, e.g. Coughanowr and Koppel [9];

$$L\{f(t)\} = F(s) = \int_0^\infty \mathrm{e}^{-st}\, f(t)\, \mathrm{d}t \tag{3.79}$$

We operate like so:

$$L\left\{a_1 \frac{\mathrm{d}y}{\mathrm{d}t} + a_0 y = 0\right\} = a_1(sY - y_0) + a_0 Y = 0$$

or

$$Y = \frac{y_0}{s + \frac{a_0}{a_1}}$$

from which

$$\frac{\mathrm{d}y}{\mathrm{d}t} = \frac{-a_0}{a_1} y \quad \textit{and} \quad y = y(0)\, \mathrm{e}^{(-a_0/a_1)/t}$$

Note that coming out of the Laplace domain is more difficult than entering it, and extensive use of partial fractions is necessary for this job.

Transfer functions

A transfer function is the relationship between the input and output of a system. In classical control systems' literature that makes use of Laplace transforms, extensive use is made of Laplace transfer functions. Table 3.2 presents the transfer functions of common process systems dynamics.

Frequency response

Frequency response analysis is another classical tool that has been used in the analysis and design of process control systems. The Laplace variable s is replaced by $\mathrm{j}\omega$, where $\mathrm{j} = \sqrt{-1}$ and ω is the radian frequency. The frequency response is then plotted using an Argand diagram approach.

For example, consider the frequency response of the first-order process given by the transfer function $G(s) = 1/(\tau s + 1)$.

Table 3.2 Transfer functions of common process systems dynamics

System	Time-domain equation	Laplace transfer function
	$\frac{\mathrm{d}y}{\mathrm{d}t} = f(y, x, t)$	$\frac{Y(s)}{X(s)}$
First-order system	$\tau\frac{\mathrm{d}y}{\mathrm{d}t} + y = K_\mathrm{p}x$	$\frac{K_\mathrm{p}}{\tau s + 1}$
Two non-interacting first-order systems in series	$\tau\frac{du}{dt} + u = K_\mathrm{p}x$ $\tau\frac{\mathrm{d}y}{\mathrm{d}t} + y = K_\mathrm{p}u$	$\frac{K_\mathrm{p}^2}{(\tau s + 1)(\tau s + 1)}$
Second-order system	$\tau^2\frac{\mathrm{d}^2y}{\mathrm{d}t^2} + 2\tau\xi\frac{\mathrm{d}y}{\mathrm{d}t} + y = K_\mathrm{p}x$	$\frac{K_\mathrm{p}}{\tau^2s^2 + 2\tau\xi s + 1}$
Dead time	$y = x(t - L)$	e^{-Ls}

The Laplace variable s is replaced by $\mathrm{j}\omega$ and converted to polar form. Then the amplitude ratio (AR) and phase α are computed via Equations 3.80 and 3.81:

$$\mathrm{AR} = \frac{1}{\sqrt{\omega^2\tau^2 + 1}} \tag{3.80}$$

$$\alpha = \tan^{-1}(-\omega\tau) \tag{3.81}$$

These can be plotted either on Bode graph paper or Black–Nichols graph paper, and if converted to decibels (db) via $\mathrm{AR_{db}} = 20\log(\mathrm{AR})$ the response can be plotted on semi-log and polar graph paper.

The concepts of gain margin and phase margin are used for control-system design using frequency-response analysis. Gain margin is the additional amount of gain that would destabilize the system. Phase margin represents the additional amount of phase lag required to destabilize the system. Both gain margin and phase margin can be obtained from the open-loop Bode diagram of the system. Typical design specifications are that the gain margin should be greater than 1.7 and the phase margin must be greater than 30°.

3.7.5 The modern modelling for control approach

The modern approach to process modelling for process control that this book takes (which is described in Chapter 1) is to make use of simulation tools and computer-based design packages that avoid the limitations imposed by the analytical design methods, namely abstraction, linearisation and simplification, e.g. Allen [13].

Mathematical models imbedded in today's simulation software provide a means to handle both variations in operating level and process nonlinearities. The important result is not the controller settings, but the ability of the designer to manipulate process parameters to meet process specifications at a minimum cost. This can only be achieved by developing a deep understanding of the process and the system, which includes the process and the controllers.

3.8 References

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4

Basic control modes

The previous chapter discussed basic feedback control concepts, including the vital role of the controller. Again, the purpose of the controller in regulatory control is to maintain the controlled variable at a predetermined set point. This is achieved by a change in the manipulated variable using a pre-programmed controller algorithm. This chapter will describe the basic control modes or algorithms used in controllers in feedback control loops.

4.1 On–off control

The most rudimentary form of regulatory control is *on–off control*. This type of control is primarily intended for use with FCEs that are non-throttling in nature, i.e. some type of switch as opposed to a valve. An excellent example of on–off control is a home heating system. Whenever the temperature goes above the set point, the heating plant shuts off, and whenever the temperature drops below the set point, the heating plant turns on. This behaviour is shown by Equation 4.1:

$$\text{MV} = 0\% \text{ for PV} > \text{SP and MV} = 100\% \text{ for PV} < \text{SP} \tag{4.1}$$

The controller output MV is equal to 0 per cent or off whenever PV exceeds the set point SP. Whenever the process variable is below the set point, the controller output is equal to 100 per cent or on.

The most useful type of process where on-off control can be successfully applied is a large capacitance process where tight level control is not important, i.e. for the case of flow smoothing. A good example of this type of process is a surge tank. A large capacitance is important owing to the nature of the controller action and its effect on the operational life of the FCE. This leads us to one of the disadvantages of an on-off type of controller. Owing to the continual opening and closing of the controller, the FCE quickly becomes worn and must be replaced. This type of control action is illustrated in Figure 4.1, which shows the typical behaviour of an on–off controller.

In this example, at time $t = 0$, PV is less than SP, and MV is equal to 100 per cent. When PV crosses the set point, MV becomes 0 per cent. The temperature rises somewhat

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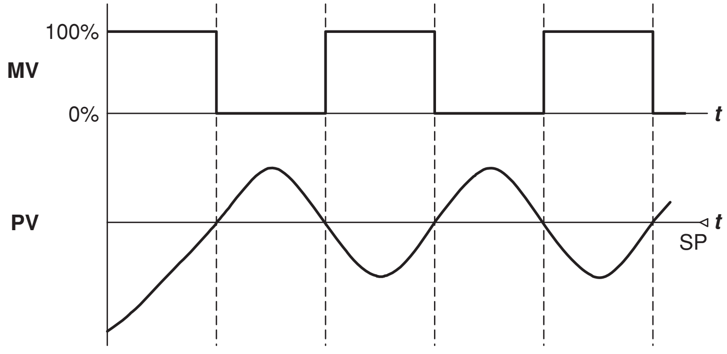

Figure 4.1 On–off controller response

above the set point before the controller turns off because of dead time, the capacitance of the heating system, and heat transfer to the ambient. These factors are termed system dynamics. When the temperature drops below the set point, the controller opens the valve. However, again due to system dynamics, the temperature drops somewhat below the set point before PV begins to rise again. It is easy to see how the FCE would quickly become worn out when this action is continually occurring.

Since the controller cannot throttle the actuator, but only turn it on or off, the primary characteristic of on–off control is that the process variable is always cycling about the set point. The rate at which PV cycles and the deviation of PV from the set point are a function of the dead time and capacitance in the system, or the system dynamics. The longer the lag time, the slower the cycling, but the greater the deviation from the set point. This can better be illustrated by using an on–off controller with a differential gap or dead band, as shown in Figure 4.2. Most on–off controllers are built with an adjustable differential gap or dead band, inside which no control action takes place. The intent of this differential gap is to minimize cycling of the controller output and extend the operational life of the FCE.

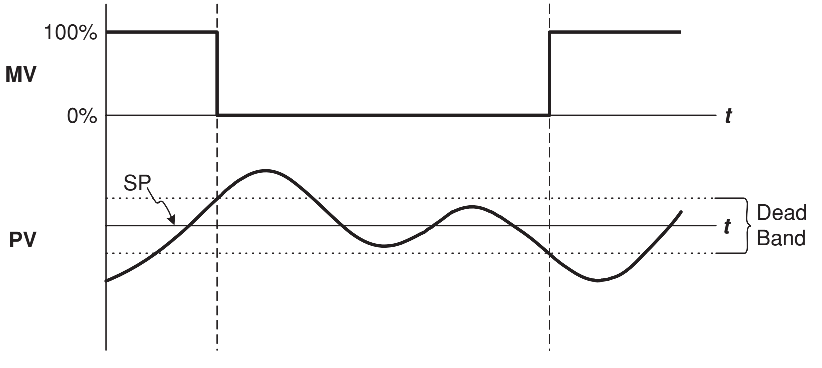

Figure 4.2 On–off controller with dead band

The controller switches off when the process variable exits the dead band on the high side and does not turn on again until PV is outside the dead band on the low side. The frequency of cycling is reduced, but the deviation from the set point is increased. If the dead band is reduced then the frequency of cycling is increased, but deviation from set point is decreased.

Typically, the dead band is adjusted as a percentage of the process variable span. Using the heating system example, suppose the temperature measurement range was from 20 to 120°C. Setting the dead band equal to 10 per cent of the span, the dead band in degrees would be 10°C. If the set point were 75°C, then the upper edge of the dead band would be $75°\mathrm{C} + 5°\mathrm{C} = 80°\mathrm{C}$ and the lower edge of the dead band would be $75°\mathrm{C} - 5°\mathrm{C} = 70°\mathrm{C}$, giving the dead band width of $80°\mathrm{C} - 70°\mathrm{C} = 10°\mathrm{C}$.

With an on–off controller, cycling cannot be eliminated. When a large lag is present in the process, the deviation from the set point may not be perceptible, since the amount of time per cycle is longer. If this is acceptable, then an on–off controller can be used. However, in order to eliminate cycling completely, another control mode would need to be implemented.

4.2 Proportional (P-only) control

Proportional control is the simplest continuous control mode that can damp out oscillations in the feedback control loop. This control mode normally stops the process variable PV from cycling, but it does not necessarily return it to the set point.

For example, consider the liquid level control situation given in Figure 4.3, in which the tank must not overflow or run dry. If the inflow F_i is equal to the outflow F_0, then the level, as seen in the sight glass, remains constant. If F_0 increases such that it is greater than F_i, then the level will begin to drop. In order to stop the level from dropping, F_i must be increased by opening the inflow valve until it is equal to F_0, and the level stops dropping. However, the tank level is no longer at the initial level; it has dropped to a new steady-state level. The amount by which the level drops depends on how much the inflow valve had to move to make F_i equal to F_0. A similar situation would occur if F_0 was less than F_i, only in this case the level would rise until the readjusted inflow equals the outflow. This scenario describes what a proportional controller would do if it were connected to the tank.

In equation form, the output of a proportional controller is proportional to the error (where error is the deviation of the measurement CV from the set point SP):

$$\mathrm{MV} = K_\mathrm{c} e \tag{4.2}$$

where K_c is the controller gain, e is the error, and MV is the manipulated variable. Remember:

$$e = \mathrm{SP} - \mathrm{CV} \qquad \text{(for reverse acting)}$$

$$e = \mathrm{CV} - \mathrm{SP} \qquad \text{(for direct acting)}$$

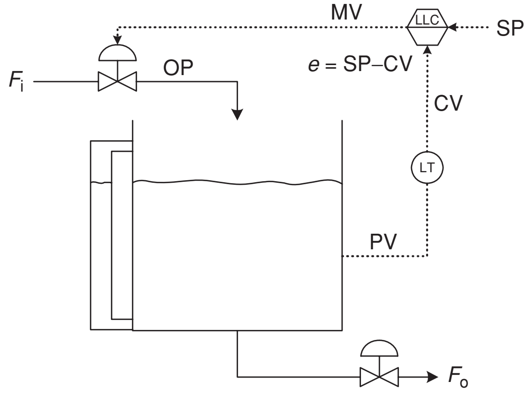

Figure 4.3 Liquid level control: proportional mode

To allegorise proportional control we will use the liquid level loop shown in Figure 4.3. Initially, the proportional controller is placed in manual and the level in the tank is manually adjusted to equal the set point. With F_i equal to F_0, the level should stay at the set point. Also, set $F_0 = 50\% = F_i$, CV = SP = 50%, and $K_c = 2$. Note that here we are using the percentage of span units. Now, if the controller is placed into automatic mode what will happen to the output? At the instant the controller is placed into automatic mode, the error will be zero since CV is equal to SP, and the controller output will also be zero:

$$\text{MV} = 2(50 - 50) = 0\,(\text{from Equation 4.2})$$

For a controller output of zero, what will the level do? The level will begin to drop. To stop this movement F_i and F_0 must equal 50 per cent again. If a linear relationship is assumed between inflow and controller output, then for $F_i = 50\%$ we will have MV = 50%.

Since	$\text{MV} = K_c e = 2(\text{SP} - \text{CV}) = 2(50\% - \text{CV})$
For	$\text{MV} = 50\%,\ e = 25\%$
Therefore	$\text{MV} = 50\%$ when $\text{CV} = 25\%$

Thus, the controller output becomes 50 per cent when the measurement CV drops by 25 per cent, creating a 25 per cent error. For this case, in order to stop the level from dropping, the proportional controller had to drop CV to create a large enough error so the controller could make $F_i = F_0$.

Suppose we set $K_c = 4$, giving MV $= 4e$. Now, the error would only need to be 12.5 per cent for MV = 4(12.5%) = 50%. Logically, it would appear that the larger the controller gain, the smaller the error. In theory, if K_c is set to infinity then the error can be reduced to zero. The problem with this extrapolation is that the controller gain K_c, is multiplied by all the gains of the other elements to give the loop gain K_L. If K_c becomes large enough, then the loop gain will be greater than unity, thus causing the loop to become unstable. Because of this loop gain limit, there is a limit to how large the controller gain can be. However, there is another approach to reducing the error to zero. Suppose another term is added to the proportional controller equation:

$$\mathrm{MV} = K_c e + b \tag{4.3}$$

This additional term is called the bias b, and is simply defined as the output of the controller when the error is zero. Using the previous example again, let us set $K_c = 2$. Also, manually adjust CV = SP = 50%, $F_i = F_0 = 50\%$, and set $b = 50\%$. Now, when the controller is set to automatic, what will happen? Since CV is equal to SP, e is equal to zero and, hence, $K_c e = 2(0) = 0$. There is no proportional contribution to the output and the output MV is equal to the bias, which is 50 per cent (Equation 4.3). Since F_0 is equal to 50 per cent and MV is also equal to 50 per cent, the level will stay the same. In general, if the bias equals the load MV, then the error will always be zero.

Now, suppose F_0 becomes 75 per cent. In order to stop the level from dropping, MV must equal F_o, which in this case is 75 per cent. From Equation 4.3, MV = $2e + 50\% = 2(50\% - \mathrm{CV}) + 50\%$, and CV must drop to 37.5 per cent to make the output MV equal to 75 per cent. When MV is equal to the outflow, the level will stop dropping. The level could also be prevented from dropping if the outflow F_0 was decreased. Suppose F_0 is equal to 25 per cent. In this case, the level will stop rising when CV is equal to 62.5 per cent, since that gives MV = $2(50\% - \mathrm{CV}) + 50\% = 25\%$.

As mentioned previously, increasing K_c can decrease the error; but remember not to increase K_c such that it makes the loop unstable. There is a limit for each feedback control loop. If K_c has a value such that the loop gain $K_L = 1$, then the loop will oscillate with a period that is a function of the natural characteristics of the process. This is called the natural period τ_n. If K_c is adjusted such that the loop gain is equal to 0.5 and a change is made in F_0, then the response shown in Figure 4.4 could be expected.

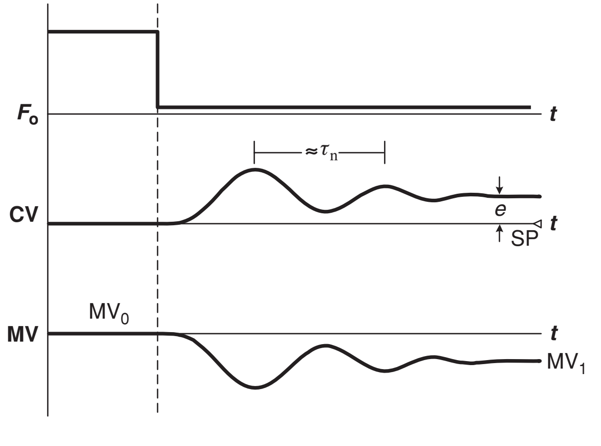

Figure 4.4 Typical proportional-only controller response

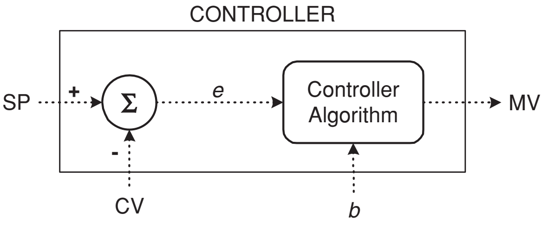

Figure 4.5 Block diagram of proportional-only controller

CV damps out with a quarter decay ratio (discussed in greater detail in Chapter 5) and a period approximately equal to the natural period. It then stabilizes with an offset that is a function of both the controller gain and the bias. The offset is the sustained error e, where CV does not return to the set point even when steady state is reached. This is a typical response for a loop under proportional-only control.

Now let us look again at Equation 4.3, and recall that the gain of any loop element is defined by

$$K = \frac{\Delta \text{ output}}{\Delta \text{ input}} \tag{4.4}$$

The block diagram of a proportional controller can be represented as shown in Figure 4.5.

The controller gain is the ratio of the change in controller output to the change in error. Hence, the gain of the proportional controller K_c is given by

$$K_c = \frac{\Delta \text{MV}}{\Delta e} \tag{4.5}$$

Since there is a one-to-one relationship between CV and e, the controller gain can be written as

$$K_c = \frac{\Delta \text{MV}}{\Delta \text{CV}} \tag{4.6}$$

The controller gain can also be defined as a change in controller output for a change in the process variable PV. This is true because the controlled variable CV is the transformed process variable from the transmitter to the controller. Therefore, CV is essentially PV, only in different units, i.e. percentage level instead of milliamperes.

Assuming that a linear relationship exists between CV and MV, as shown in Figure 4.6, Equation 4.6 may be written as

$$K_c = \frac{\Delta \text{MV}}{\Delta \text{CV}} = \frac{100\%}{\Delta \text{CV}} \tag{4.7}$$

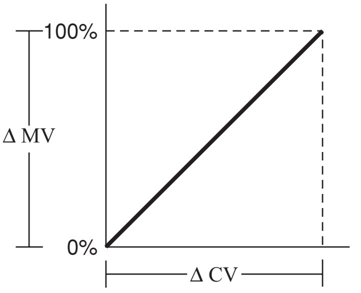

Figure 4.6 Controller input–output relationship

The controller gain K_c in Equation 4.7 is the amount that CV must change to make the controller output change by 100 per cent. The gain of the transmitter is similar and is given by

$$K_T = \frac{\Delta \text{out}}{\Delta \text{in}} = \frac{100\%}{\text{span}} \tag{4.8}$$

In other words, the input of the transmitter changes by the amount of the span to make the transmitter output change by 100 per cent. The span is the difference between the upper and lower values of the range.

The case of the controller is analogous to that of the transmitter, but instead of calling ΔCV the span, it is called PB, the proportional band. The proportional band is defined as the change in CV that will cause the output of the controller to change by 100 per cent. Using this definition of PB, we can be define the controller gain as

$$K_c = \frac{\Delta \text{MV}}{\Delta \text{CV}} = \frac{100\%}{\text{PB}\%} \tag{4.9}$$

If the proportional band setting on the controller is set to 40 per cent, the output of the transmitter, which is the input to the controller, changes over 40 per cent of its output span. The output of the controller would change by 100 per cent, or the controller gain K_c would be

$$K_c = \frac{100\%}{40\%} = 2.5$$

Virtually all modern controllers use a gain adjustment; however, a few older controllers exist that still use a proportional band adjustment. Remember that $K_c = 100\%/\text{PB}\%$, or as the PB gets larger, the gain gets smaller and vice versa. The equation for a proportional controller in terms of PB can be written as

$$\text{MV} = \frac{100}{\text{PB}} e + b \tag{4.10}$$

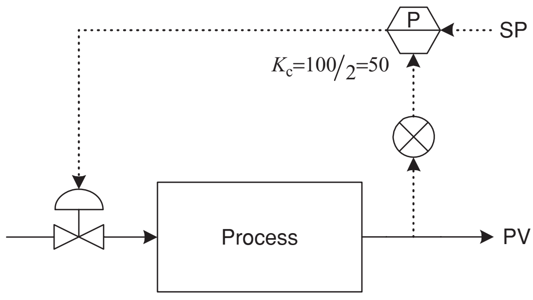

Figure 4.7 SISO feedback control loop

Note that:

$$e = \text{SP} - \text{CV} \text{ (for reverse acting)}$$

$$e = \text{CV} - \text{SP} \text{ (for direct acting)}$$

In order to make the error equal to zero, one of the following two possibilities must occur:

1 set PB = 0 ($K_c = \infty$)

2 set b = MV.

The first option, as discussed previously, is not plausible since as PB $\rightarrow$ 0, $K_c \rightarrow \infty$ and the loop becomes unstable. Furthermore, it is not possible to set PB = 0, because on many controllers the minimum setting is usually 2 to 5 per cent. However, if PB is very small, i.e. 2 per cent or $K_c = 50$, then the error would certainly be minimised, provided the loop remained stable. This case can be illustrated using Figure 4.7.

If in Figure 4.7 $K_v K_p K_T < 1/50$, then the loop would be stable since the loop gain $K_L < 1$ (Equation 2.6). If the process had a lower gain K_p, then a higher controller gain or smaller PB in the P-only controller could be used to minimize the error. One type of process where this is the case is a very large capacitance process, i.e. a large surge tank. Owing to the low process gain, a P-only controller is often used for level control.

The second option to make the error zero is to set the bias equal to the controller output MV. Some controllers have an adjustable bias, hence making this option viable, as in Equation 4.11:

$$e = \frac{1}{K_c}(\text{MV} - b) \tag{4.11}$$

However, this approach is only an option for processes that experience few load upsets, since a manual readjustment of the bias is required each time there is a load upset. There would be no error as long as the bias was equal to the load. Hence, if the process had infrequent load upsets, the operators could readjust the bias to give zero error, and it would be possible to use a P-only controller.

In general, a proportional controller provides a fast response when compared with other controllers, but a sustained error occurs where the PV does not return to the set point even when steady state is reached. This sustained error is called offset and is undesirable in most cases. Therefore, it is necessary to eliminate offset by combining proportional control with one of the other basic control modes.

4.3 Integral (I-only) control

The action of *integral control* is to remove any error that may exist. As long as there is an error present, the output of this control mode continues to move the FCE in a direction to eliminate the error. The equation for integral control is

$$\mathrm{MV} = \frac{1}{T_\mathrm{i}} \int e \, \mathrm{d}t + \mathrm{MV}_0 \tag{4.12}$$

MV_0 is defined as either the controller output before integration, the initial condition at time zero, or the condition when the controller is switched into automatic. The block diagram for an integral-only controller is given in Figure 4.8.

The action or response of the integral control algorithm for a given error is shown in Figure 4.9, assuming increase/decrease action.

If the measurement CV is increased in a step-wise fashion at time t_1 and then returned to the set point at t_2, the output would ramp up over the interval $t_1 < t < t_2$ since the controller is in effect integrating the step input. When the measurement is returned to the set point at $t = t_2$, the output would hold the value that the controller had integrated to, since the controller would think this was the correct value or the set point, i.e. $e = 0$.

The rate at which the controller output ramps is a function of two parameters: the integral time T_i, and the magnitude of the error. Obviously, the controller output MV would ramp in the opposite direction if CV had been moved below the set point.

The integral time T_i is defined as the amount of time it takes the controller output to change by an amount equal to the error. In other words, it is the amount of time required to duplicate the error. Thus, T_i is measured in minutes per repeat. Because of

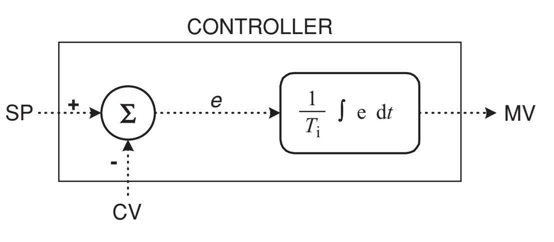

Figure 4.8 Block diagram of integral-only controller

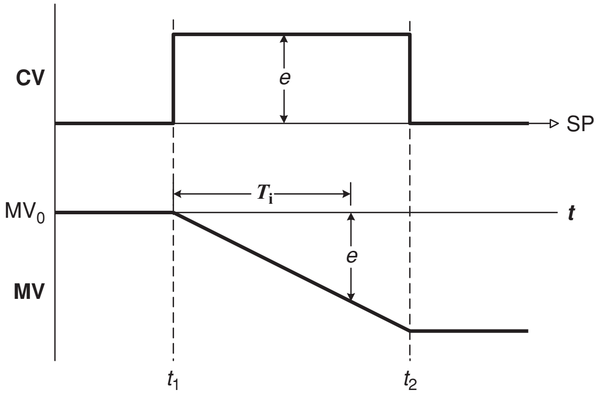

Figure 4.9 Integral controller response to square wave input

the form of Equation 4.12 some manufacturers measure the reciprocal of T_i or repeats per minute in a controller:

$$\frac{1}{T_i}[=]\left[\frac{1}{\text{min/repeat}}\right][=][\text{repeats/min}] \tag{4.13}$$

As a result of this reciprocal relationship, if the controller is adjustable in min/repeat, then increasing the adjustment gives less integral action, whereas in repeats/min, increasing the number produces greater integral action. Therefore, it is important to be aware of how an individual controller adjusts T_i. The rate of change of MV also depends on the magnitude of e as shown in Figure 4.10, in which T_i is fixed.

Figure 4.11 illustrates the responses of P-only, I-only and PI controllers to a step input. Although an integral-only controller provides the advantage of eliminating offset, there is a significant difference in its response time when compared with a proportional-only controller. As mentioned earlier, the output of the proportional-only controller changes

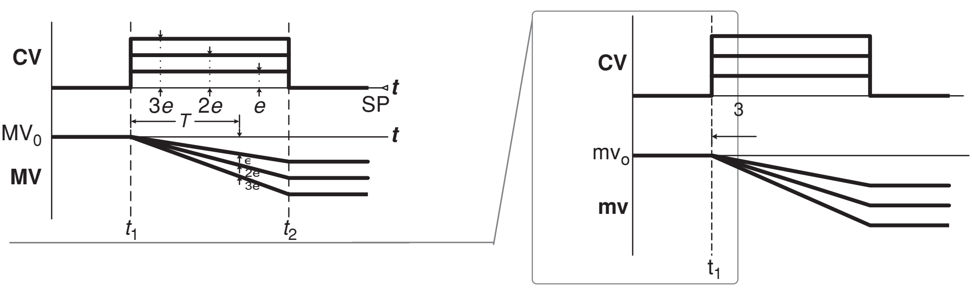

Figure 4.10 Effect of error magnitude on integral control response

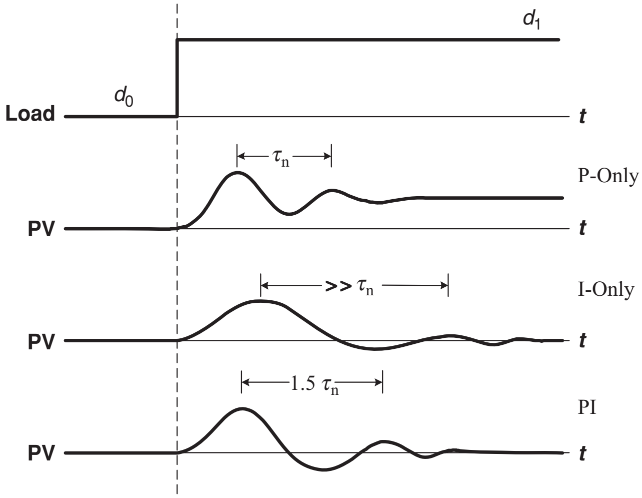

Figure 4.11 Response of P-only, I-only and PI controllers

as quickly as the measurement changes; in other words, the controller tracks the error. So, if the measurement changes as a step, then the controller output also changes as a step by an amount depending on the controller gain. For a step input to an integral controller the output does not change instantaneously, but rather by a rate that is affected by T_i and e.

Hence, integral-only control, due to the additional lag introduced by this mode, has an overall response that is much slower than that for proportional-only control. The period of response for the PV under integral-only control can be up to 10 times that for proportional-only; so a trade-off is made when using an I-only controller. If no offset is required, then a slower period of response must be tolerated. If the requirement is a return to the set point with no offset, and a faster response time is necessary, then the controller must be composed of both proportional and integral action.

As a result of the above, controllers with both proportional and integral action are more common. However, a few examples of integral-only controllers do occur. In the Claus sulphur plant air-demand controller [1] the trim air valve position error may be used to drive the main air valve position using only integral action. The combination of small (trim) and large (main) valves permits fine control of the Claus sulphur plant furnace air demand. In energy or 'BTU' control of a coal-fired power station [2] the integral-only control compares the energy leaving the boiler (steam) with the energy entering the boiler (coal). If there is a sustained difference, then the integral-only controller modifies the energy content of the coal feed until there is an energy balance. The integral time is very small, and is intended to compensate for the energy content of the coal supplied. Typically, it takes some 12 h for the BTU content of the coal to change a significant amount.

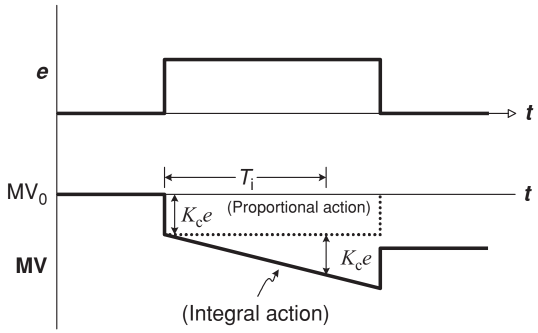

Figure 4.12 Proportional plus integral controller response to square wave input

4.4 Proportional plus integral (PI) control

A *proportional plus integral controller* will give a response period that is longer than a P-only controller but much shorter than an I-only controller. Typically, the response period of the process variable PV under PI control is approximately 50 per cent longer than for the P-only ($1.5\tau_n$, Figure 4.11). Since this response is much faster than I-only, and only somewhat longer than P-only control, the majority (>90 per cent) of controllers found in plants are PI controllers. The equation for a PI controller is

$$\mathrm{MV} = K_c\left(e + \frac{1}{T_i}\int e \, dt\right) = K_c e + K_c\frac{1}{T_i}\int e \, dt \tag{4.14}$$

The PI controller gain has an effect not only on the error, but also on the integral action. When we compare the equation for a PI controller (Equation 4.14) with that for a P-only controller (Equation 4.11) we see that the bias term in the P-only controller has been replaced by the integral term in the PI controller. Thus, the bias term for PI control is given by

$$b = K_c\frac{1}{T_i}\int e \, dt \tag{4.15}$$

Therefore, the integral action provides a bias that is automatically adjusted to eliminate any error. The PI controller is faster in response than the I-only controller because of the addition of the proportional action, as illustrated in Figure 4.12.

As shown in Figure 4.9, it takes T_i minutes for the output of the I-only controller to duplicate the error. With the addition of proportional action there is an immediate proportional step followed by integral action. The integral time in this case is defined as the amount of time it takes for the integral portion of the controller to replicate the proportional action. When the measurement is returned to the set point, the proportional action is lost, since $e = 0$, and the controller output is determined solely by integral action.

As can be seen from Equation 4.16, K_{PI}, the gain of the PI controller, is the sum of the two component gains. These component gains are proportional action K_P and integral action K_I.

$$K_{PI} = K_P + K_I = K_c + \frac{K_c}{T_i} \tag{4.16}$$

The K_c and T_i are used to adjust the PI controller gain to give the loop a desired response. Suppose $T_i = \infty$, which would result in $K_I = 0$, regardless of the value of K_c. In effect, the response would be that of a P-only controller with a period equal to τ_n and a sustained error. Although $T_i = \infty$ is not realizable, it can be set to a very large number in min/repeat to minimize the integral action.

Now, suppose T_i were set to a very small value. In this case, the PI controller gain would approach that of an integral-only controller, since $K_I \gg K_P$. The control action in the loop would now be that of an I-only controller with a return to the set point, but with a long response period.

These are two extremes, and somewhere in between is a T_i that will give a return to the set point with a reasonable response period of $1.5\tau_n$. The selection of T_i will be discussed in more detail under controller tuning in Chapter 5.

In general, starting with only proportional action, as more integral action is added, the PV begins to return to the set point. We only want enough integral gain to return to the set point, since a K_I greater than this will only serve to lengthen the response period. As more integral action is added by reducing T_i, we must compensate for the increased integral gain by reducing the proportional gain. Adjusting T_i will have an effect on K_I, and thus affects K_{PI}; this, in turn, affects both the damping and the response period. Adjusting K_c affects both K_I and K_P equally; thus K_c only has an effect on K_{PI}, affecting the damping and not the response period. These interacting effects will be considered in more detail under controller tuning in Chapter 5.

Although the response period of a loop under PI control is only 50 per cent longer than that for a loop under P-only control, this may in fact be far too long if τ_n is as large as 3 or 4 h. In order to increase the speed of the response, it may be necessary to add an additional control mode.

4.5 Derivative action

The purpose of *derivative action* is to provide lead to overcome lags in the loop. In other words, it anticipates where the process is going by looking at the rate of change of error $\mathrm{d}e/\mathrm{d}t$. For derivative action, the output equals the derivative time T_d multiplied by the derivative of the input, which is the rate of change of error:

$$\text{output} = T_d \frac{\mathrm{d}e}{\mathrm{d}t} \tag{4.17}$$

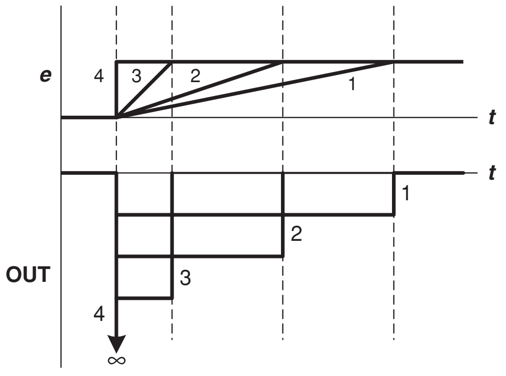

Figure 4.13 Effect of error on derivative mode output

Figure 4.13 shows how the output from a derivative block would vary for different inputs given a fixed value of T_d.

As the rate of change of the input gets larger, the output gets larger. Since the slope of each of these input signals is constant, the output for each of these rate inputs will also be constant. However, what happens as the slope approaches infinity as in the case of a step change, (4) in Figure 4.13. Theoretically, the output should be a pulse that is of infinite amplitude and zero time long. This output is unrealizable, since a perfect step with zero rise time is physically impossible, but signals that have short rise and fall times do occur. These types of signal are referred to as noise. Thus, the output from the derivative block would be a series of positive and negative pulses, which would try to drive the FCE either full open or full close. This would result in accelerated wear on the FCE and no useful control.

Consider a temperature measurement with a small-amplitude and high-frequency noise. One might assume that since the noise is of such small amplitude in comparison with the average temperature signal that a controller would not even notice it. This is only the case if the controller does not have derivative action. If the controller contains derivative action, then the temperature signal would be completely masked by the noise into the derivative mode of the controller, and the controller output would be a series of large-amplitude pulses, entirely masking any output contributed by the other control modes. Fortunately, in a case such as this the noise is either easily filtered out or is eliminated by modifying the installation of the primary sensor.

However, there are cases where noise is inherent in the measurement of PV and the rise and fall time of the noise is of the same magnitude as that of the measurement itself. In such a case, noise filtering would only serve to degrade the accuracy of the measurement of PV. A good example of a situation like this is a flow control loop. Flow measurement by its very nature is noisy; therefore, derivative action cannot be successfully applied.

It is important to note that derivative control would never be the sole control mode used in a controller. The derivative action does not know what the set point actually is and hence cannot control to a desired set point. Derivative action only knows that the error is changing.

4.6 Proportional plus derivative (PD) controller

The minimum controller configuration containing derivative action is the combination of *proportional plus derivative action* shown in Equation 4.18. This combination is not used very often and is primarily applied in batch pH control loops. However, it will help in the definition of derivative time T_{d}.

$$\mathrm{MV} = K_{\mathrm{c}}\left(e + T_{\mathrm{d}}\frac{\mathrm{d}e}{\mathrm{d}t}\right) + b \tag{4.18}$$

In Equation 4.18, the PD controller equation contains a bias term. A bias term will normally appear in any controller algorithm that does not contain integral action. This bias term does not appear when integral action is present, since integral action is in effect an automatic adjustment of bias. As with the PI controller, the proportional gain acts on the error as well as the derivative time T_{d}. Figure 4.14 shows the controller output MV for a typical input e test signal for the proportional and derivative portions of a PD controller.

In Figure 4.14, MV_{P} is the proportional portion of the output and MV_{D} is the derivative portion. In the example, the measurement changes at a fixed rate of change; therefore, the derivative portion of the output is constant and depends on the rate of change, the derivative time T_{d}, and proportional gain K_{c}. This dependency is evident from Equation 4.18. The proportional output is a ramp whose slope is a function of the proportional controller gain K_{c}.

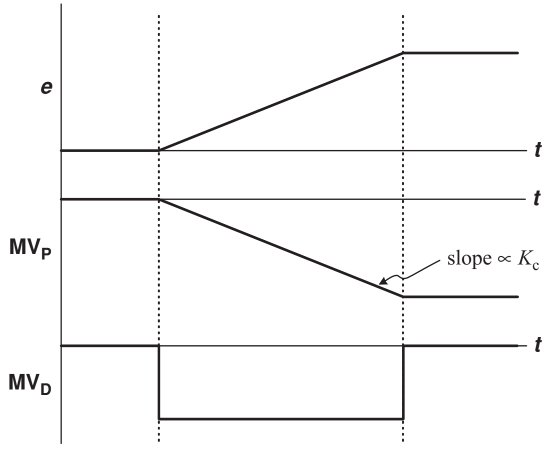

Figure 4.14 Responses for P-only and D-only portions of a PD controller

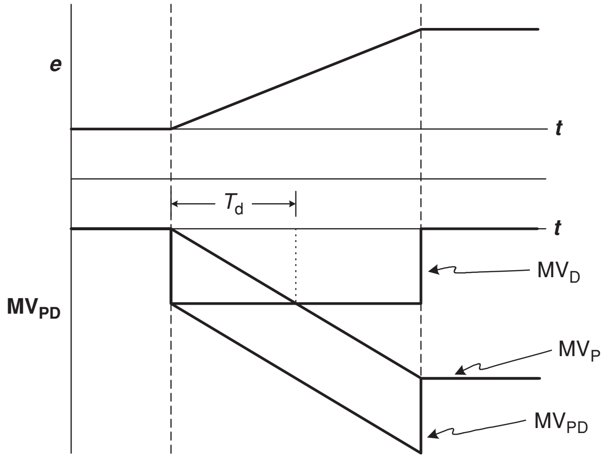

Figure 4.15 Combined response of a PD controller

Now, let us superimpose MV_P and MV_D to get the actual output for a PD controller, as shown in Figure 4.15.

For a ramp input it takes a period of time for the proportional action to reach the same level as the derivative action. This period of time is called the derivative time T_d and is measured in minutes. Increasing the derivative time T_d increases MV_D, or the contribution of the derivative action to the movement of the final control element.

In Equation 4.18, for the PD controller the derivative action acts on the error. Since $e = \mathrm{SP} - \mathrm{CV}$ for I/D action, de/dt is a function of both the derivative of the set point dSP/dt, and the derivative of the controlled variable dCV/dt:

$$\frac{\mathrm{d}e}{\mathrm{d}t} = \frac{\mathrm{dSP}}{\mathrm{d}t} - \frac{\mathrm{dCV}}{\mathrm{d}t} \tag{4.19}$$

If there is a load upset to the process, then the process variable PV will change at some rate dCV/dt which will result in the error also changing at the same rate ($\mathrm{d}e/\mathrm{d}t = -\mathrm{dCV}/\mathrm{d}t$), assuming there is no set-point change. Now, if a set-point change of even a few per cent is made and if the set point is changed quickly, then dSP/dt can become very large. This would cause a large pulse to be generated at the output of the controller. To overcome this potential problem, the controller can be made so that the derivative mode simply ignores set-point changes, as shown in Equations 4.20–4.22.

$$\frac{\mathrm{d}e}{\mathrm{d}t} = \frac{\mathrm{dSP}}{\mathrm{d}t} - \frac{\mathrm{dCV}}{\mathrm{d}t} \tag{4.20}$$

Ignoring set-point changes gives

$$\frac{\mathrm{d}e}{\mathrm{d}t} = \frac{-\mathrm{dCV}}{\mathrm{d}t} \tag{4.21}$$

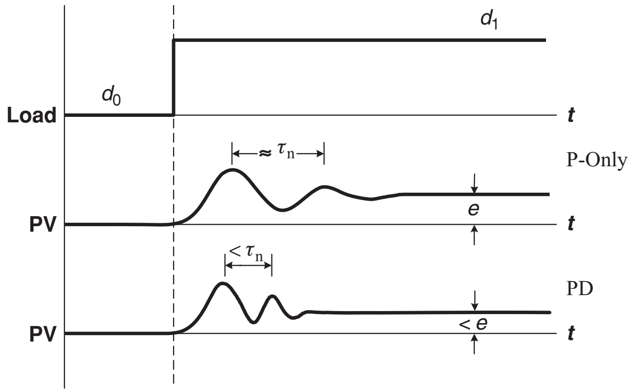

Figure 4.16 Proportional derivative controller response to a load disturbance

Hence

$$\mathrm{MV} = K_\mathrm{c}\left(e - T_\mathrm{d}\frac{\mathrm{dCV}}{\mathrm{d}t}\right) + b \tag{4.22}$$

In other words, there is no derivative action on a set-point change, only proportional action. On a load upset, both proportional and derivative actions are enabled. (Note also that, in some controller implementations, the proportional action is also decoupled from set-point changes, as the kick from a set-point change is also considered to be too aggressive.)

Figure 4.16 shows a comparison of the control-loop response to a load upset for both P-only and PD control. The response of the measurement PV under PD control is faster and results in a smaller offset than the loop under P-only control. This faster response is due to the addition of the derivative action.

In a PI controller, in order to minimize the integral action, T_i was made a large number. This makes the integral gain approach zero, and the controller then behaves essentially like a P-only controller. However, in the PD controller, even by setting T_d to a very small value, there is still the possibility of a sizeable derivative contribution if there is a noisy input, i.e. if $\mathrm{dCV}/\mathrm{d}t$ is large.

In electronic controllers and DCSs the derivative action can be eliminated by setting T_d to zero. In a pneumatic controller the derivative action cannot be eliminated, but it can be reduced to a minimum value of approximately 0.01 min. If a PD controller is installed on a flow loop there will still be considerable derivative action due to the noisy flow measurement. Therefore, it is important when applying a pneumatic controller to a noisy loop, such as a flow loop, to make certain the controller does not contain a derivative block.

The main reason for interest in derivative action is to combine it with proportional and integral action to produce a three-mode controller, a PID.

4.7 Proportional integral derivative (PID) control

The primary purpose of a *proportional integral derivative controller* (see Equation 4.23) is to provide a response period τ_n that is much the same as with proportional control but which has no offset. The derivative action adds the additional response speed required to overcome the lag in the response from the integral action.

$$\mathrm{MV} = K_c \left(e + \frac{1}{T_i} \int e \, dt - T_d \frac{\mathrm{CV}}{dt} \right) \tag{4.23}$$

Figure 4.17 presents a comparison of the responses for P-only, PI, and PID controllers to a step change in load.

The addition of the derivative mode in the PID controller provides a response similar to that of a P-only controller, but without the offset because of the integral action. Therefore, a PID controller provides a tight dynamic response, but since it contains a derivative block it cannot be used in any processes in which noise is anticipated.

4.7.1 Digital electronic controller forms

Controller algorithms are implemented in digital electronics using digital or 'discrete-time' forms of the analog or 'continuous-time' controller algorithms presented above. There are two basic digital electronic controller algorithms: the positional form and the velocity or differential form.

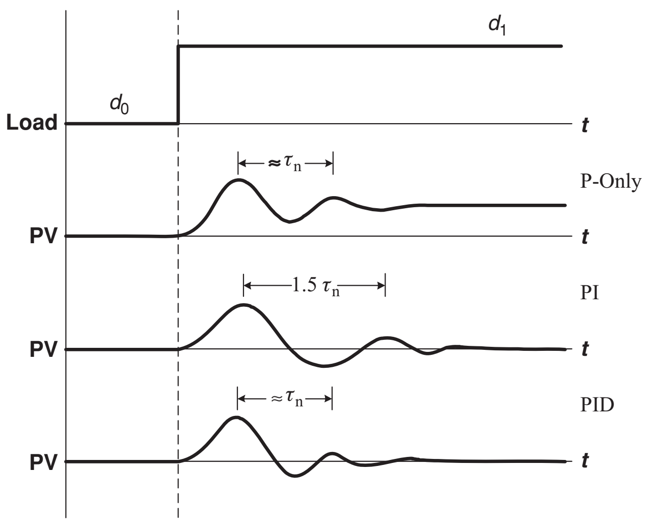

Figure 4.17 P-only, PI, and PID controller responses to a load disturbance

The positional form of the PID controller algorithm is

$$\mathrm{MV}(t) = K_c \left[e(t) - e(t-1) + \frac{1}{T_i} \sum_{k=0}^{n} e(kh) - T_d \frac{\mathrm{CV}(t) - \mathrm{CV}(t-1)}{h} \right] \tag{4.24}$$

where MV(t), $e(t)$ and CV(t) are the current controller output, error and controlled variables respectively, t is the enumerated sampling instant in time, MV(t – 1), $e(t$ – 1) and CV(t – 1) are the values of the controller output, error and controlled variables respectively one sampling period ago, and h is the sampling period.

The velocity or differential form of the PID controller algorithm is

$$\mathrm{MV}(t) = \mathrm{MV}(t-1) + K_c \left[e(t) - e(t-1) + \frac{1}{T_i} e(t) \right.$$
$$\left. - T_d \frac{\mathrm{CV}(t) - 2\mathrm{CV}(t-1) + \mathrm{CV}(t-2)}{h} \right] \tag{4.25}$$

For the positional form it is important to note how to handle the summation term associated with the integral action properly. The integral term in Equation 4.24 could grow to become a very large value if the output device was saturated and the CV was not able to return to the set point. For situations such as this, it is important to reset the value of the summation to ensure that the output of the algorithm will be equal to the (upper or lower) limit of the controller output. Then, when the set point is changed to a region where the controller can control effectively, the controller will respond without having to decrease the summation term from a value that has grown way beyond the upper or lower limit of the output. This automatic resetting of the controller integral term is commonly called anti-reset windup.

The velocity or differential form does not suffer from reset windup and is, therefore, the preferred form of controller equation when integral action is required. However, the positional form is preferred when there is no integral term, because this is the fail-safe form (in that in the advent of a failure the controller output will fail fully open or closed depending on the design) – whereas the failure mode for the velocity or differential form is the last value of the output.

4.8 Choosing the correct controller

Now that the various basic control modes have been described, it is desirable to be able to choose a particular control mode for a specific process. Figure 4.18 graphically outlines a procedure for control mode selection.

Starting at the top of the flow diagram, the first decision block asks the question: ‘Can offset be tolerated?’ If the answer is yes, then a proportional-only controller can be used. If the answer is no, then proceed to the next block, which asks: ‘Is there noise present?’ If there is noise, then use a PI controller. If there is no noise, then proceed

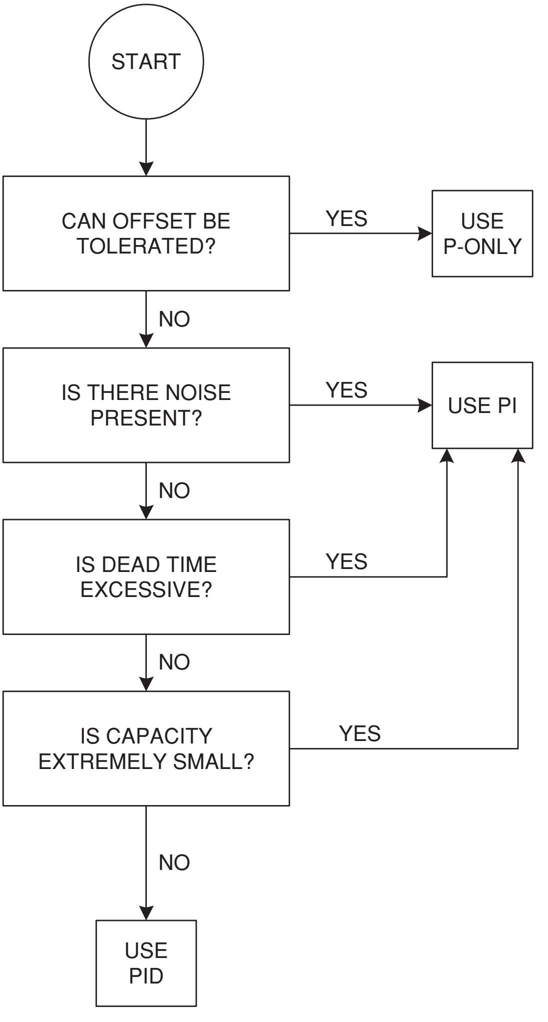

Figure 4.18 Flow chart for controller selection

to the next block, which asks: 'Is dead time excessive?' If the ratio of the dead time to the process time constant is greater than 0.5, then the process can be assumed to be dead-time dominant and requires a PI controller. If the process has no excessive dead time, then the next block asks: 'Is the capacitance extremely small?' If the answer is yes, then a PI controller can be used. A process with a short dead time and small capacitance does not require derivative action to speed up the response since it is already fast enough, as is the case for a flow loop. In this instance we might even consider an I-only controller, since the loop is so fast that slowing down the response through the use of integral-only action will still provide a fast enough response for the majority

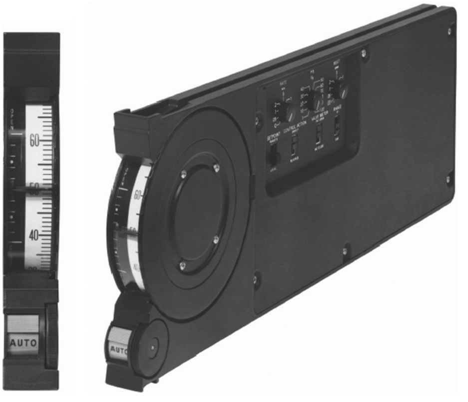

Figure 4.19 Electronic analog controller (reproduced by permission of Emerson Process Management)

of applications in the fluid processing industries. Finally, if the process capacitance is large, then a PID controller can be used.

It was mentioned earlier that the PI controller is the most common controller found in the plant. Looking at this flow chart one can see why. There are three possible paths to the PI controller, whereas there are four decision blocks that must be passed through to reach a PID controller.

4.9 Controller hardware

Now that we have covered how the controller works it is necessary to discuss controller hardware. Figures 4.19 and 4.20 are examples of single-loop stand alone controllers. Figure 4.19 is an electronic analog controller from the 1970s.

Figure 4.20 is a more recent digital version of the electronic controller from the late 1980s. It contains additional functionality, such as alarm limits for process value, deviation output signal and set point, set point ramping, auto- or self-tuning, signal filter time constant adjustment, start up values, on/off modality, and gain schedule limits.

Figure 4.21 shows a screenshot from a modern DCS. Simply put, a DCS is an electronic digital control system where computers spread functionality over multiple processes in large-scale plants. The advantage of a DCS is that it allows operators to monitor and control entire plants from a central control room.

DCSs were introduced in the mid 1970s with the advent of the microcomputer. DCSs enabled more flexible and complex control, monitoring, alarming, and historic data trending than local, single-loop control, or the centralized control previously possible with mini-computers.

Modern DCSs feature the use of digital, multi-drop communications that can interconnect sensors, actuators and the control room. Control can be allocated to digital

Figure 4.20 Electronic digital controller: Fisher DPR series (reproduced by permission of Emerson Process Management)

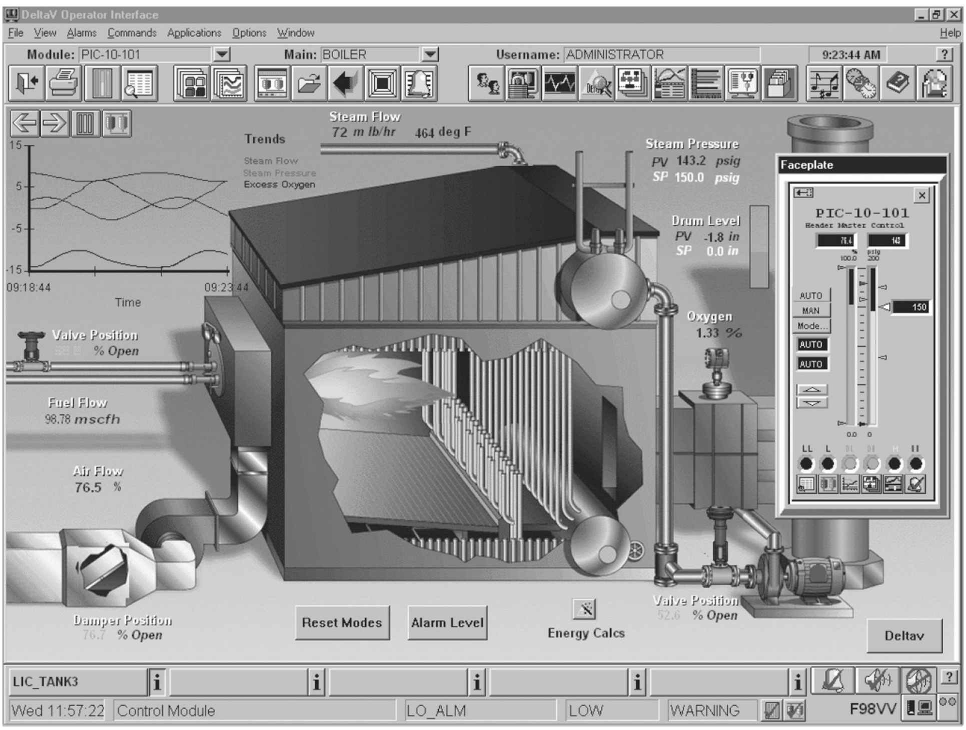

Figure 4.21 Screenshot from a Delta V distributed control system (reproduced by permission of Emerson Process Management)

devices that can communicate directly with each other, fully exploiting each other's capabilities with remote diagnostics and supervisory control and data acquisition (SCADA). This control technology is known as Fieldbus [3,4].

The PC explosion of the 1980s and 1990s has also impacted modern DCSs with the advent of PC-based control systems [5,6] which feature object linking and embedding (OLE) software for process control (OPC) [7].

4.10 References

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5

Tuning feedback controllers

There is no absolute right way or, for that matter, absolute wrong way to tune a controller. Controller settings depend on what the engineer/operator deems to be good performance in terms of the desired response to process upsets. The type of process, the process gain, the time constant, and dead time all play a role in determining the controller settings.

The settings also depend on the anticipated types of disturbance that the process will encounter. A controller would be tuned differently for stability due to set-point changes (servo control) than for load disturbances (regulatory control). In process control systems, load disturbances are most frequently encountered; hence, most systems are optimised for regulatory control.

This chapter will discuss control quality and optimisation, including the performance criteria that need to be considered when tuning a controller. A number of methods that can be used to determine controller settings in order to achieve the desired control are also described.

5.1 Quality of control and optimisation

Controller tuning can be defined as an optimisation process that involves a performance criterion related to the form of controller response and to the error between the process variable and the set point. When tuning a controller, some of the questions that may be asked include:

- Can offset be tolerated?
- Is no overshoot desired?
- Is a certain decay ratio required?
- Is a fast rise time needed?

These questions address some of the performance criteria that are used in the tuning of a controller, including overshoot, decay ratio, and error performance.

A Real-Time Approach to Process Control, Second Edition W. Y. Svrcek, D. P. Mahoney and B. R. Young
 ISBN: 978-0-470-02533-8

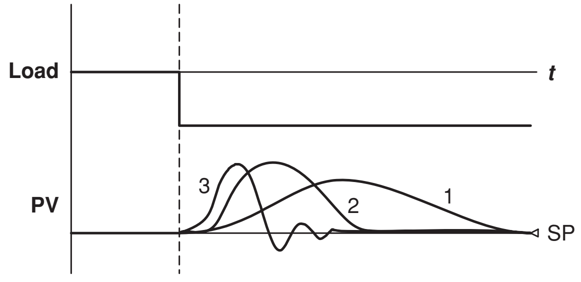

Figure 5.1 Typical responses to a load change

5.1.1 Controller response

Depending on the process to be controlled, the first consideration is to decide what type of response is optimal, or at least acceptable. Typical process responses to a load change are illustrated in Figure 5.1.

The three possible general extremes of response that exist, as shown in Figure 5.1, are:

1 overdamped: slow response with no oscillation;

2 critically damped: fastest response without oscillation;

3 underdamped: fast return to set point but with considerable oscillation.

From these three general extremes, we can see that selection of good control is a trade off between the speed of response and deviation from the set point. A highly tuned controller may become unstable if large disturbances occur, whereas a sluggishly tuned controller provides poor performance but is very robust. What is typically required for most process control loops is a compromise between performance and robustness.

When examining the response, there are several common performance criteria that can be used for controller tuning, which are based on characteristics of the system's closed-loop response. Some of the more common criteria include overshoot, offset, rise time and decay ratio. Of these simple performance criteria, control practitioners most often use decay ratio.

Cyclic radian frequency

The cyclic radian frequency ω is defined as

$$\omega = 2\pi f \tag{5.1}$$

and

$$f = \frac{1}{\text{period}} \tag{5.2}$$

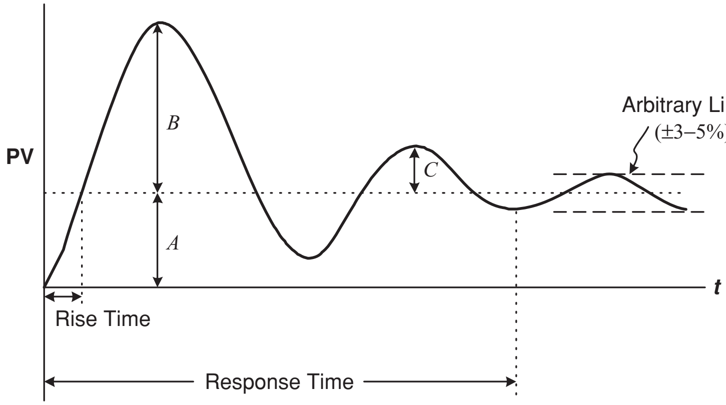

Figure 5.2 Second- or higher-order typical response to a set-point change

If Equation 5.2 is substituted into Equation 5.1, we obtain

$$\omega = \frac{2\pi}{\text{period}} \tag{5.3}$$

The cyclic radian frequency can also be related to the undamped natural frequency ω_n and the damping coefficient ξ.

$$\omega = \omega_n\sqrt{1-\xi^2} \tag{5.4}$$

Overshoot

Overshoot is the amount the response exceeds the steady-state final value. Referring to Figure 5.2, the overshoot can be defined as

$$\frac{B}{A} = e^{-\pi\xi/\sqrt{1-\xi^2}} \tag{5.5}$$

Decay ratio

The decay ratio is the ratio of the amplitude of an oscillation to the amplitude of the proceeding oscillation, C/B in Figure 5.2. More specifically, we can define the quarter decay ratio (QDR), which lies between critical damping and underdamping:

$$\text{QDR} = \frac{C}{B} = \frac{1}{4} \tag{5.6}$$

The decay ratio is often used to establish whether the controller as tuned is providing a satisfactory response. The QDR has been shown through experience to provide a good trade off between minimum deviation from the set point after an upset and the fastest return to the set point. The penalty of QDR is that some oscillation does occur. For a second-order system it can be shown that

$$\frac{C}{B} = \mathrm{e}^{-2\pi\xi/\sqrt{1-\xi^2}} \tag{5.7}$$

Rise time

The rise time is the initial time required by the transient response to reach the final steady-state value.

Response time

The response time is the time required for the response to settle within the specified arbitrary limits. These limits are typically set at ±3–5 per cent of the PV steady-state value.

5.1.2 Error performance criteria

The previously discussed simple performance criteria, i.e. decay ratio, overshoot, etc., use only a few points in the response and therefore are simple to use. On the other hand, error performance criteria are based on the entire response of the process but they are also more complicated.

Integrated error

The curve shown in Figure 5.3 represents the response of a loop due to a process upset. This graphical representation of the controlled variable's return to the set point is known

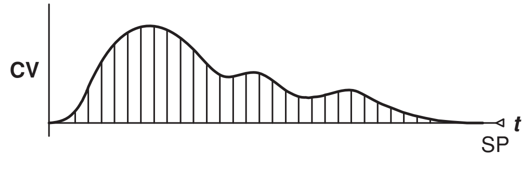

Figure 5.3 General response curve

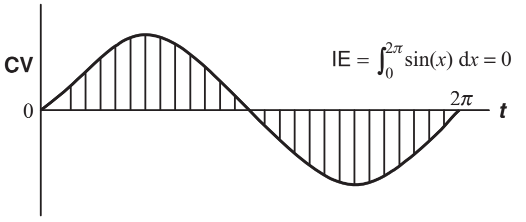

Figure 5.4 Sinusoidal error response

as a response curve. The integrated error (IE) is the area under the response curve, and the idea of using this as an error criterion is to attempt to minimise this area.

In mathematical terms, with e representing the error as a function of time, we can write

$$\mathrm{IE} = \int_0^{\infty} e \, \mathrm{d}t \tag{5.8}$$

It may not always be possible to minimise IE without paying a penalty in some respect. For example, underdamping produces a minimum area under the response curve but has considerable oscillation.

The method of integrated error may not be 100 per cent reliable if there is no averaging elsewhere in the process. For example, if there is a sinusoidal oscillation about the set point, then the positive and negative areas tend to cancel each other out over time, which presents a misleading conclusion, as shown in Figure 5.4. However, barring this situation, integrated error is a perfectly adequate error criterion.

Integrated absolute error

Integrated absolute error (IAE) essentially takes the absolute value of the error. Negative areas are accounted for when IAE is used, thus dismissing the problem encountered with IE regarding sinusoidal responses.

$$\mathrm{IAE} = \int_0^{\infty} |e| \, \mathrm{d}t \tag{5.9}$$

Integrated squared error

The integrated squared error (ISE) criterion uses the square of the error, thereby penalizing larger errors more than smaller errors. This gives a more conservative response, i.e. faster return to the set point.

$$\mathrm{ISE} = \int_0^{\infty} e^2 \, \mathrm{d}t \tag{5.10}$$

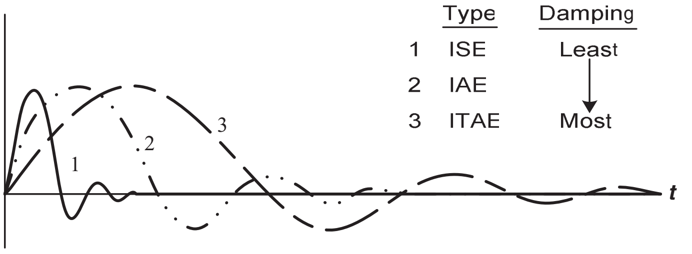

Figure 5.5 Response to various error criteria

Integrated time absolute error

The integrated time absolute error (ITAE) criterion is the integral of the absolute value of the error multiplied by time. ITAE results in errors existing over time being penalised even though they may be small, which results in a more heavily damped response.

$$\mathrm{ITAE} = \int_0^{\infty} t\ |e|\ \mathrm{d}t \tag{5.11}$$

Figure 5.5 shows the various responses of a loop that is tuned to the above criteria, including those for IAE and ISE.

5.2 Tuning methods

The following presents a very brief description of some of the various accepted methods used for controller tuning. In each case the suggested controller settings are optimised for a particular error performance criterion, often QDR. The first method described is based entirely on trial and error, whereas the rest are based upon some understanding of the physical nature of the process to be controlled. This understanding is generated from either open- or closed-loop process testing.

5.2.1 'Trial-and-error' method

As the name suggests, tuning by trial and error is simply a guess-and-check type method. The following is a list of practical suggestions for tuning a controller by trial and error. These suggestions are also useful for fine tuning controllers tuned by other methods.

1 Proportional action is the main control. Integral and derivative actions are used to trim the proportional response.

2 The starting point for trial-and-error tuning is always with the controller gain, integral action, and derivative action all at a minimum.

3 Make adjustments in the controller gain by using a factor of two, i.e. 0.25, 0.5, 1.0, 2.0, 4.0, etc.

4 The optimal response is the QDR.

5 When in trouble, decrease the integral and derivative actions to a minimum and adjust the controller gain for stability.

Rules of thumb

The following rules of thumb should not be taken as gospel or as a methodology; rather, they are intended to indicate typical values encountered. As such, these rules can be useful when tuning a controller using the trial-and-error method. However, it is important to remember that controller parameters are strongly dependent upon the individual process and may not always abide by the rules outlined below.

Flow When dealing with a flow loop, P-only control can be used with a low controller gain. For accuracy, PI control is used with a low controller gain and high integral action. Derivative action is not used, because flow loops typically have very fast dynamics and flow measurement is inherently noisy.

$K_c = 0.4–0.65$
$T_i = 0.1\text{min (or 6 s)}$

Level Levels represent material inventory that can be used as surge capacity to dampen disturbances. Hence, loosely tuned P-only control is sometimes used. However, most operators do not like offset, so PI level controllers are typically used.

The following P-only settings ensure that the control valve will be 50 per cent open when the level is at 50 per cent, wide open when the level is at 75 per cent, and shut when the level is at 25 per cent:

$K_c = 2$
Bias term $b = 50\%$
Set point SP $= 50\%$

Typical PI controller settings are:

$K_c = 2–20$
$T_i = 1–5\text{ min}$

Pressure Pressure control loops show large variation in tuning, depending on the dynamics of the pressure response. Typical ranges are as follows:

Vapour $K_c = 2\text{–}10$
$T_i = 2\text{–}10$ min
Liquid $K_c = 0.5\text{–}2$
$T_i = 0.1\text{–}0.25$ min

Temperature Temperature dynamic responses are usually fairly slow, so PID control is used. Typical parameter values are:

$K_c = 2\text{–}10$
$T_i = 2\text{–}10$ min
$T_d = 0\text{–}5$ min

5.2.2 Process reaction curve methods

In the process reaction curve methods a process reaction curve is generated in response to a disturbance. This process curve is then used to calculate the controller gain, integral time and derivative time. These methods are performed in open loop, so no control action occurs and the process response can be isolated.

To generate a process reaction curve, the process is allowed to reach steady state or as close to steady state as possible. Then, in open loop, so that there is no control action, a small disturbance is introduced and the reaction of the process variable is recorded. Figure 5.6 shows a typical process reaction curve generated using the above method for a generic self-regulating process. The term self-regulating refers to a process where the controlled variable eventually returns to a stable value or levels out without external intervention. The process parameters that may be obtained from this process reaction curve for the initial step disturbance P (%) are as follows: the lag time L (min), the time constant T (min), the change in PV in response to step disturbance ΔC_p (%), i.e. (change in PV)/(PV span) $\times$ 100, the reaction rate N (% min^{-1})

$$N = \frac{\Delta C_p}{T}$$

and the lag ratio R (dimensionless)

$$R = \frac{L}{T} = \frac{NL}{\Delta C_p}$$

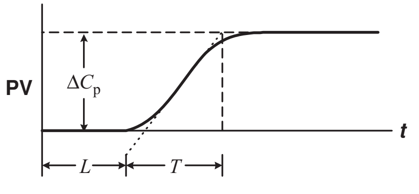

Figure 5.6 Process reaction curve

Methods of process analysis with forcing functions other than a step input are possible, and include pulses, ramps, and sinusoids. However, step function analysis is the most common, as it is the easiest to implement.

Ziegler–Nichols open-loop rules

In 1942, Ziegler and Nichols [1] changed controller tuning from an art to a science by developing their open-loop step function analysis technique. They also developed a closed-loop technique, which is described in the next section on constant cycling methods.

The Ziegler–Nichols open loop recommended controller settings for the QDR are as follows:

$$\text{P-only} \quad K_c = \frac{P}{NL} \tag{5.12}$$

$$\text{PI} \quad K_c = 0.9\frac{P}{NL} \tag{5.13}$$

$$T_i = 3.33L \tag{5.14}$$

$$\text{PID} \quad K_c = 1.2\frac{P}{NL} \tag{5.15}$$

$$T_i = 2.0L \tag{5.16}$$

$$T_d = 0.5L \tag{5.17}$$

These settings should be taken as recommendations only and tested thoroughly in closed loop, with fine tuning of the parameters to obtain the QDR.

Cohen–Coon open-loop rules

In 1953, Cohen and Coon [2] developed a set of controller tuning recommendations that correct for one deficiency in the Ziegler–Nichols open-loop rules. This deficiency is the sluggish closed-loop response given by the Ziegler–Nichols rules on the relatively rare occasion when process dead time is large relative to the dominant open-loop time constant.

The Cohen–Coon recommended controller settings are as follows:

$$\text{P-only} \quad K_c = \frac{P}{NL}\left(1 + \frac{R}{3}\right) \tag{5.18}$$

$$\text{PI} \quad K_c = \frac{P}{NL}\left(0.9 + \frac{R}{12}\right) \tag{5.19}$$

$$T_{\mathrm{i}} = L\,\frac{30+3R}{9+20R} \tag{5.20}$$

$$\text{PID} \quad K_{\mathrm{c}} = \frac{P}{NL}\left(1.33 + \frac{R}{4}\right) \tag{5.21}$$

$$T_{\mathrm{i}} = L\,\frac{32+6R}{13+8R} \tag{5.22}$$

$$T_{\mathrm{d}} = L\,\frac{4}{11+2R} \tag{5.23}$$

As with the Ziegler–Nichols open-loop method recommendations, the Cohen–Coon values should be implemented and tested in closed loop and adjusted accordingly to achieve the QDR.

Internal model control tuning rules

Many practitioners have found that the Ziegler–Nichols open loop and Cohen–Coon rules are too aggressive for most chemical industry applications, since they give a large controller gain and short integral time. Rivera *et al*. [3] developed the internal model control (IMC) tuning rules with robustness in mind. The tuning parameter from the IMC method (the closed-loop speed of response) relates directly to the closed-loop time constant and the robustness of the control loop. As a consequence, the closed-loop step load response exhibits no oscillation or overshoot. Lambda tuning, e.g. [4], is a term that is also used to refer to controller tuning methods that are based on a specified closed-loop time constant.

Since the general IMC method is unnecessarily complicated for processes that are well approximated by first-order dead time or integrator dead-time models, simplified IMC rules were developed by Fruehauf *et al*. [5] for PID controller tuning (see Table 5.1).

Of course, these recommendations need to be tested in the closed-loop situation and the final settings arrived at through the use of fine tuning.

Table 5.1 Simplified IMC rules [4]

	$\frac{\tau}{L} > 3$	$\frac{\tau}{L} < 3$	$L < 0.5$
K_{c}	$\frac{P}{2NL}$	$\frac{P}{2NL}$	$\frac{P}{N}$
T_{i}	$5L$	τ	4
T_{d}	$\leq 0.5L$	$\leq 0.5L$	$\leq 0.5L$

5.2.3 Constant cycling methods

Ziegler–Nichols closed-loop method

The closed-loop technique of Ziegler and Nichols [6] is a technique that is commonly used to determine the two important system constants: ultimate period and ultimate gain. It was one of the first tuning techniques to be widely adopted.

When tuning using the Ziegler–Nichols closed-loop method, values for proportional, integral, and derivative controller parameters may be determined from the ultimate period and ultimate gain. These are determined by disturbing the closed-loop system and using the disturbance response to extract the values of these constants.

The following is a step-by-step approach to using the Ziegler–Nichols closed-loop method for controller tuning:

1. Attach a proportional-only controller with a low gain (no integral or derivative action).

2. Place the controller in automatic.

3. Increase the controller gain until a constant-amplitude limit cycle occurs.

4. Determine the ultimate period P_u and ultimate gain K_u from the constant-amplitude limit cycle (Figure 5.7):

$$\begin{aligned} P_u &= \text{period taken from limit cycle} \\ K_u &= \text{controller gain that produces the limit cycle} \end{aligned}$$

5. Calculate the tuning parameters using the following equations:

$$\text{P-only} \quad K_c = \frac{K_u}{2} \tag{5.24}$$

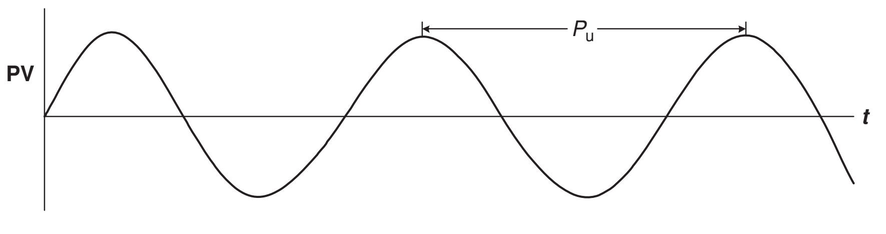

Figure 5.7 Constant-amplitude limit cycle

$$\text{PI} \quad K_c = \frac{K_u}{2.2} \tag{5.25}$$

$$T_i = \frac{P_u}{1.2} \tag{5.26}$$

$$\text{PID} \quad K_c = \frac{K_u}{1.7} \tag{5.27}$$

$$T_i = \frac{P_u}{2} \tag{5.28}$$

$$T_d = \frac{P_u}{8} \tag{5.29}$$

6 Fine tune by adjusting K_c, T_i, and T_d as required to find the QDR.

Auto-tune variation technique

The auto-tune variation (ATV) technique of Åström and Hagglund [6] is another closed-loop technique used to determine the two important system constants, i.e. the ultimate period and the ultimate gain. However, the ATV technique determines these system constants without unduly upsetting the process. Tuning values for proportional, integral and derivative controller parameters can be determined from these two constants. Here, we recommend the use of Tyreus–Luyben [7] settings for tuning that is suitable for chemical process unit operations. All methods for determining the ultimate period and ultimate gain involve disturbing the system and using the disturbance response to extract the values of these constants.

In the case of the ATV technique, a small limit-cycle disturbance is set up between the manipulated variable (controller output) and the controlled variable (process variable). Figure 5.8 shows the instrument set-up, and Figure 5.9 shows the typical ATV response plot with critical parameters defined. It is important to note that the ATV technique is

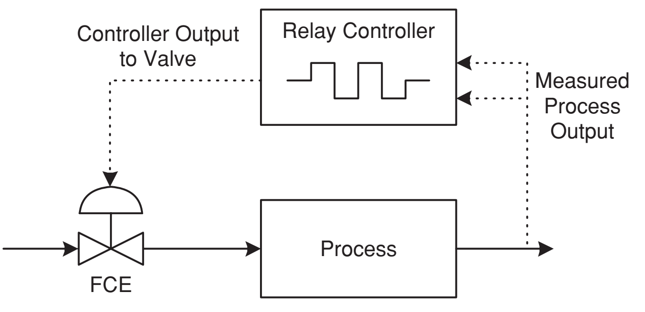

Figure 5.8 ATV tuning instrument set-up

applicable only to processes with significant dead time. The ultimate period will just equal the sampling period if the dead time is not significant. The general ATV tuning method for a PI controller is as follows:

1. Determine a *reasonable* value for valve change h (typically 0.05, i.e. 5 per cent). The value for h should be small enough that the process is not unnecessarily upset, but large enough that the amplitude a can be measured.
2. Move the valve $+h$ units.
3. Wait until the process variable starts to move, then move the valve $-2h$ units.
4. When the process variable (PV) crosses the set point, move the valve $+2h$ units.
5. Repeat until a limit cycle is established, as illustrated in Figure 5.9.
6. Record the value of a by picking it off the response graph.
7. Perform the following calculations to determine the ultimate period P_u, ultimate gain K_u, and the controller gain K_c and integral time T_i:

$$P_u = \text{period taken from limit cycle}$$
$$K_u = \frac{4h}{3.14a}$$

$$K_c = \frac{K_u}{3.2} \tag{5.30}$$

$$T_i = 2.2P_u \tag{5.31}$$

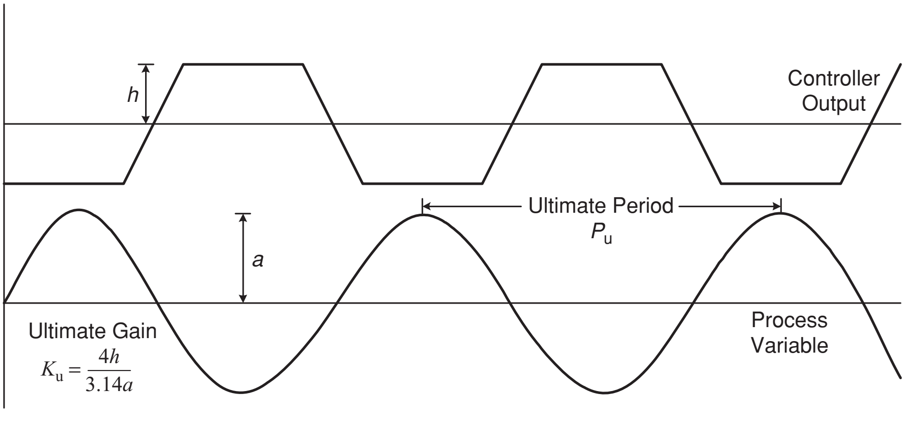

Figure 5.9 ATV critical parameters

Table 5.2 Tuning comparison

	Z–N CL	ATV	(Z–N/ATV) ratio
Controller gain K_c	$\frac{K_u}{2.2}$	$\frac{K_u}{3.2}$	1.45
Integral time T_i	$\frac{P_u}{1.2}$	$2.2\,P_u$	0.38

Comparison of ATV and Ziegler–Nichols closed-loop tuning techniques

Table 5.2 compares the tuning constants between the ATV and Ziegler–Nichols closed-loop (Z–N CL) tuning techniques. Notice that the Ziegler–Nichols tuning is more aggressive, with a larger controller gain and shorter integral time. This technique was originally developed for electromechanical systems control and is based on the more aggressive QDR criterion. ATV tuning was developed for fluid and thermal processes and emphasises minimizing overshoot. ATV, therefore, is often the preferred technique for process control.

5.3 References

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6

Advanced topics in classical automatic control

Up to this point the discussion has been restricted to feedback control loops, the most common control method used in process industries. However, there are a number of more complex control techniques that can provide improved and more economic process control that should also be considered [1]. Some of the control schemes that are discussed in this chapter include cascade, feedforward, ratio, and override control. These schemes are classified as 'advanced classical' [1] topics and are widely used in industry.

6.1 Cascade control

Cascade control is a common control technique that uses two controllers with one feedback loop nested inside the other [2–4]. The output of the primary controller acts as the set point for the secondary controller. The secondary controller controls the FCE. A typical cascade control loop is illustrated in Figure 6.1.

To understand cascade control better, we will examine a typical feedback control scheme and consider how it may be improved through the use of cascade control. Let us consider the feedback control loop for a heat exchanger shown in Figure 6.2.

In this example of a heat exchanger, the energy is provided by steam flow F_s, and is used to heat an entering fluid F_w from an inlet temperature T_1 to an outlet temperature T_2. If the controller is properly tuned and there is a change in F_w, then the change in T_2 will be sensed and the temperature controller will then change its output by repositioning the valve to bring the outlet temperature back to the set point. In other words, the controller moves the variance from one stream to another and thus mitigates changes in the process variable PV.

Let us now consider another possible type of upset: a supply side upset. If the steam F_s, is coming from a supply header that is also servicing other users, then there is a possibility that, as the other users' needs vary, pressure upsets will occur and cause changes in the steam supply F_s. Suppose that another user demanded steam quantities that caused a pressure drop in the header, thus resulting in a drop in steam flow to the exchanger. The only way this drop in F_s could be measured would be as a drop in the

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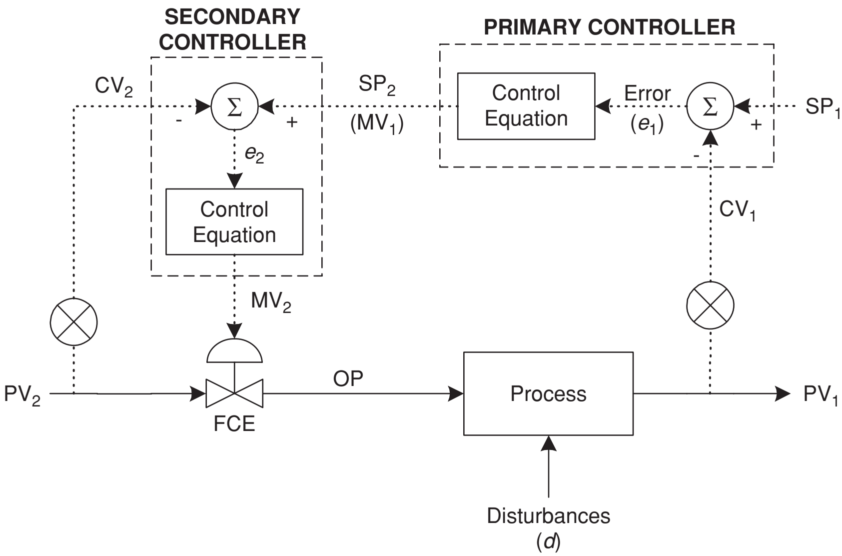

Figure 6.1 Cascade control system

outlet temperature T_2. This deviation from set point would be sensed and compensated for in the same manner as for the stream load upset F_w. The response of the process variable T_2 to the two situations is shown in Figure 6.3.

In both cases the process variable T_2 would dampen out in the period τ_n. However, for the supply upset situation the feedback control may cause the temperature to be in a constant state of flux. For instance, if the valve (when open to 50 per cent) supplies the amount of steam needed, then the outlet temperature will be at the set point. However, then the flow through the valve is a function of the pressure drop across it. Therefore, if

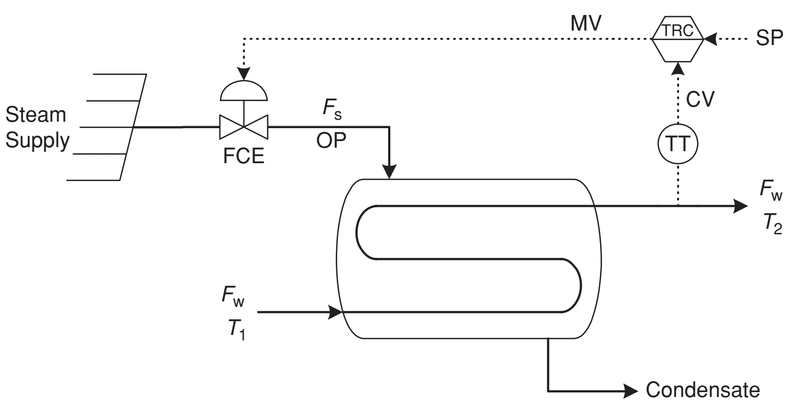

Figure 6.2 Feedback control loop for a heat exchanger (steam supply loop)

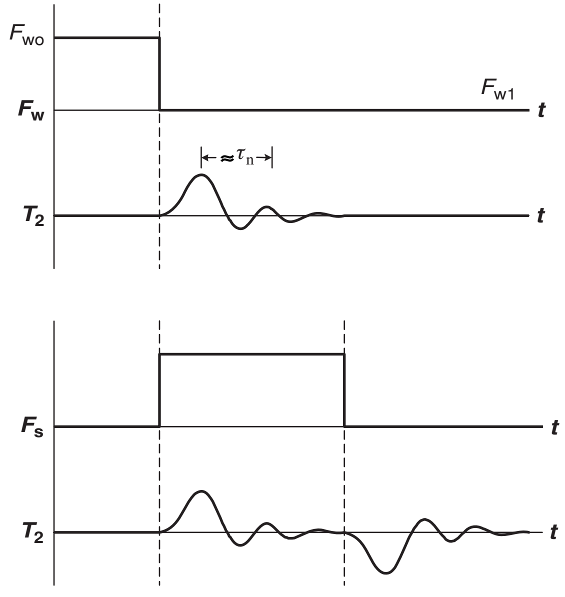

Figure 6.3 Heat exchanger outlet temperature response

there is a decrease in inlet pressure then the flow will decrease, even though the valve is 50 per cent open. This disturbance will propagate through the dead time and capacitive lag of the heat exchanger before it emerges as an outlet temperature deviation, at which time the necessary control action will be taken to bring the temperature back to the set point. The response will damp out with a period τ_n back to the set point. If τ_n were a minute or two or even longer, then the process might be in a constant state of flux and never settle to the desired set point.

The problem is that the controller output sets a valve opening rather than setting an energy supply requirement. With a constant pressure drop across the valve, the relationship between valve position and steam flow is constant, but if the pressure drop changes then this relationship changes. Thus, it is better to set the steam requirement rather than valve position. As long as the valve can supply the energy requirement, it does not matter how wide open the valve is, only how much energy is being delivered. Hence, for this process, better control can be achieved by using cascade control.

Now we will apply cascade control to the same heat exchanger to control supply-side upsets. In Figure 6.4, the temperature loop (primary loop or master) is used to adjust the set point of the flow control loop (secondary loop or slave).

With cascade control in place, if a set-point change is made in the temperature controller, or if a load upset occurs that changes F_w, then the output of the temperature controller will change the steam flow controller set point. The flow loop operates so much faster than the temperature loop that the temperature controller does not in fact know whether its output is going directly to a valve or as a set point to another controller. In general, the control loop closest to the controlled variable is called the

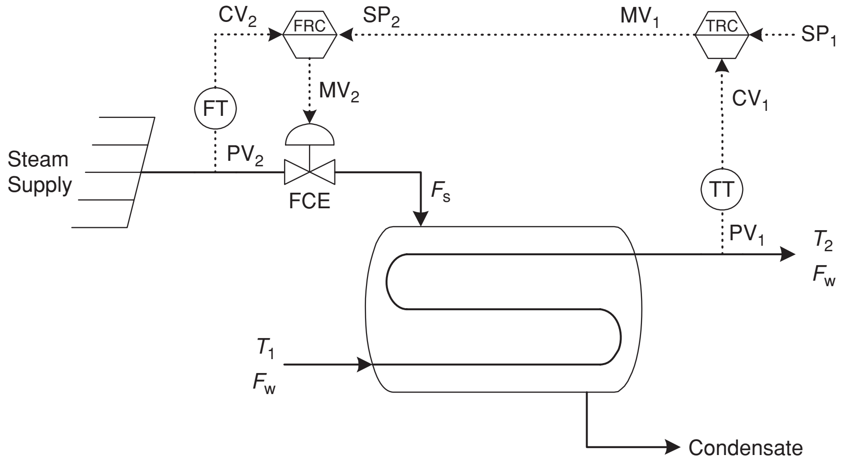

Figure 6.4 Typical cascade control loop for a heat exchanger

primary, outer, or master loop. The control loop closest to the supply to the process is called the secondary, inner, or slave loop. In the heat exchanger example given, the temperature loop is the primary loop and the flow loop is the secondary loop.

Both the primary and secondary loops have their own response period, independent of whether they are in a cascade configuration or not. The response period of the primary loop is τ_1 and that of the secondary loop is τ_2. In order for cascade control to work effectively, $\tau_1 \geq 4\tau_2$. What this rule of thumb implies is that the primary loop should never know that there is a secondary loop. The secondary loop should be able to respond as quickly as the FCE. If this rule is followed, then there will be little interaction between the two loops and the control scheme will function properly.

6.1.1 Starting up a cascade system

To put a cascade system into operation:

1 Place the primary controller in manual or the secondary controller to the local set point. This will break the cascade and allow the secondary controller to be tuned.

2 Tune the secondary controller as if it were the only control loop present.

3 Return the secondary controller to the remote set point and/or place the primary controller in auto.

4 Now tune the primary loop normally. If the system begins to oscillate when the primary controller is placed in auto, reduce the primary controller gain.

Note, when tuning the primary controller there should be no interaction between the primary and secondary loops. If there is, it means that the primary loop is not slow enough in comparison with the secondary.

One of the most common forms of cascade is the output of a primary controller acting as a set point to a valve positioner.

6.2 Feedforward control

One of the disadvantages to using feedback control is that a disturbance must propagate through the process before it is detected and action is taken to correct it. This type of control is sufficient for processes in which some deviation from the set point is acceptable. However, there are certain processes where this set-point deviation must be minimized. Feedforward control can accomplish this because it corrects and/or minimizes disturbances before they enter the process [3,4]. A typical feedforward control system is shown in Figure 6.5.

In its simplest form, a feedforward controller merely proportions the corrective action to the size of the disturbance. In other words, the control equation is merely a gain based on steady state, i.e. mass or energy balance at steady state. This does not take into account any of the process dynamics of the system. If there is a difference, or lag, in the speed of the process response to the control action when compared with that of the disturbance, then it may be necessary to introduce some dynamic compensation into the control equation. The dynamic compensation correctly times the control action and response, thus giving increased accuracy in the feedforward control.

In general, the feedforward dynamic elements will not be physically realizable. In other words, they cannot be implemented exactly. For instance, if the process disturbance measurement contains dead time, or lag, then the feedforward dynamic compensation would have to be a predictor, which of course is impossible unless an exact

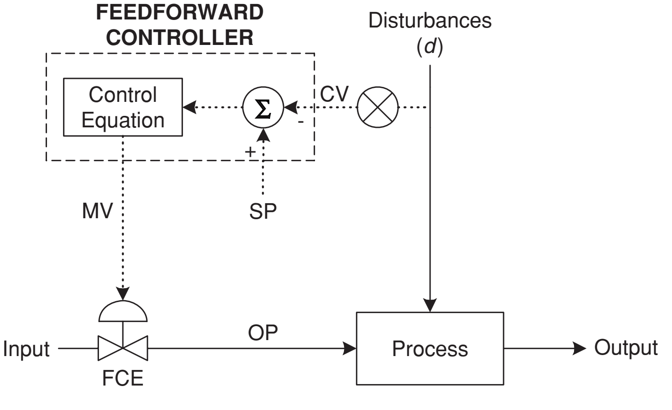

Figure 6.5 Feedforward control system

and very fast dynamic model of the process is available. In practice, the feedforward dynamic elements are approximated by a lead–lag network [3,4] that is adjusted to yield as much disturbance rejection as possible over as wide a range as possible.

When feedforward control is used, equations are needed to calculate the amount of the manipulated variable needed in order to compensate for the disturbance. This sounds simple enough; however, the equations must incorporate an understanding of the exact effect of the disturbances on the process variable. Therefore, one disadvantage of feedforward control is that the controllers often require sophisticated calculations, as even steady models can be nonlinear and thus need more technical and engineering expertise in their implementation.

Another disadvantage of feedforward control is that all of the possible disturbances and their effects on the process must be known precisely. If unexpected disturbances enter the process when only feedforward control is used, then no corrective action is taken and the errors will build up in the system. If all the disturbances were measurable and their effects on the process precisely known, then a feedback control system for regulatory purposes would not be needed. However, such complete and error-free knowledge is never available, so feedforward is generally combined with feedback, as illustrated in Figure 6.6. The intent of this union is that the feedforward mitigates most of the effects of the principal disturbances and the feedback loops provide residual control and set point tracking.

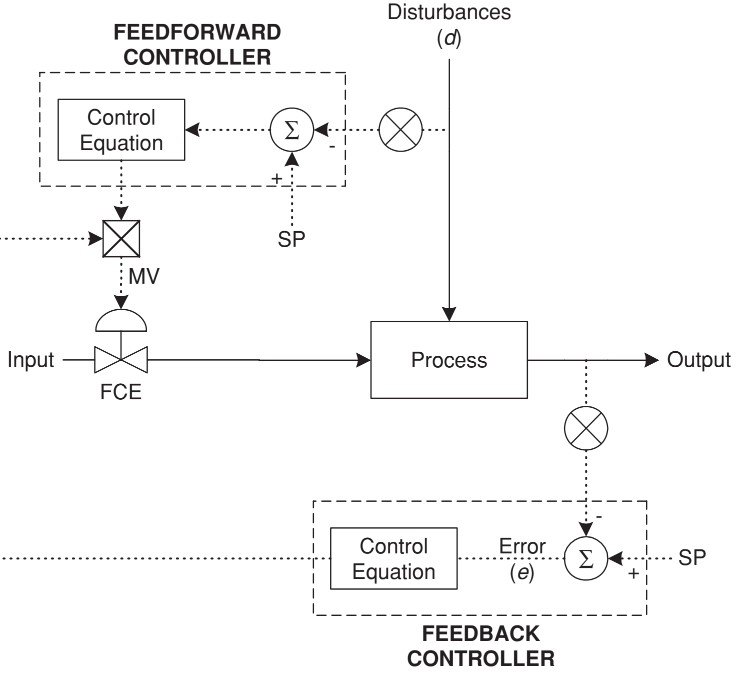

Figure 6.6 Feedforward/feedback control system

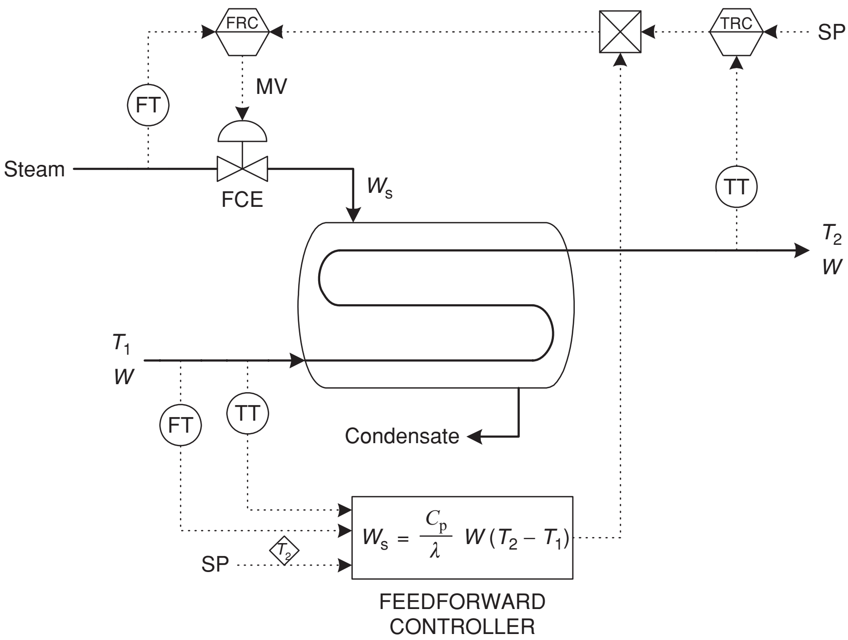

Figure 6.7 Combined feedforward and cascade control of a heat exchanger

Consider the following example of the feedforward control of a heat exchanger with cascaded feedback trim control (shown in Figure 6.7). The addition of feedback and cascade control serves to eliminate offset due to modelling inaccuracies and other non-measured disturbances.

At steady state an overall heat balance can be written for the process as shown in Equations 6.1–6.3:

$$q_{\text{in}} - q_{\text{out}} = 0 \tag{6.1}$$

$$W_s\lambda - WC_p(T_2 - T_1) = 0 \tag{6.2}$$

Or

$$W_s = \frac{C_p}{\lambda} W(T_2 - T_1) \tag{6.3}$$

where λ (kJ kg^{-1}) is the enthalpy transferred by the steam condensing to form condensate and C_p (kj kg^{-1}k^{-1}) heat capacity of the process fluid.

In this example, the inlet flow of liquid W and the temperature T_1 are measured to determine the amount of steam required as per Equation 6.3. The desired outlet temperature T_2 is the set point into the feedforward controller. The feedback temperature

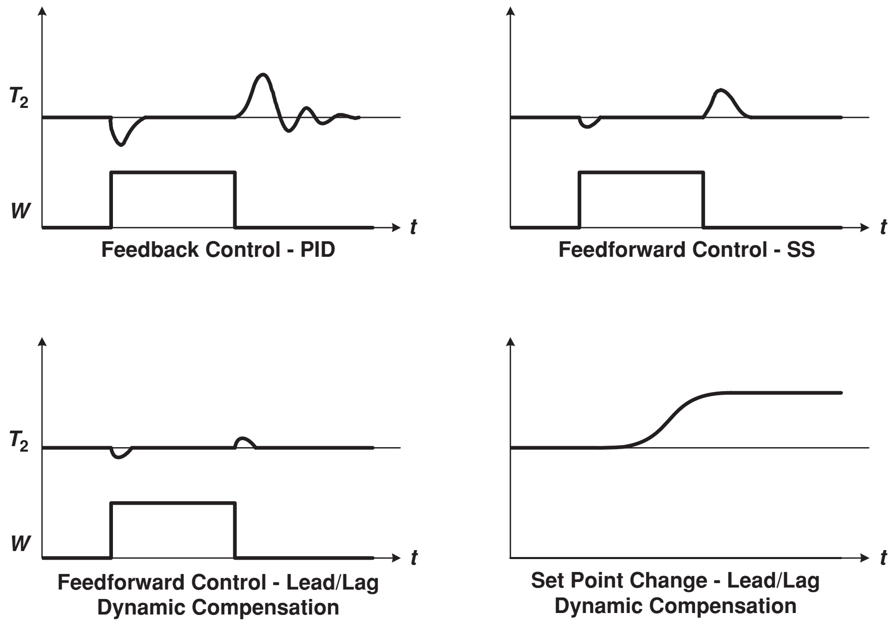

Figure 6.8 Typical responses of the heat exchanger

controller on the liquid stream measures T_2 to adjust for any disturbances that are not corrected by the feedforward controller.

Typical response curves for a load upset would appear as shown in Figure 6.8. Included in this figure for comparison is the response curve for feedback control only (with a PID controller) on the same process.

The response of the outlet temperature T_2 for the base case of the feedback control shows the type of improvement in control that can be achieved with even a simple steady-state feedforward controller. The lead/lag dynamic compensation [3,4] shows further improvement over the steady-state feedforward control.

6.3 Ratio control

Ratio control involves keeping the ratio between two variables fixed [2,4], as illustrated in Figure 6.9. Typically, these two variables, y and y_w, are flow rates, where y_w is the wild or uncontrolled flow rate and y is the manipulated or controlled flow rate. The wild flow rate is measured, and the controlled flow rate is then adjusted to maintain a fixed ratio between the two.

Ratio control can be considered a form of feedforward control. This is obviously true since, in ratio control, the process variable is measured upstream of the process, as is the case in feedforward control (Figure 6.4). Take, for example, a reactor with two liquid feed streams. Ratio control would ensure that these streams were being

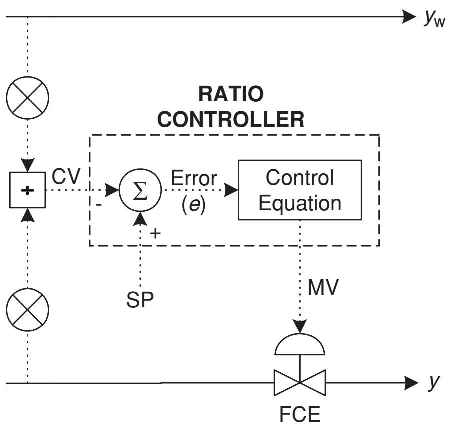

Figure 6.9 Typical ratio control system

stoichiometrically fed to the reactor by measuring the flow of one stream and adjusting the other accordingly. The product stream would be of no real use in determining whether the stoichiometric ratio was met.

There are two methods by which the ratio between the two variables can be fixed when only one stream is being manipulated. The first is shown in Figure 6.9.

In this first scheme, both flows are measured and divided to obtain the actual ratio. This is then compared with the set point and the flow of y is adjusted based on the difference. The set point to the ratio controller is the desired ratio.

In the scheme shown in Figure 6.10, the flow of the wild stream y_w is measured and then multiplied by the desired ratio. The output from the multiplier is the set point for the controller, which compares it with the measured flow and adjusts the flow of y

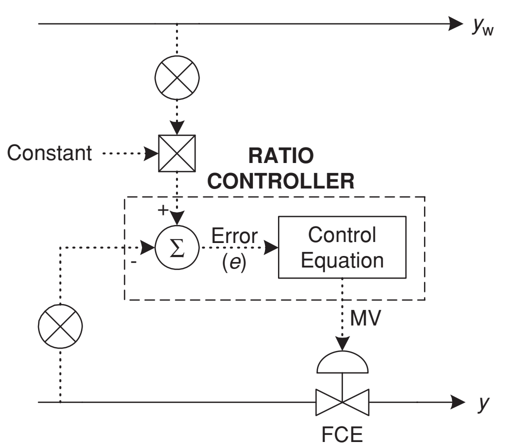

Figure 6.10 Typical ratio control system

accordingly. In this scenario, the desired ratio is a constant variable in the multiplier, and if a new value for the ratio is needed then it must be set in the multiplier.

The ratio control configuration shown in Figure 6.9 will not have a steady loop gain because the ratio calculation itself is in the loop. The loop in Figure 6.9 may become nonlinear, making the control configuration in Figure 6.10 a more reliable model, since its loop gain is constant.

A common example of ratio control is the case of an adsorption column, where a fixed ratio of *V/L* is desired. The wild flow rate is the vapour feed V to the column, and the controlled flow rate is the liquid flow rate L. The ratio control seeks to maintain constant absorption factors in the column by keeping a constant V/L profile.

6.4 Override control (auto selectors)

Frequently, a situation is encountered where two or more variables must not be allowed to exceed specified limits for reasons of economy, efficiency, or safety. If the number of controlled variables is greater than the number of manipulated variables, then a selection must be made for control purposes (SISO). A selector is used to accomplish this. Selectors are available in both electronic and pneumatic versions. The only difference between selectors is the number of inputs a particular hardware implementation may be able to accommodate. In this section, specific examples of such selectors will be discussed. It must be kept in mind that these are only a few examples of such auto selectors [4].

The two basic building blocks for selector systems are the high and the low selector. The high selector, shown in Figure 6.11, will pass the highest value of the multiple input to the output signal, ignoring all other inputs.

The low selector, shown in Figure 6.12, will choose the lowest of inputs to pass through as the output while ignoring all other inputs.

By using combinations of these basic building blocks it is possible to build other types of selector, such as a median value selection, shown in Figure 6.13. The selector output for a median value selector is a signal that falls between the highest and lowest input.

Let us investigate some typical applications of these selectors in four areas:

1 protection of equipment;

2 auctioneering (choosing from several signals);

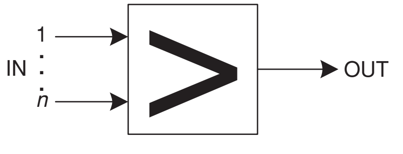

Figure 6.11 High selector

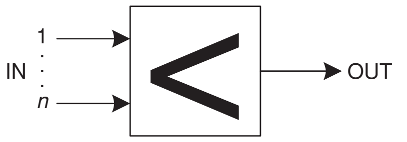

Figure 6.12 Low selector

3 redundant instrumentation (used commonly with process analytical equipment);

4 artificial measurements (establishing artificial limits).

6.4.1 Protection of equipment

To illustrate how selectors can be used to protect equipment, examine the pump system shown in Figure 6.14. The pump system demonstrates a situation where there are multiple measurements, multiple controllers and only one manipulated variable that can provide the following protection:

- Surge protection. When P_{in} drops below a certain minimum value, close the valve.

- High temperature. When the temperature of the pump exceeds a certain maximum temperature, close the valve.

- Excessive downstream pressure. When P_0 exceeds a certain maximum pressure, close the valve. (Assume $P_0 > P$ shut off.)

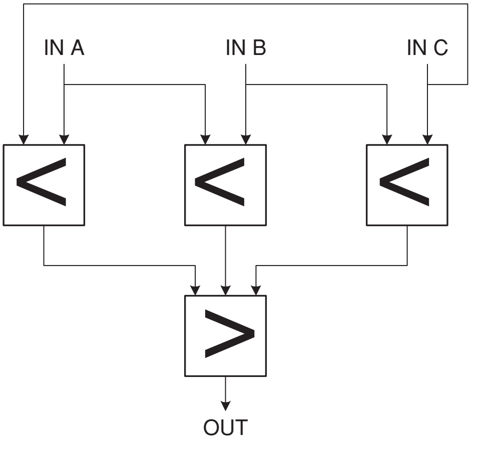

Figure 6.13 Median value selector

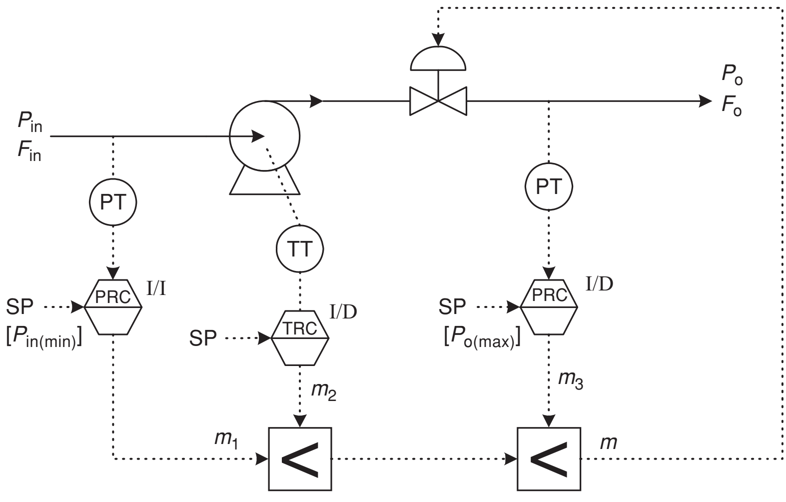

Figure 6.14 Protection of equipment: pump

Surge protection

As P_{in} begins to drop, the output m_1 will also decrease (note increase/increase action on pressure controller). The output m_1, will be selected by the first and second low selectors and will be passed through as the manipulated variable, causing the valve to close.

High temperature and excessive downstream pressure

If either the pump temperature or outlet pressure begins to increase, both outputs m_2 and m_3 begin to decrease (note increase/decrease action on both of these controllers). The smallest value will be chosen and passed through to manipulate the valve. In general, the smallest output from either of the controllers will always be operating the valve.

6.4.2 Auctioneering

The objective of autioneering is to protect against the highest temperature sensed by one of many temperature transmitters. In the example shown in Figure 6.15, the control equipment consists of one controller, four transmitters, and one FCE. The highest temperature will be selected by the high selectors and will be used as the measurement for controlling the fuel to the oven.

6.4.3 Redundant instrumentation

For an exothermic reactor (shown in Figure 6.16), too much catalyst can prove disastrous. By implementing a fail-safe scheme that consists of two composition transmitters

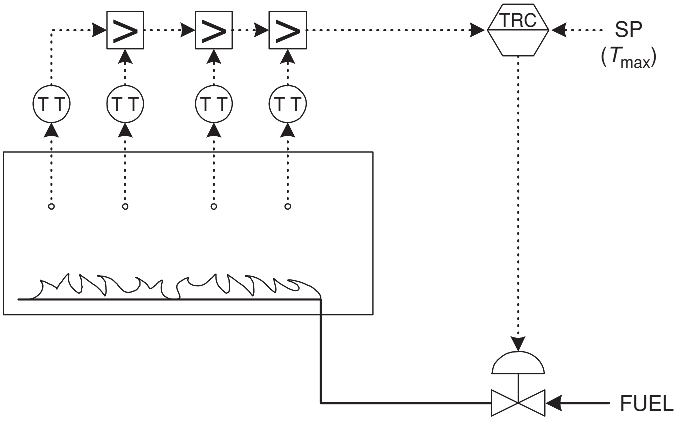

Figure 6.15 Auctioneering: temperature control in an oven

that are analysers and a high selector, the highest reading from the analysers will be utilized by the composition controller to control catalyst flow. The following actions will occur in the event of catastrophic failure of the analysers:

1 Downscale failure of analyser. If one analyser fails to zero, then the other will be selected to control catalyst flow and production will not be interrupted.

2 Upscale failure of analyser. If one analyser fails to full scale, then it will get selected and the catalyst flow will be stopped. Production is stopped and a possible hazardous situation is avoided.

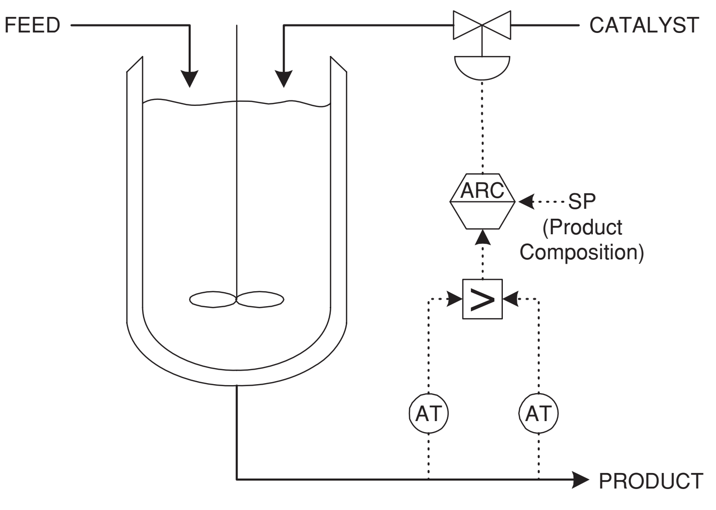

Figure 6.16 Redundant instrumentation: reactor

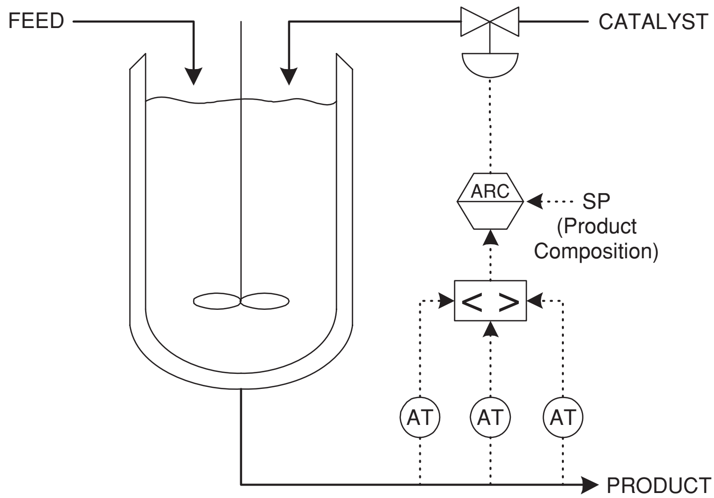

Figure 6.17 Redundant instrument: reactor/median selector

An alternate scheme, shown in Figure 6.17, implements analysers with a medium selector that will keep the process operating regardless of the failure mode of one of the analysers. The measurement variable to the controller will always be the median transmitter output. If one of the analysers fails, then either upscale or downscale the selector will still choose the median value.

In summary, the amount or quantity of redundancy depends on the importance of the process unit (reactor, distillation column, etc.). This is because the higher the quantity of redundancy the higher the cost (capital/operating) becomes and, therefore, the economics must be justified.

6.4.4 Artificial measurements

Some processes require certain operating constraints to be set. These are referred to as artificial measurements. These operating constraints can be set through the use of selectors. For example, consider a distillation column whose feed versus steam characteristic is shown in Figure 6.18.

Instead of operating the steam versus feed flow on a staight line, operating constraints are set. The operating constraints require a minimum steam rate of 10 per cent, even if the feed rate drops to zero. This sets the low limit of the steam flow. Furthermore, at maximum feed rate the steam rate should not exceed a maximum flow of 90 per cent, the high limit. These constraints can be implemented as shown in Figure 6.19.

If the feed flow is within the safe operating region, then the signal from the multiplier will pass through the high selector, since it is higher than the low limit. It will also pass

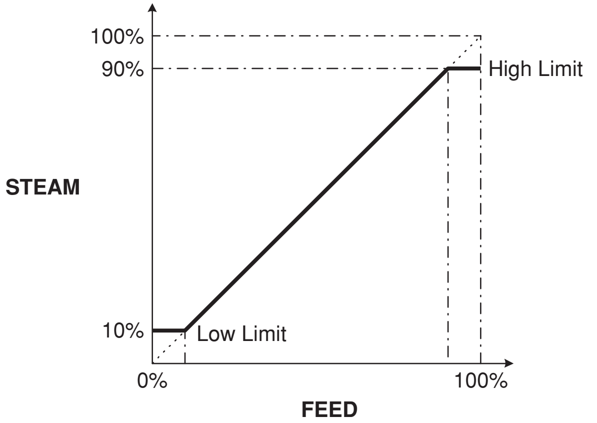

Figure 6.18 Feed/steam characteristic of a distillation column

through the low selector, since it is lower than the high limit, and then act as the set point for the steam flow controller. If the feed signal falls below the low limit or above the high limit, then the proper limit will be selected and that limit will be a constant high or low signal to the steam flow controller. This prevents the high and low limits from being exceeded.

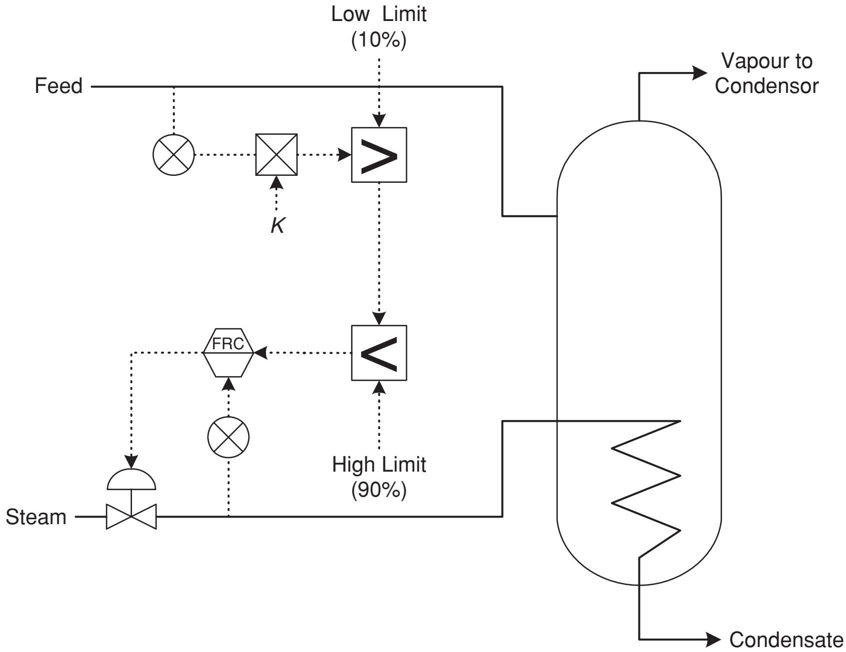

Figure 6.19 Artificial constraints

6.5 References

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2. Eckman, D. P. *Principles of Industrial Process Control*. Wiley, New York, 1945, pp. 194–9.
3. Murril, P. W. *Automatic Control of Processes.* International Textbook Company, Scranton, PA, 1967, pp. 431–44 (cascade/ratio), 405–25 (feedforward control).
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7

Common control loops

This chapter will describe some common loops encountered in process control. The loop characteristics, type of controller to use, response, tuning, and limitations will all be examined.

7.1 Flow loops

A typical flow control loop is shown in Figure 7.1. This process is very fast responding, and, even for long lengths of pipe, the dead time and the capacitance are very small. Typically, the process response is limited by the valve response (time constant).

As shown in Figure 7.1, the flow sensor/transmitter is always placed upstream of the valve for several reasons. First, many flow measuring devices have upstream and downstream straight-run pipe requirements. Usually, the upstream straight run is longer than the downstream straight run. Therefore, the flow measuring device can be placed closer to the valve upstream than downstream, where there might be problems with additional pressure drop through piping if a head flow device is used. Some examples of head flow devices are orifice plates, venturi tubes, and flow nozzles. Second, when the flow sensor is upstream from the valve there is a more constant inlet pressure, since it is closer to the source. Finally, there might be pressure fluctuations introduced to elements installed downstream from the valve as a result of valve stroking. Valve stroking results when the valve moves up and down, causing pressure changes that can affect downstream units or elements.

Another consideration when using head flow devices, besides additional pressure drop, is their nonlinear response, illustrated in Figure 7.2.

It is important to examine all the elements in the loop when determining what type of response is expected. For example, a differential pressure transmitter, also known as a d/p cell, has a linear response. However, when the head flow element, i.e. orifice, and the d/p cell are connected together the response is nonlinear, as shown in Figure 7.2.

A desirable attribute of a control loop is a response that is independent of the operating point, or a linear response. To this end, good practice requires offsetting nonlinearities in the loop to create an overall linear response. For example, when using a head flow device, a square-root extractor is used to linearize the flow signal. The square-root extractor is a device that simply takes the square root of the signal in order to linearize it.

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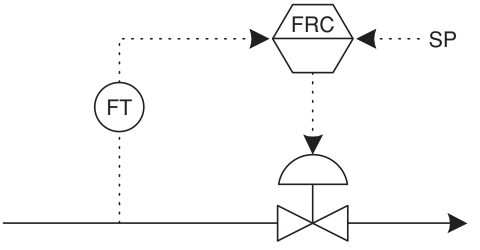

Figure 7.1 Flow control loop

Another factor that affects the linearity of the response is the characteristic of the valve selected. Three of the most common valve characteristics include quick opening, linear, and equal percentage. If the majority of the system pressure drop is taken across the valve, then a linear valve should be used since its installed characteristic will also be linear, giving the linear response desired. However, if the pressure drop across the valve is a small part of the total line drop and is not constant, then an equal percentage valve can be used since its installed characteristic will be close to linear. Quick-opening valves are most commonly used with on–off controllers where a large flow is needed as soon as the valve begins to open. More information on these valve flow characteristics can be found in Chapter 2.

Flow measurement by its very nature is noisy. Therefore, derivative action cannot be used effectively in the controller because the noisy signals cause the loop to become unstable. Flow control is one type of loop where an integral-only controller can be used. One drawback to I-only control is that it can greatly slow down the response of the loop, but the flow process is so fast that this slowing down may not be significant. To understand just how fast a flow loop is, consider again the heat exchanger cascade control scheme shown in Figure 6.4, where the primary loop may have a response period of several minutes. However, the secondary flow loop, even under I-only control, is fast enough for effective cascade control. Despite this, PI controllers are typically most often employed in flow loops because they are standard in the process control industry.

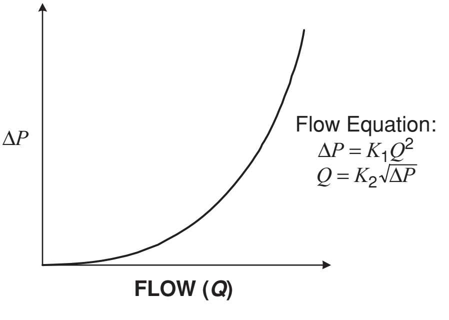

Figure 7.2 Head flow device response curve

Tuning a flow loop for a PI controller is rather easy in comparison with other types of loop. The flow loop is so fast that quarter amplitude damping (also called the QDR; see Chapter 5 for more detail) cannot be observed. The objective is for the flow measurement to track the set point very closely. To achieve this, the gain should be set between 0.4 and 0.65 (PB $\approx$ 200%) and the integral time T_i between 0.05 and 0.25 min. If there is an instability limit over the operating range due to nonlinearities in the loop, then the controller gain can be reduced, but the integral time should not. For a fast loop, such as flow control, an offset may persist because more gain is contributed from the proportional action than from the integral action.

7.2 Liquid pressure loops

A liquid pressure loop has the same characteristics as a flow loop. The objective of the loop is to control the pressure P at the desired set point by controlling the flow F as the needs of the process change (see Figure 7.3).

The pressure loop shown in Figure 7.3 is, in fact, a flow control loop, except that the controlled variable is pressure rather than flow. Since the flow is an incompressible fluid, the pressure P will change very quickly. The process behaves like a fixed restriction, i.e. an orifice plate, whose ΔP is a function of flow through the process. The process gain K_p can be determined from Equations 7.1–7.4.

$$P = \Delta P + P_o \tag{7.1}$$

where P_o is the downstream pressure at zero flow. Also:

$$\Delta p = \frac{F^2}{R^2} \tag{7.2}$$

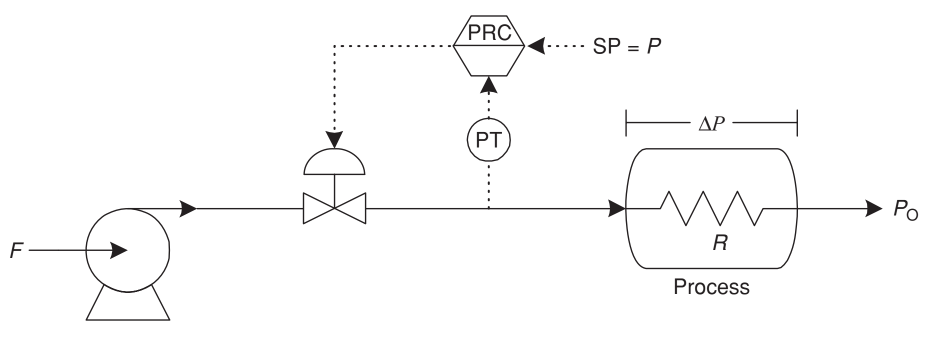

Figure 7.3 Liquid pressure control loop

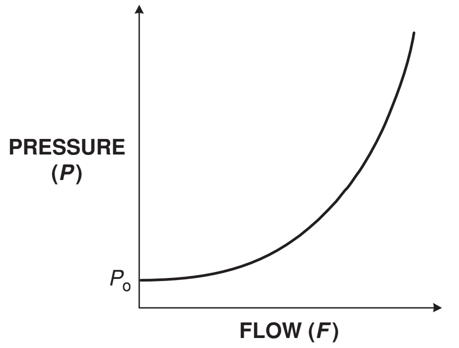

Figure 7.4 Process gain of pressure flow loop

so

$$P = P_o + \frac{F^2}{R^2} \tag{7.3}$$

As illustrated in Figure 7.3, R is the process flow resistance. The gain of the pressure loop is calculated using

$$K_p = \frac{dP}{dF} \tag{7.4}$$

Substituting Equation 7.3 into Equation 7.4 results in

$$K_p = \frac{dP}{dF} = \frac{2F}{R^2} \tag{7.5}$$

Plotting $P = P_o + (F^2/R^2)$ results in Figure 7.4, where the slope of the curve at any point is the process gain K_p as calculated using Equation 7.5.

The response of pressure to flow is exactly the same shape as the head flow device response discussed previously and shown in Figure 7.2. Therefore, the same rules apply for a liquid pressure loop as for the flow loop. The only difference between the two is that the pressure varies from P_o to 100 per cent, and not from 0 to 100 per cent as for the head flow device. For this case, the process gain is somewhat smaller than that for the flow process; thus, a higher controller gain can be used, i.e. between 1 and 2.

Other considerations for the liquid pressure loop are as follows:

- The controller can be proportional plus integral (PI) or integral only (I-only) and is tuned similarly to the flow controller.

- K_p is not constant; therefore, a square-root extractor should be used or the highest loop gain should be used in tuning the controller. The reason behind using the highest loop gain is to prevent the loop from ever becoming unstable. This concept is explained in more detail in Section 7.3.

- Since the liquid pressure loop is similar to a flow loop, it is also noisy. Therefore, derivative action in the controller is not advisable.

7.3 Liquid level control

A liquid level control loop, shown in Figure 7.5, is essentially a single dominant capacitance without dead time. Typical hold-up times are from 5 to 15 min. Typically, processes that are dominated by a single, large capacitance are the easiest to control. However, liquid level processes are not necessarily as simple as they first appear to be. In many liquid level control situations, considerable noise in the measurement is present as a result of surface turbulence, stirring, boiling liquids, etc. The fact that this noise exists often precludes the use of derivative action in the controller. Still, some applications use unique methods of level measurement to minimize the noise in the measurement in order to apply derivative action in the controller.

The first example of using a unique measurement method to minimize noise is using a displacer in a stilling well, shown in Figure 7.6. The intention of this arrangement is effectively to filter out high-frequency noise due to turbulence in the measurement by using the stilling well. However, one caution is that the tank and stilling well form a U-tube and the result could be low-frequency movement of the liquid from the tank to the well and back. This will make the transmitter believe that the level is slowly moving up and down. If control action were taken, then the controller would actually be aggravating the situation.

Another example of noise filtering is to use ultrasonic level measurement with electronic filtering of the level signal, shown in Figure 7.7. This method works because

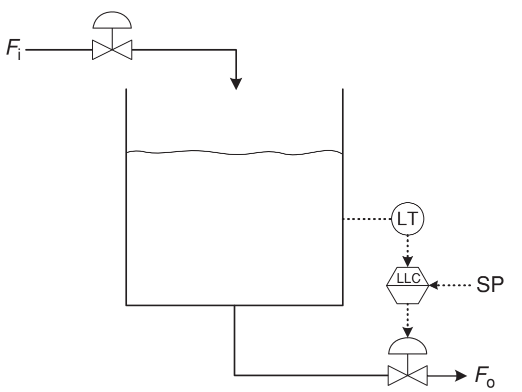

Figure 7.5 Liquid level control loop

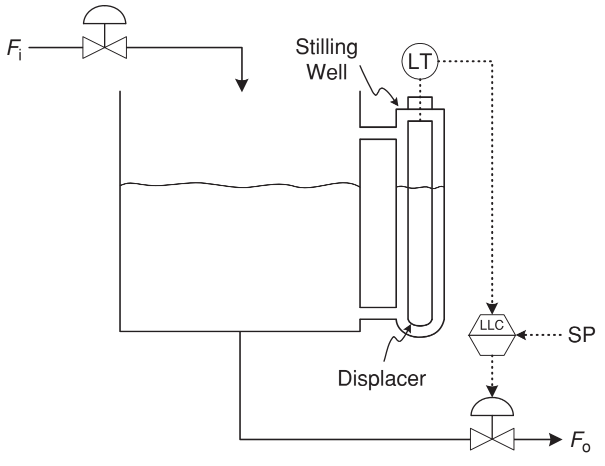

Figure 7.6 Liquid level measurement with a stilling well

the noise frequency is much higher than the period of response of the tank level. The electronic filtering in this case is a relatively simple matter.

Another technique that has been employed effectively to minimize noise is to use some sort of tank weighing method, illustrated in Figure 7.8. In this application, a load cell is placed under each tank support in order to measure the mass of the tank. The outputs are sent to an averaging weight transmitter, and then to a weight/level converter

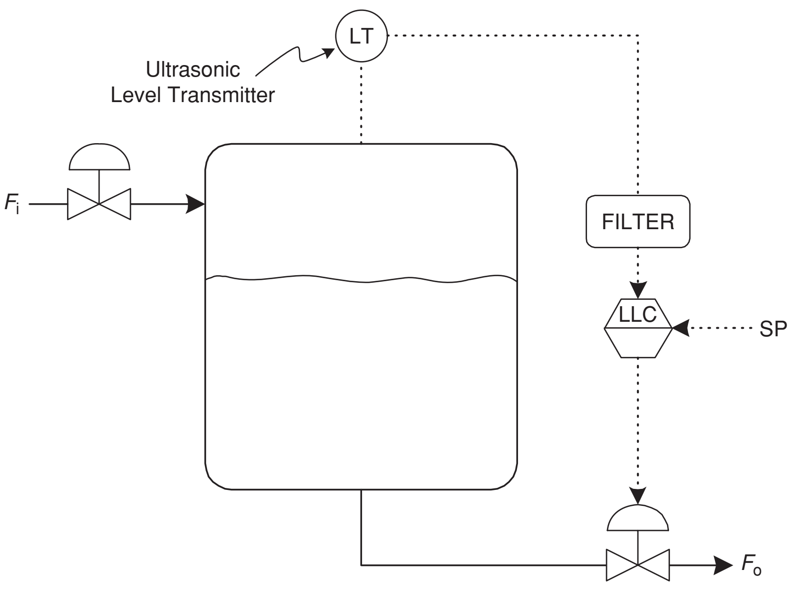

Figure 7.7 Ultrasonic liquid level measurement

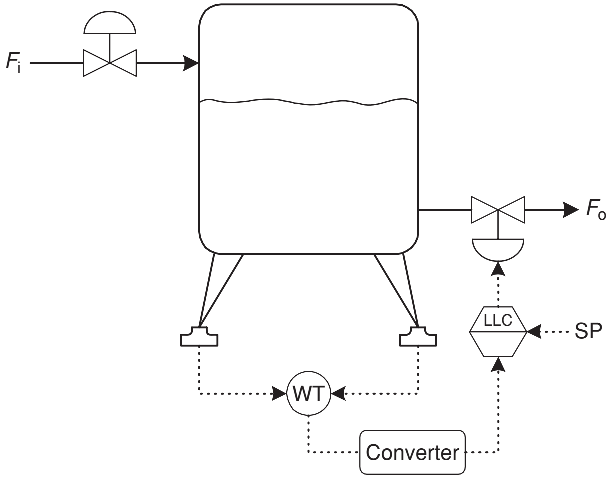

Figure 7.8 Using a load cell to measure mass in tank

before entering the controller. Obviously, this method is effective in eliminating noise in the measurement because the turbulence in the tank does not affect the weight measurement.

The three methods suggested (Figures 7.6–7.8), are ways to minimize noise in the level signal so that derivative action can be used in the controller. Derivative action in the controller will overcome the sluggish response caused by the integral action. Integral control is required to maintain the level at the set point, which cannot be accomplished by a proportional-only controller. Just how large of an offset results from applying P-only action to a level process and is it small enough to justify use of a P-only controller for liquid level control? The equation for the error resulting from P-only control of a process is

$$e = \frac{\mathrm{MV} - b}{K_\mathrm{c}} \tag{7.6}$$

As seen in the above equation, increasing the controller gain K_c will minimize the error. If the process gain is low, then the larger controller gain will result in a small error while still maintaining a stable loop.

Therefore, applying a P-only controller to control the liquid level in large tanks should definitely be considered. In many cases, an acceptably small error of only a few per cent will result. The response will be just as fast as with a PID controller, and noise in the measurement does not need to be a consideration. Effective application of P-only control is possible in this case because of the low process gain of the large-capacity tank, which allows for a high controller gain and thus smaller error. P-only control should be considered whenever a single large dominant capacitance with very little or no dead time is present. Keep in mind, however, that tight level control is not always

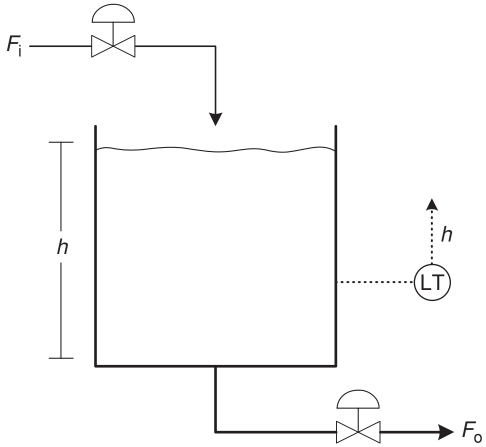

Figure 7.9 Liquid level process

desirable. Deviations from the level set point can sometimes be tolerated in exchange for a smoother flow in the manipulated stream feeding other, more sensitive equipment. This is termed averaging level control [1–4].

Another interesting problem to be considered in a liquid level process is the dependence of the process gain on load, which is a problem that exists in any single dominant capacitance situation. Load is defined as anything that will affect the controlled variable under a condition of constant supply. Consider the open-loop case of a liquid level process shown in Figure 7.9. The process gain K_p for a constant outflow F_o is calculated using

$$K_p = \frac{\Delta \text{Out}}{\Delta \text{In}} = \frac{\Delta \text{Level}}{\Delta \text{Inflow}} = \frac{\Delta h}{\Delta F_i} \quad \text{for } F_o = \text{Constant} \tag{7.7}$$

Since the outflow is set at a constant value, the inflow is considered to be a load on the process. Figure 7.10 shows the effect that the inflow has on the process gain. In this figure, the process gain is the slope of the curve and F_i is the relative opening of the inlet valve.

In the first case, the outlet valve is closed, $F_{01} = 0$, and $F_i \neq 0$, causing the tank level to increase. The level will theoretically increase to infinity or $K_p = \infty$. However, in reality, the tank will overflow and the level h will never saturate at the maximum tank capacity. Also, regardless of what the level is in the tank, if F_{01} is set to zero and then F_i is set to the original flow, the level will continue rising at the same rate. This is not the case if $F_0 \neq 0$. As F_0 gets larger, the steady-state level is a lower value. The reason that this occurs is because, as F_0 is made larger, the head in the tank does not have to be as large to make the inflow equal to the outflow. Also, for any given outlet valve setting, i.e. F_{02}, F_{03}, etc., if $F_i = F_0$ and then F_i is moved to its original setting, the level in the tank will rise, and Δh for a given ΔF_i will be less.

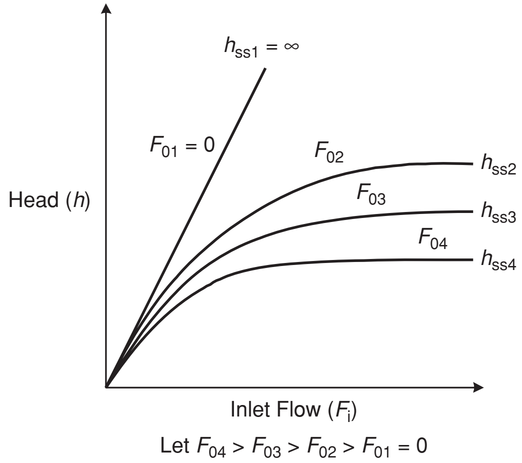

Figure 7.10 Head versus inlet flow for liquid level process

This behaviour is generally true of a class of processes dominated by a single capacitance. The process gain is a function of the load, with the gain decreasing as the load increases, as shown in Figure 7.11.

However, the inflow is not the only load on this process. The set point, i.e. fixed value of F_o, is also considered to be a load, as illustrated in Figure 7.12.

As the set point is increased, $SP_3 > SP_1$, the process gain, which is the slope of the curve at SP_1, SP_2, and SP_3, would decrease. Again, the process gain shows a reciprocal relationship to the load. In this case, the load is the set point.

Why is the dependence of process gain on load a consideration? The previous discussion shows that, when a process contains a capacitance and the controller gain is adjusted to give a particular response at a given load, the response will change as the load changes. If the load on the process is reduced, then the process gain rises and,

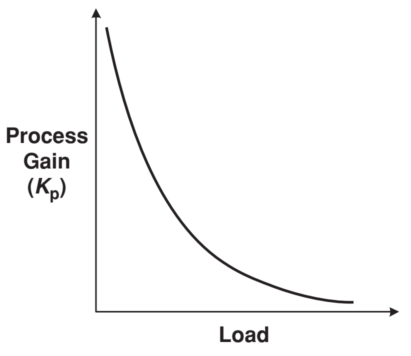

Figure 7.11 Relationship between load and process gain

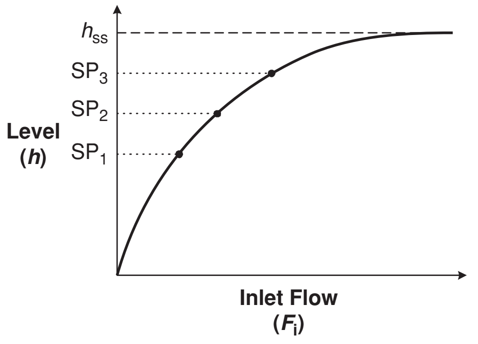

Figure 7.12 Gain and set-point relationship

therefore, the loop response tends to be more oscillatory. If the process gain increases enough, then the loop could become unstable. On the other hand, if the load increases, then the process gain decreases, and the loop responds sluggishly. The important fact to remember is that the loop must not become unstable, i.e. the loop gain K_L must be less than unity. Therefore, for the situation where the process gain is a function of the load, the simplest thing to do is tune the loop at the highest process gain and live with a sluggish response for the situation of the process gain decreasing with increasing load.

Another approach to this situation would be to put a component in the loop that would have a complementary gain to the process gain. An example of this is using a square-root extractor with a head flow meter in the flow control loop. If the pressure drop across the valve remained fairly constant, then the valve and installed characteristic would be nearly the same. An equal percentage valve could be used to complement the process, and the product of the valve and process gain ($K_V K_P$) would almost be constant, as illustrated in Figure 7.13.

Yet another approach to this situation would be to adjust the controller parameters, the controller gain for example, with the variation in load of set point to compensate for the variation in process gain. This approach is termed gain scheduling [5,6] or programmed adaptation [7,8] and can be considered to be form of adaptive control [5].

Another level control situation where a nonlinear element might be introduced into the loop is the case of level control of a cylindrical tank lying on its side, shown in

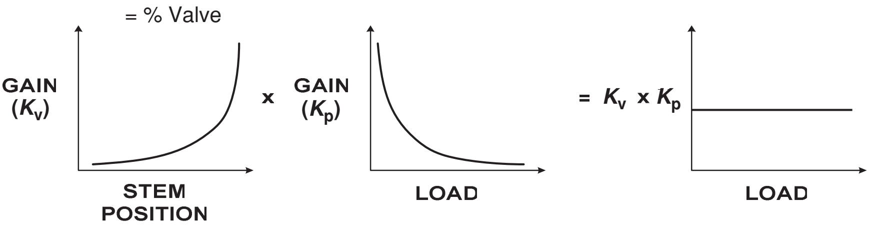

Figure 7.13 Load versus gain

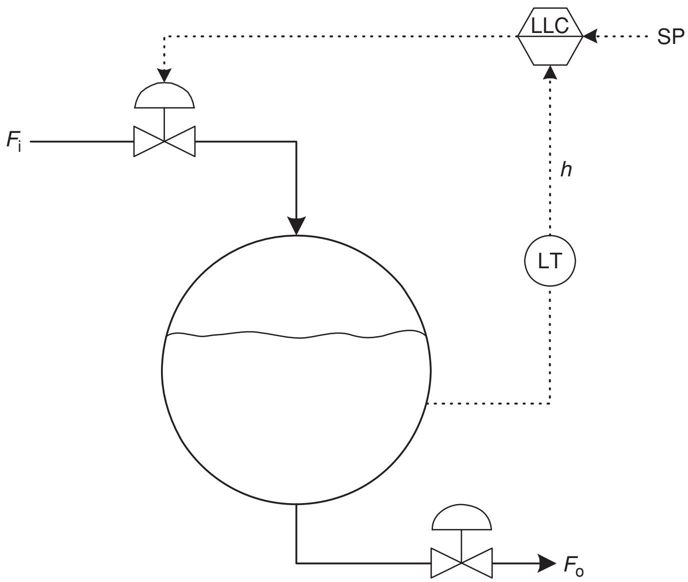

Figure 7.14 Cylindrical tank

Figure 7.14. The previous liquid level processes were shown as having the outlet flow as the manipulated variable. Although this is the common method for liquid control, the inlet flow can also be used to control the level in the tank, as shown in this example.

The response of level in the cylindrical tank shown in Figure 7.14 is given in Figure 7.15. Obviously, owing to the differences in tank geometry, the response for the horizontal, cylindrical tank differs from those previously discussed.

Figure 7.16 represents the process gain qualitatively as it varies with load. In this case the load is the height h in the tank. The gain of the tank is high at both low and high levels and is low at normal levels in the cylindrical tank.

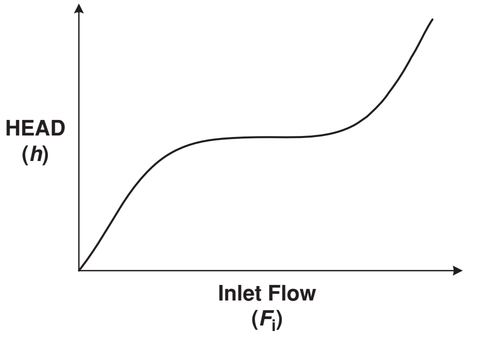

Figure 7.15 Cylindrical tank level response

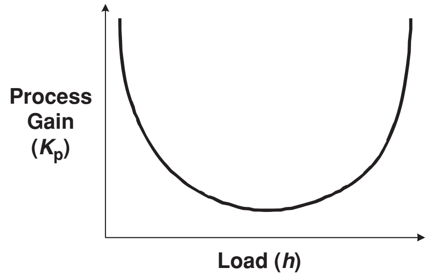

Figure 7.16 Process gain versus load for a cylindrical tank

To make the loop gain independent of the tank level, a signal compensator must be added to the liquid-level loop. The signal compensator has a response that can be varied as shown in Figure 7.17; thus, when the process gain is high, the compensator gain is low and vice versa, thus giving an overall linear response. Note that current practice is to use digital controllers and the compensation can be done within a DCS function block.

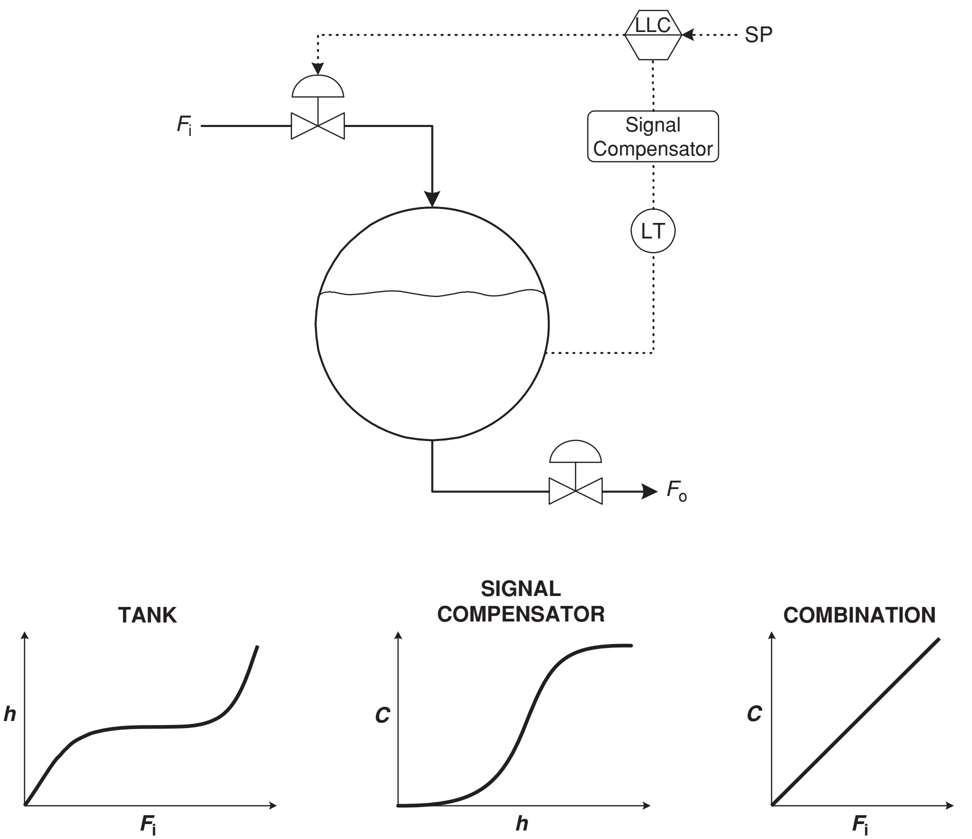

Figure 7.17 Cylindrical tank level compensator

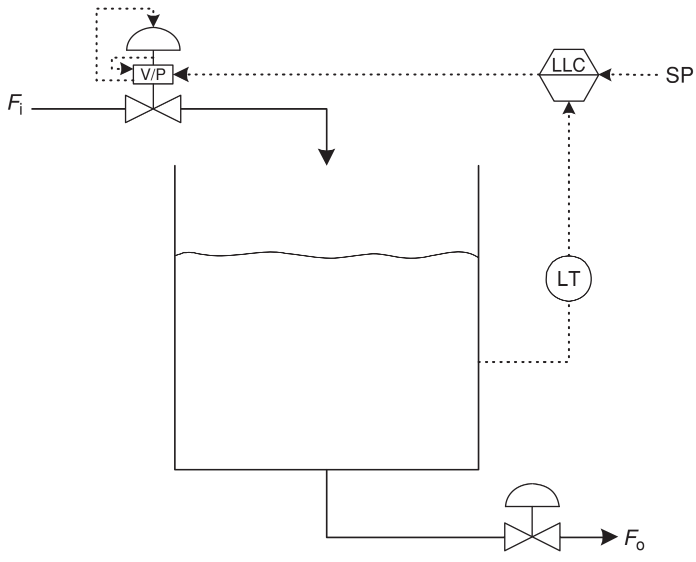

Figure 7.18 Liquid level control loop

Now consider the case, shown in Figure 7.18, of a tank with a P-only controller and some valve hysteresis. All valves have some hysteresis, but excessive valve hysteresis typically occurs when the valve sticks as it tries to open and close. This can happen for a number of reasons, including overtightened packing, etc.

The response for this process might not be as good as desired, since any misposition in the supply valve will show up as an incorrect level to the controller. If this were a large tank, then the response might be very slow because the level would change very slowly due to the large surface area. By applying a valve positioner, V/P, a cascade system is set up that will effectively minimize the effects of valve hysteresis and improve loop response. In this cascade system the level controller acts as the primary loop and the valve positioner is the secondary loop. A valve positioner can be used in this case because the response of the level control loop is much slower than that of the valve positioner loop and, therefore, the rule $\tau_{01} > 4\tau_{02}$ is obeyed, giving an effective cascade system.

Another level control situation commonly encountered is that of using the capacity of a tank to prevent surging or pulsing of flow from upstream processes to downstream process units. This is the general case of an integrating process that does not require tight control. An example of an integrating process is a buffer tank with pumped outflow, as illustrated in Figure 7.19. The buffer tank provides exit flow smoothing in the face of incoming flow disturbances. The objective is not to provide tight level control, but to let the level swing. Since the tank and pump are in effect an integrator, if a PI level controller were used the result would be a double integrator with the potential for continuous cycling.

There are two popular approaches to tuning integrating processes to provide flow smoothing of the exit stream. One involves proportional-only action, and the other uses proportional plus integral action.

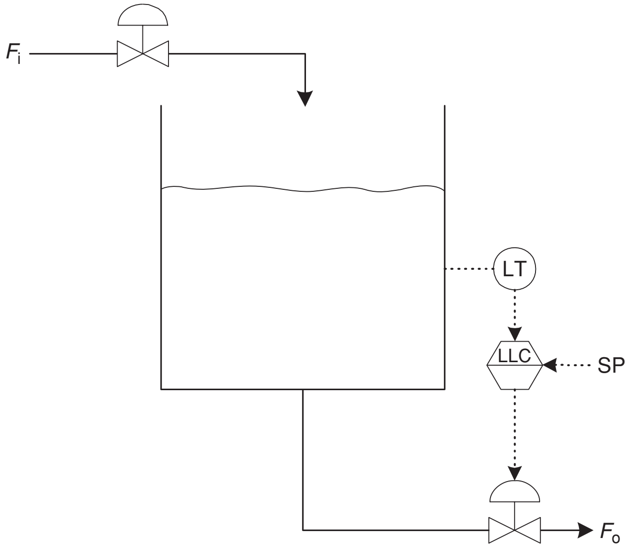

Figure 7.19 Integrating process: flow smoothing

7.3.1 Proportional-only control for integrating processes

The following defines the desired conditions for proportional only control of an integrating process with flow smoothing:

$$\text{Proportional controller gain} = 2\%$$

$$\text{Bias} = 50\%$$

$$\text{Set point} = 50\%$$

To set the bias at 50 per cent, put the controller into manual, set the output to 50 per cent, and then switch to automatic. With no integral action, the bias will remain at whatever the valve position was when last in manual mode.

These settings offer the following benefits:

1 Output flow will shut off before the level drops below 25 per cent; *thus, the tank cannot run dry.*

2 Output flow reaches the maximum before the level exceeds 75 per cent; *thus, the tank level will never exceed 75 per cent for maximum throughput.*

3 Effective buffering is increased when compared with using a PI controller. In other words, at low flow the level is low and there is room for the most likely change, a large dump of fluid into the tank. In fact, there is approximately 50 per cent more room, i.e. 25/50 or 50 per cent.

Using Equation 4.3 for a proportional-only controller (for a direct acting controller, $e = \mathrm{CV} - \mathrm{SP}$) and inserting the above values for the set point SP, the bias b, and the controller gain K_c, and rearranging the results gives

$$\mathrm{MV} = 2\mathrm{PV} - 50 \tag{7.8}$$

Therefore, at a 25 per cent level the output is zero, and for a 75 per cent level the output is 100 per cent. What may not yet be obvious is how the effective buffering is increased. Consider a lower than normal throughput, say MV = 5%. From Equation 7.8, the level PV would line out at about 27.5 per cent. For a higher than normal throughput, say MV = 95%, the level would line out at about 72.5 per cent. Therefore, P-only control will significantly decrease disturbance transmission downstream [4]. At high throughput, the level runs high; and for low throughput, the level runs low, always providing maximum room for the most likely disturbance. The increased buffering comes from the fact that:

- at low throughputs there is more ‘headroom’ to buffer sudden increases in throughput when the level is low;
- at high throughputs, there is more inventory in the tank to buffer sudden decreases in throughput when the level is high.

PI controllers will all eventually drive the level to the set-point value, and therefore lack this particular benefit. However, P-only level controllers have two important drawbacks:

1 The proportional gain of 2 required here is sometimes higher than desired and provides insufficient flow smoothing, i.e. small holdups. Lower gains offer improved flow smoothing; however, there is the risk of running the tank dry or overflowing the tank during maximum throughput.

2 Without integral action, P-only controllers typically never operate at the set point, so there is always an offset between the level and its set point. While this is actually what provides some of the benefits described above, many operators dislike seeing this sustained offset and resist its use.

7.3.2 PI controller tuning for integrating process

If the limitations of the P-only controller preclude its use, then the following outlines a tuning procedure for a PI control scheme:

1 Select a value for controller gain that is less than 2. Try a gain between 0.5 and 1.0. Only if $K_c < 2$ does it make sense to use a PI controller at all; otherwise, a P-only controller is used.

2 Determine the total holdup time T_{HU} of the tank by dividing the volume of the tank, as measured between the minimum and maximum level control points, by the maximum flow through the control valve (Equation 7.9). It is important to note that the volume in Equation 7.9 is the volume of the tank between minimum and maximum controlled levels, not the total tank volume.

$$T_{HU} = \frac{\text{Volume}[\text{ft}^3]}{\text{MaxFlow}[\text{ft}^3\ \text{min}^{-1}]} = \frac{V}{Q_{max}} \tag{7.9}$$

3 Calculate the integral time using

$$T_i = 4\ \frac{T_{HU}}{K_c} \tag{7.10}$$

7.4 Gas pressure loops

The characteristics of the gas pressure loop are almost the same as that of a liquid level control loop. A typical gas pressure loop is shown schematically in Figure 7.20.

Varying the flow of a compressible fluid controls the pressure in a large volume. This process is dominated by a single large capacitance with no dead time. The measurement is normally noise free and, owing to its capacitive nature, is characterized by a slow response and a small process gain. As shown for liquid level control, a proportional controller is more than adequate for gas pressure control.

The gas pressure loop is perhaps the easiest type of process loop to control. Owing to the low gain in the process, a high controller gain will result in good control with very little offset and very little possibility of instability. It is perhaps the only loop in the fluid processing industry that is very close to being unconditionally stable. As with the level control loop, a valve positioner can be used to improve loop response for a valve with hysteresis.

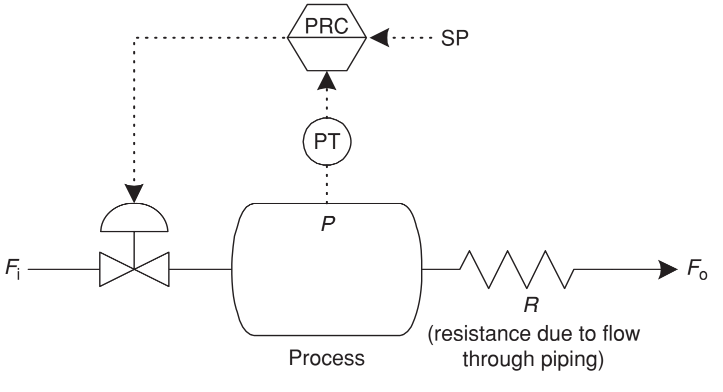

Figure 7.20 Gas pressure control loop

The gain of this process is a function of the load F_o. However, since the loop is almost unconditionally stable, it is not necessary to tune the controller at the highest process gain. The process gain changes, but even at the lowest load, it is stable. It is simply not possible to increase the controller gain to a high enough value to cause cycling.

7.5 Temperature control loops

Temperature loops may be divided into two main categories:

1 endothermic, i.e. requiring heat energy;

2 exothermic, i.e. generating heat energy.

Both of these processes have similar characteristics, in that they are typically comprised of one large and many small capacities, i.e. valve actuator, transmitter, etc. The net result is a response indicative of a process with a dominant capacitance plus a dead time. Both of the above categories will be investigated and their specific differences and similarities will be identified.

For both of these processes, one of the following devices for measuring temperature is used:

- thermocouple (TC);
- filled thermal system (FTS);
- RTD.

Although the overall loop response is characteristic of a large dominant capacitance plus dead time, care should be taken when installing temperature measuring elements. The temperature measuring devices should be selected so that the devices add a minimum lag to the process lag. It is common practice to insert the measuring element into a thermal well to protect it from the process fluid and to facilitate change out of the element if a problem should occur. Thermal wells are typically made of metal or ceramic, depending on the environment. Figure 7.21 illustrates a thermal well.

Every component in the measuring system, as shown schematically in Figure 7.21, has an associated time constant τ, where $\tau = RC$ (see Equation 3.28). So, each component in the measuring system will increase the measurement lag depending on the size of its time constant. Good practice dictates that the dominant time constant is the process fluid time constant, $\tau_{\mathrm{PF}} = R_{\mathrm{PF}}C_{\mathrm{PF}}$. The other time constants, i.e. thermal well $\tau_{\mathrm{w}} = R_{\mathrm{w}}C_{\mathrm{w}}$, thermal well fluid $\tau_{\mathrm{AG}} = R_{\mathrm{AG}}C_{\mathrm{AG}}$, and measuring device $\tau_{\mathrm{B}} = R_{\mathrm{B}}C_{\mathrm{F}}$, should be as small as possible.

These time constants may be minimized in various ways. For instance, τ_{w} can be reduced by choosing a thermal well made from a material of low thermal resistance and also made as small as possible to reduce C_{w}. To minimize τ_{AG}, the air gap is filled with a

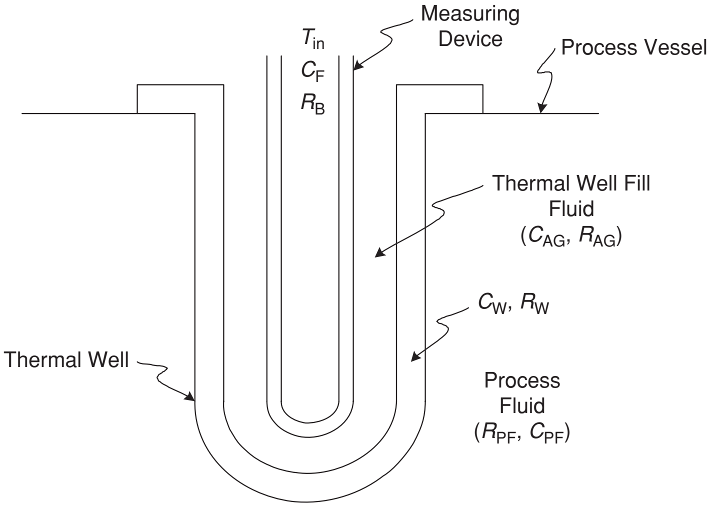

Figure 7.21 Typical temperature measurement device installation

highly thermal conductive fluid to decrease R_{AG} or the measuring device is attached via welding to the well. Using the smallest bulb and shortest capillary possible to reduce both R_B and C_F will minimize τ_B. With respect to short capillary runs in an FTS, to increase response speed a transmitter is mounted close to the process and a gas-filled bulb is used. This is done to minimize capillary length and gas is used because its thermal capacity C_F is low. This, in turn, results in a small measurement time constant, since the pressure signal from the FTS is changed to an electronic signal of 4–20 mA sooner without a long capillary run.

When using an RTD or a TC the time constant considerations are similar, but the actual response times of the devices will vary. The FTS and RTD will have response times of nearly the same magnitude, whereas the TC is somewhat faster. For a thermocouple, τ_B varies with the device's construction and length of the extension wires. Hence, a TC made of small wire with short extension wires will give a fast response. A typical TC response is around 0.5 s.

Let us put some numbers on the response times of the other two temperature measurement devices. The FTS and RTD have similar response times, except that the response of the RTD in water is generally longer than the FTS owing to its greater internal resistance. But, for the same size bulb, either a liquid-filled FTS or RTD has a time constant of about 3 s.

For an FTS the capillary has a time constant of 0.55 s per 10 ft, and thus it is obvious why a transmitter is often used, since some capillaries can be up to 100 ft long. Using a short capillary, with a gas-filled system and a transmitter, results in a response that is two times faster than the liquid-filled system with a long capillary.

The response times of RTD or FTS bulbs in a thermal well depend on the material of the well and the clearance between the bulb and the well. For a bulb in a dry well a typical time constant is 1–2 min, whereas for a bulb with thermal fluid in the well the following apply:

- *in a gas stream* – τ is the same as for the dry well, owing to the high thermal resistance of the gas (τ_{PF} is very large);

- *in a liquid stream* – τ should be about two to three times that of the bulb alone, since the thermal resistance of the liquid is very small (τ_{PF} is very small) and the response improves with the lowering of the thermal well fluid resistance, i.e. make τ_{AG} as small as possible.

If it is necessary to use a thick metal or ceramic well because of corrosive process fluid, then the response time can increase to 10 times that of the bulb alone. In addition, a large well creates a static error as a result of conduction along the wall of the thermal well. The addition of a large well can increase the measuring device plus well time constant by approximately 1.5 min. This increase can be detrimental in certain processes, i.e. in an exothermic reactor, but is not significant in others.

The following are general rules of thumb for reducing temperature measurement lag:

1 Use a small-diameter bulb or thermal well.

2 Increase the velocity of flow past the measuring device by using a small pipe or a restriction orifice near the bulb. Be cautious though, because there is the possibility of thermal well fracture as the velocity increases.

3 When measuring temperature in two-phase flow situations, place the measuring element in the liquid phase, if possible, to gain the benefit of faster heat transfer from the process fluid to the measuring element.

4 Consider using a transmitter with derivative action. Some manufacturers make a gas-filled thermal system connected via a short capillary to a temperature transmitter with derivative action in it. The derivative action acts to cancel out some of the lag in the measuring element. Reduction of the derivative gain in the controller is required to accommodate this added derivative gain in the transmitter.

The whole point of the preceding discussion on minimizing temperature measurement lag in the temperature control loop is to make you aware that this is important in both slow and fast loops.

7.5.1 The endothermic reactor temperature control loop

A good example of an endothermic process is a process heat exchanger being used to heat a fluid from the inlet temperature T_1 to an outlet temperature T_2, as shown previously in Figure 6.2. This heat exchanger's response will be that of a single large dominant capacitance with at least 30 s of dead time. Typically, either a PI or PID controller is used. Derivative action can be used, since the temperature measurement is not noisy. The response of the loop under PID control will be equal to that of P-only control, except the temperature will be maintained at the desired set point.

The steady-state gain of the heat exchanger is calculated using

$$K_\mathrm{p} = \left(\frac{\Delta T_2}{\Delta F_\mathrm{s}}\right)_{F_\mathrm{w}} = \text{constant} \tag{7.11}$$

The process gain K_p is a function of the load F_w, as in the case of liquid level control previously discussed. However, F_w is not the only load; it is one of several. Other loads include the inlet and outlet temperatures of the cold fluid stream. The steady-state equation describing the behaviour of the heat exchanger is given by Equations 7.12 and 7.13:

$$F_\mathrm{s} = KF_\mathrm{w}(T_2 - T_1) \tag{7.12}$$

$$K = \frac{C_p}{\lambda} \tag{7.13}$$

where C_p is the specific heat of F_w, the flow of cold fluid (load variable), λ is the heat of condensation, T_1 is the inlet temperature of the cold fluid (load variable), and T_2 is the outlet temperature of the cold fluid (load variable).

The heat exchanger energy balance equation can be solved for the heat exchanger gain K_p thus:

$$K_\mathrm{p} = \frac{\mathrm{d}T_2}{\mathrm{d}F_\mathrm{s}} = \frac{1}{KF_\mathrm{w}} = \frac{K'}{F_\mathrm{w}} \tag{7.14}$$

As expected, the gain is inversely proportional to the load F_w. This result is identical to that for the liquid level process and may be minimized with similar approaches.

A valve positioner, V/P, can also be added to the flow valve as described in the case of liquid level control. However, if there is a chance of a supply upset to the heat exchanger, then a temperature on flow cascade is used instead, as shown in Figure 6.4. Another approach to minimizing supply upsets is to use a pressure regulator ahead of the steam. The pressure regulator is used to make the steam supply pressure constant. This scheme negates the need for another flow loop while still providing protection against supply upsets.

There are many methods used in controlling heat exchangers. Figures 7.22–7.25 show several methods, in addition to the basic feedback loop, in which the flow of steam was directly throttled by the temperature controller.

Figure 7.22 shows a situation in which F_s is a wild flow and T_2 is controlled by controlling the condensate level in the heat exchanger, i.e. overhead condenser. When the temperature is too high, the valve closes, which causes condensate to cover more

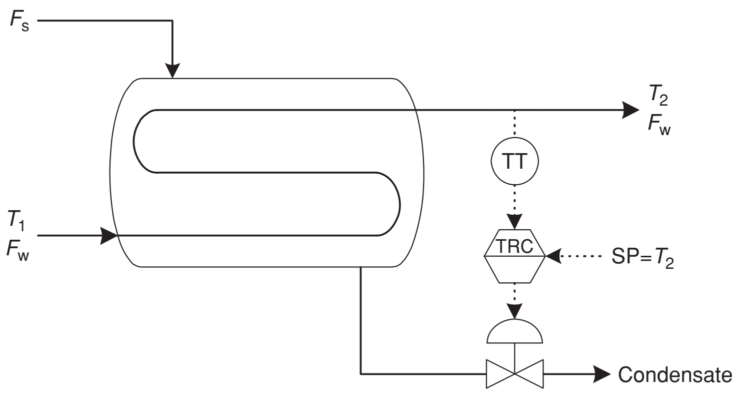

Figure 7.22 Temperature control via level control

tubes and reduce heat transfer to the cold fluid. Because the condensation time is large, the response is slower than for other systems. Also, because of condensate splash, T_2 can show significant fluctuations.

Figure 7.23 shows a scheme employed when temperature control is critical and the response time τ_1 of the heat exchanger is very long.

In this approach, a sidestream of the input F_w, is bypassed and mixed with the outlet F_w. This gives fast response, with an energy penalty of first heating up and then cooling down F_w. It is also necessary to ensure good mixing at the output and to ensure a fast response in the temperature measurement, since a flow loop is being used. Another variation on the control scheme shown in Figure 7.23 is shown in Figure 7.24.

The scheme shown in Figure 7.24 provides control over a wider range of F_w and gives a nearly constant process gain. Hot and cold F_w are blended together to maintain

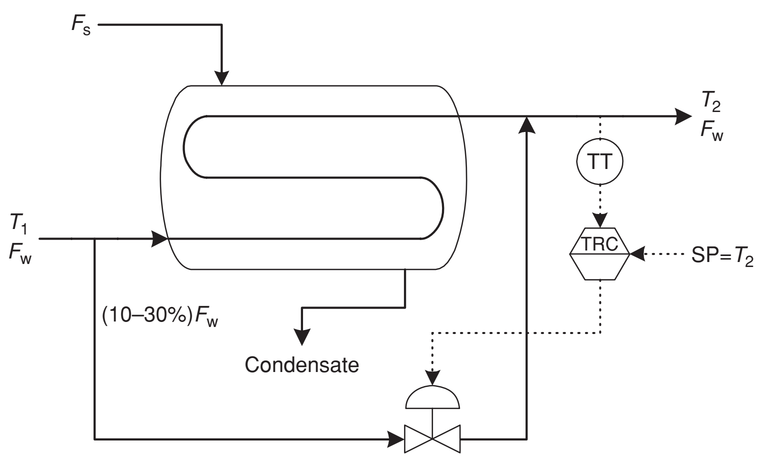

Figure 7.23 Control scheme for critical temperature control

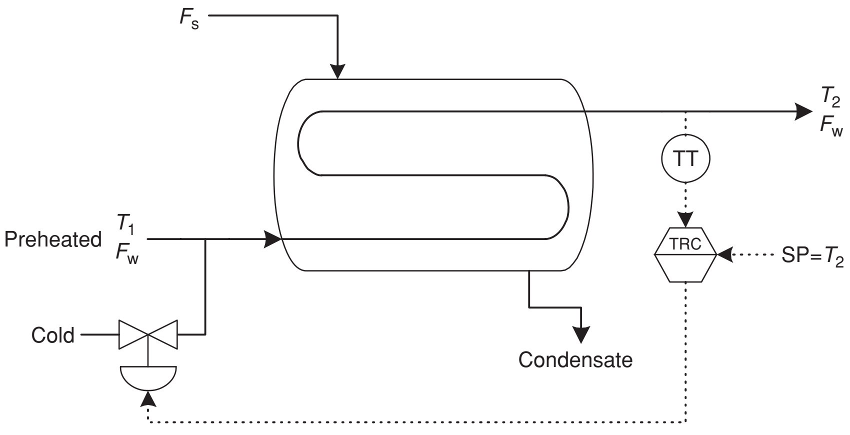

Figure 7.24 Variation of critical temperature control scheme

a uniform T_1. The energy penalty for this scheme is the cooling of the stream for which energy has been expended in heating up.

The scheme shown in Figure 7.25 throttles F_w, to maintain T_2, and is usually used on heat exchangers that are capacity limited. It is more important to maintain T_2 at the set point than to maintain F_w at a given demand.

A detailed discussion of tube and shell, aerial coolers, and fired heaters is found in the *Hydrocarbon Processing* articles by Driedger [9,10].

7.5.2 The exothermic reactor temperature control loop

The exothermic reactor is perhaps the most difficult process to control, owing to its instability and extreme nonlinear response [8]. A chemical reactor is quite often an exothermic process where some feedstock and catalyst are mixed together, and the

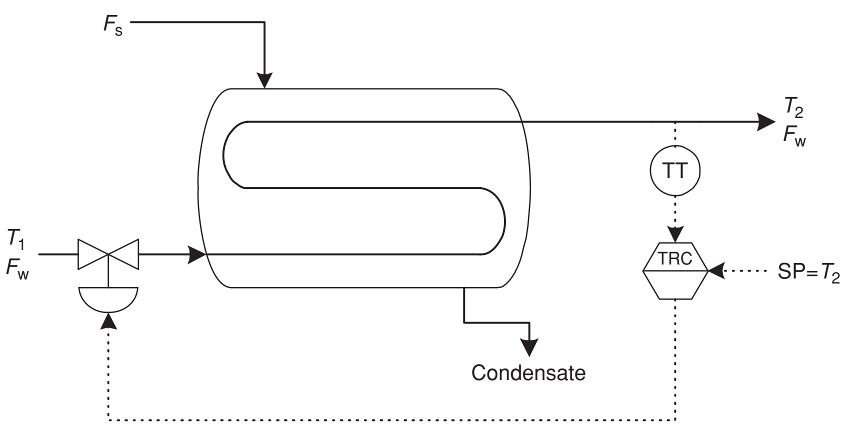

Figure 7.25 Temperature control scheme for capacity-limited exchangers

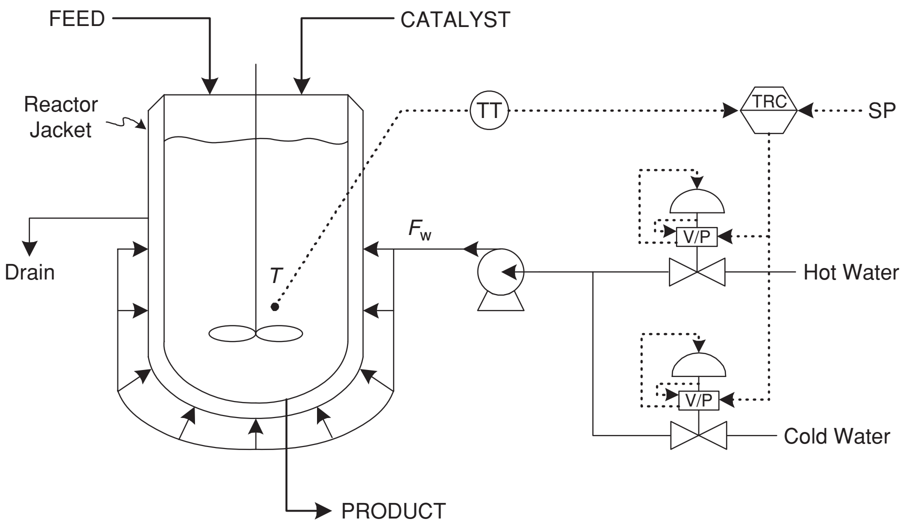

Figure 7.26 Control scheme for an exothermic reactor

temperature must be controlled at a specific set point. A typical temperature control scheme is illustrated in Figure 7.26.

The degree of stability that can be achieved in this temperature control loop depends on the rate at which the heat can be removed from the reactor. In other words, the reactor can be stabilized if the reaction temperature changes fairly slowly when compared with the rate at which the jacket temperature changes. The idea of the control loop shown in Figure 7.26 is that, once the feedstock and catalyst are added, hot water in the jacket is used to initiate the reaction. As the reaction temperature increases, the controller output decreases, closing the hot-water valve (which is air to open) and opening the cold-water valve (which is air to close). A valve positioner, V/P, is used to minimize valve hysteresis. The pump and multiple water inputs to the jacket are used to minimize dead time and to change the jacket temperature as quickly as possible, i.e. minimize the time constant of the jacket.

Typically, a PID controller is used, but using a proportional-only controller may stabilize the reactor provided that the reactor is the dominant single capacitance in the loop and there is no appreciable dead time (similar to a gas pressure loop).

Extremely fast reactions are sometimes carried out in semi-batch fashion to prevent a runaway temperature. In this case, one reagent is continuously added to a cooled reactor containing the other reagent or catalyst, and the addition rate is controlled to maintain a given batch temperature or a given heat removal rate. For safe operation, the temperature is kept high enough to ensure that a low concentration of the added reactant is required.

Often, an emergency control system (override controls) is implemented to stop the reaction by dumping the charge or stopping the catalyst flow in case the main control fails to halt a runaway reaction temperature. Other typical reactor control schemes can be found in the references [3,8].

In addition to these problems, there is also the problem of the extreme nonlinear process response. As before, the gain is a function of the temperature operating level. For the control system shown, the gain is defined by

$$K_{\mathrm{p}} = \frac{\Delta T}{\Delta F_{\mathrm{w}}} \tag{7.15}$$

In this equation, T is the reactor temperature and F_{w} is the cooling/heating water flow to the reactor jacket. Assuming perfect mixing in the reactor and a constant rate of heat evolution, the gain can be approximated as shown in Equation 7.16, where Q is the rate of heat generation:

$$K_{\mathrm{p}} \approx \frac{Q}{F_{\mathrm{w}}^2} \tag{7.16}$$

Owing to this nonlinearity, as well as to the problems mentioned earlier, some exothermic reactors are controlled with advanced control techniques, such as feed-forward, model reference or adaptive control [11].

7.6 Pump control

The flow and pressure of streams discharging from pumps must be controlled. Throttling a discharge valve on a centrifugal pump or manipulating the recirculating valve on a reciprocating pump are both reasonable means of control. The most efficient control method is to use a variable-speed motor to control the output of the pump [12].

7.7 Compressor control

Simply stated, compressors are employed whenever a gas at a certain pressure in one location is required to be at a higher pressure at another location. However, this belies the fact that compressors are major ticket items in the capital cost of a chemical or petroleum plant. For example, a large centrifugal compressor with a gas turbine is an investment of many millions of dollars.

Two major types of compressor are commonly used in chemical and petroleum plants, i.e. reciprocating and centrifugal compressors.

7.7.1 Reciprocating compressor control

Control of reciprocating compressors [13] involves the control of compressor capacity, engine load and speed, the control of auxiliary items on the compressor package, and the control of compressor safety.

Control of compressor capacity, engine load and speed

Compressor capacity is controlled by varying the driver speed, opening or closing fixed- or variable-volume clearance pockets, activating pneumatic suction valve unloaders, bypassing gas back to suction, or varying suction pressure. Driver speed control is not always possible with synchronous AC motors, although solid-state devices are now available for varying input frequency and speed. There is not space, and it is outside the scope of this book, to go into the details of these electromechanical control mechanisms. The interested reader is directed to the Gas Processors Suppliers Association's *Engineering Data Book* [14].

Control of auxiliary items on the compressor package

Oil, water and gas temperatures, oil, water and scrubber liquid levels and fuel and starting gas pressures need to be controlled.

Control of compressor safety

Safety shutdown controls must also be provided in case of harmful temperatures, pressures, speed, vibration, engine load and liquid levels.

7.7.2 Centrifugal compressor control

As mentioned previously, a large centrifugal compressor with a gas turbine as a driver is typically a multi-million dollar investment. A dedicated computer control system is usually employed to monitor multiple operating parameters, and that is specifically designed for the purpose. Such a control system is shown in Figure 7.27. The control of a centrifugal compressor involves the control of capacity, the prevention of surge and the protection of equipment.

Capacity control

The means available for controlling compressor capacity are suction throttling, discharge throttling, recirculation, variable guide vanes and motor speed control. Most of these controls are used in practice, and which is best depends on the application. Some of the pros and cons of these alternative control methods are as follows:

1 *Suction throttling.* The spinning vanes of a centrifugal compressor sling the gas outward. The centrifugal force develops a pressure proportional to the density and to the speed squared. Suction throttling reduces the density, and hence the ΔP; thus, the machine operates as a constant-pressure ratio machine.

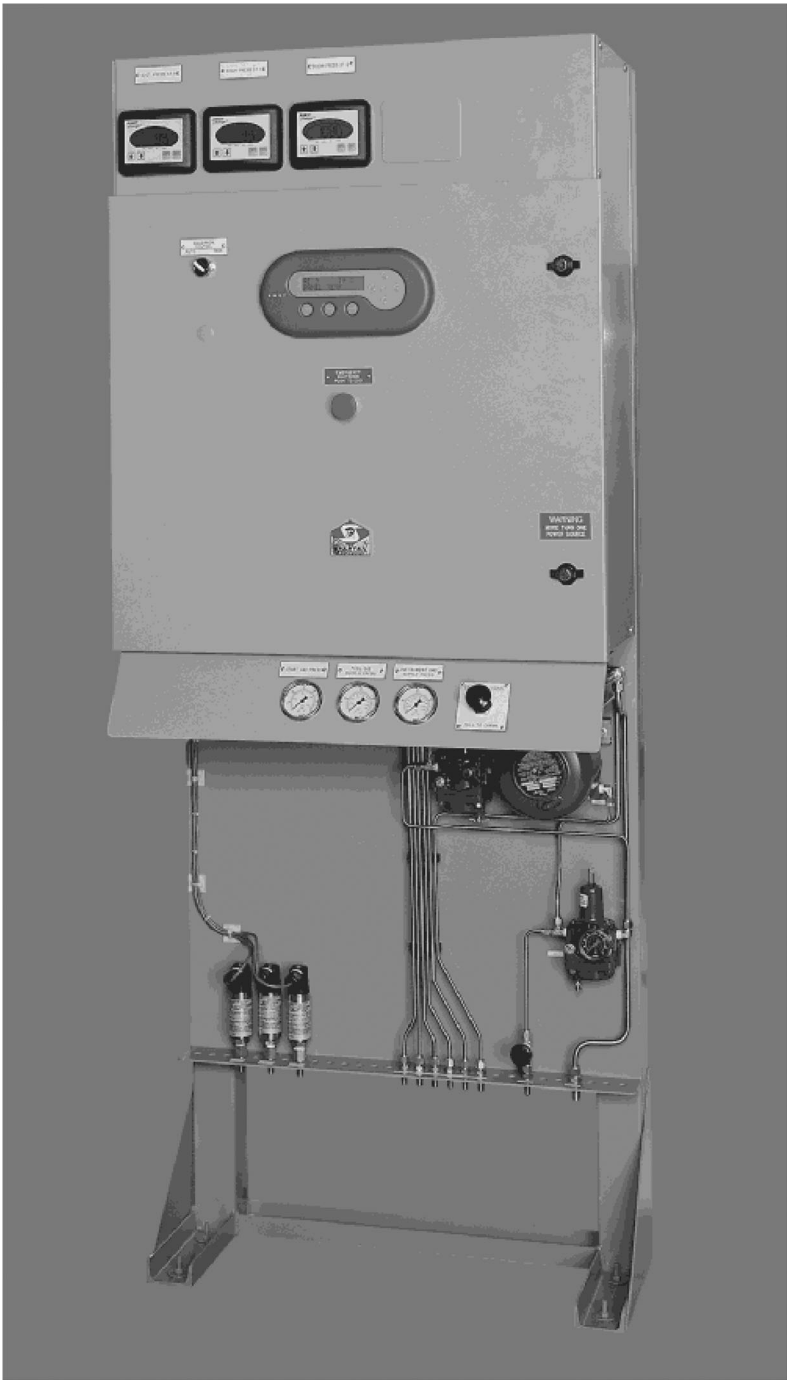

Figure 7.27 A compressor control station (courtesy of Spartan Controls Ltd)

2 *Discharge throttling.* This is much like suction throttling, but it is less efficient because the increased temperature means that the volume is more than P_1/P_2; therefore, $V\,\Delta P$ loss is greater at discharge. Therefore, discharge throttling is never used.

3 *Recirculation.* Recirculation has a much lower efficiency for similar reasons, but is essential for low turndown (low flow rates).

4 *Variable guide vanes.* These work by directing the gas flow with respect to blade rotation. Theoretically, there is no efficiency loss, but they tend also to act as inlet throttlers. Although excellent, this approach involves extra expense.

5 *Motor speed control.* This is cheap on turbine and engine drives, but it is expensive for electric motors.

Each of the methods of control affects the compressor curve to produce a set of curves called the compressor map (Figure 7.28). The various curves show the compressor characteristic at different values of the parameter being varied, such as inlet valve setting or speed. At each crossing of a compressor curve and a system curve, specific operating points occur which, collectively, establish the controllable range.

If the load on the compressor changes, i.e. the system curve changes, as shown in Figure 7.29, then the operating point moves along the compressor curve. The range within the compressor curve is called the operating range (shown as points 1 to 2 in Figure 7.29).

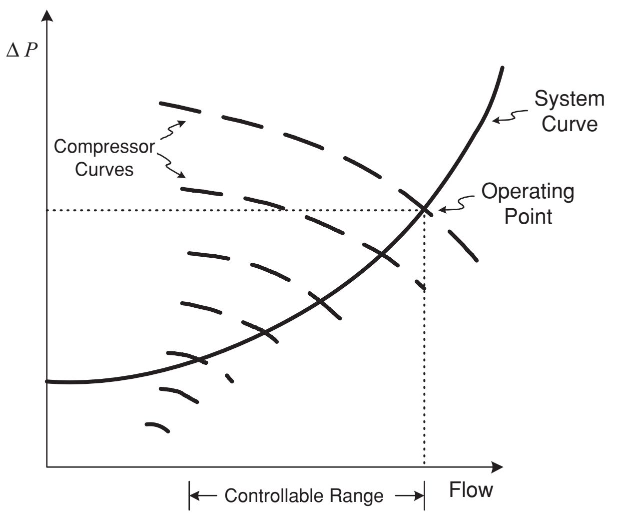

Figure 7.28 A typical compressor map

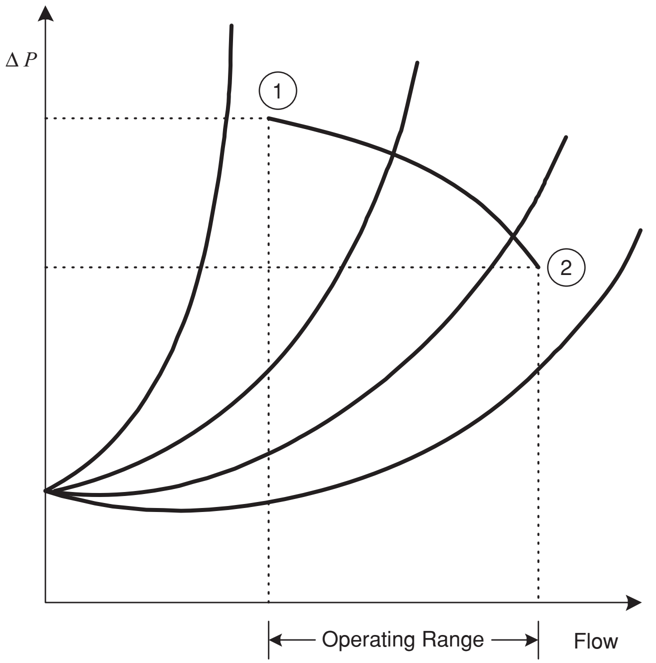

Figure 7.29 Compressor load changes

Returning to Figure 7.28, it can be seen that each compressor curve shows two end points. The lower right end point is relevant when discussing capacity control. Beyond the lower right end point the volume is so great that the internal flow velocity approaches sonic. A further drop in discharge pressure cannot affect the inlet flow; therefore, the flow rate no longer increases – this phenomenon is termed *stonewalling*.

Surge prevention

Beyond the upper left end point of Figure 7.29 the ΔP drops to a minimum and then rises again. This causes severe oscillations, known as *surge*. As the pressure rises up the curve it eventually reaches a maximum. The pressure cannot fall unless some of the gas flows out of the discharge volume or into the inlet volume.

The symptoms of surge are pulsating pressure, rapid flow reversals, a drop in motor current and a jump in turbine speed. Continuous, rapid flow reversals can cause severe damage to the compressor. In axial compressors the blades may touch, resulting in instant destruction. However, centrifugal compressors are more rugged, and only seal damage results [15].

The frequency of surge varies from 5 to 50 Hz. Suction and discharge volumes also influence surge. Minimizing the volume that has to be depressured can mitigate surge.

Preventing ΔP from getting too high also prevents surge. Surge protection involves the determination of the surge limit line, i.e. the limiting values of ΔP versus throughput that can initiate surge. Surge control keeps the compressor from crossing a surge control line that is arbitrarily set at a safe distance from the surge limit line [16].

The above compressor control theory is applied in the following example [15]. For more details on centrifugal compressors and their control, the interested reader is again directed to the Gas Processors Suppliers Association's *Engineering Data Book* [17] or the ISA Instructional Resource Package on *Centrifugal and Axial Compressor Control* [18].

Application example

The application to be considered is a plant with a compressor drawing vapour from the top of a distillation column and moving this vapour to downstream processing units. The plant also has a considerable amount of waste heat in the form of steam; therefore, it is economically worthwhile to use steam turbines as drivers with superheated steam as the motive force. The schematic of the example plant is shown in Figure 7.30.

In order to control the compressor, its purpose in terms of a process variable needs to be known. The purpose of the compressor in this application example is to control the pressure at the top of the column. A suitable measuring instrument would be a pressure transmitter located at the knock out (KO) drum. The compressor throughput is controlled by speed control on the steam turbine. Steam turbines generally have

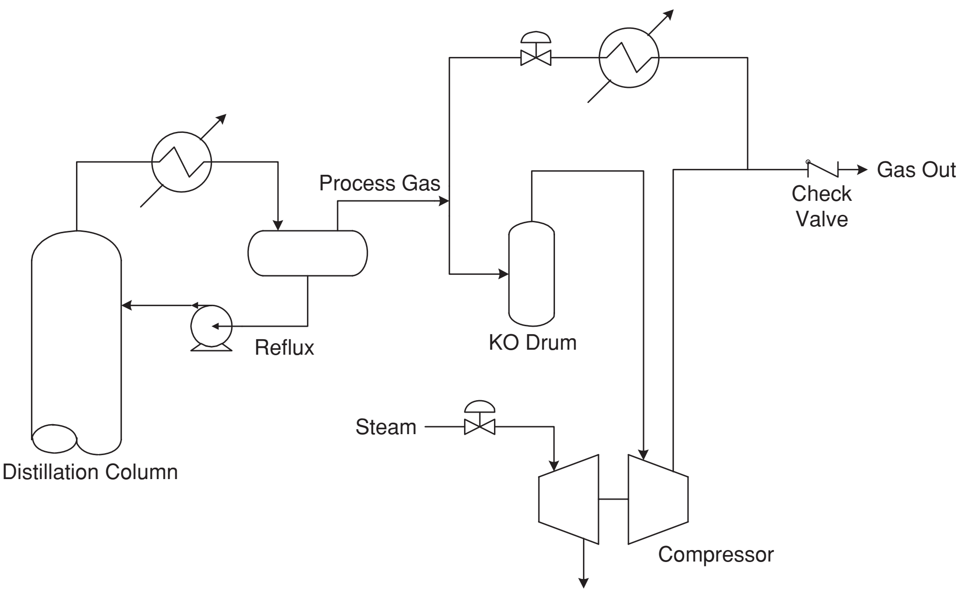

Figure 7.30 Compressor control example process schematic

special control valves that are an integral part of the machine and usually have their own governors. The pressure controller provides a set point for the governor.

Excess flow control (stonewall) protection is not needed as long as the compressor is not grossly oversized and the downstream process will provide sufficient backpressure to prevent excess flow. The process fluid is a light hydrocarbon and is never vented directly to atmosphere.

However, minimum flow (surge) protection is needed, as every compressor needs surge protection. The surge loop is placed as close as possible to the discharge. A check valve is placed downstream of the recycle *tee* to prevent recycling the entire downstream process flow.

The recycle line returns to the suction KO drum. A cooler must also form part of the recycle loop, as there is no other way of removing the energy that accumulates as heat of compression.

In order to control surge, the compressor map must be known. From the fan laws we know that flow varies proportionally with speed and ΔP varies with speed squared:

$$F \propto n \tag{7.17}$$

$$\Delta P \propto n^2 \tag{7.18}$$

From this we can calculate a family of curves based on the original compressor curve. These curves can be well fitted by a cubic equation. Surge occurs at the maximum or flat part of the curve. Applying the fan laws and solving for the maxima results in a quadratic equation called the *surge line*. To avoid surge, the compressor never operates to the left of the surge line, so the square of the flow must be greater than proportionality constant times the ΔP.

$$F^2 > k\Delta P \tag{7.19}$$

In order to provide surge control, the suction flow and ΔP must be measured. Suction flow must be in terms of actual units at the inlet, not standard volume units. The effects causing surge are based on gas velocity, not mass flow. These measurements are made as follows. ΔP is measured across the compressor. A venturi, which has by definition output proportional to the square of the flow, is placed in the compressor suction. It is important that the flow transmitter does not apply a square root to provide a linear signal, so that it may be used directly in the surge controller without further squaring.

In order to apply these process measurements, the compressor map is cast into a new form, ΔP versus the square of the flow, which results in a straight line for the surge line. However, it is not a good idea to use the surge line as the set point to the surge controller because of instrument error, transmitter, controller and valve delays, compressor variations with time and molecular weight variations. Instead, a surge control line is established, perhaps 5 per cent to the right of the actual surge line, as a safety factor.

The resulting, complete compressor control system with pressure/speed and surge loops is shown in Figure 7.31.

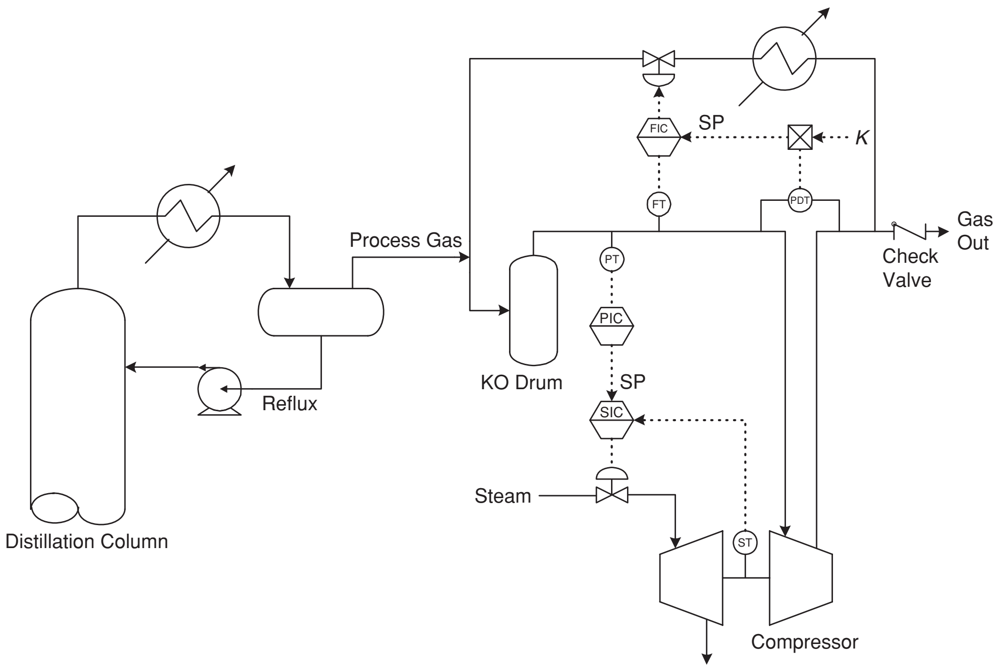

Figure 7.31 Complete compressor control system schematic

As always, it is important to verify the control scheme dynamically with the use of a suitable dynamic simulator. Other application examples that are documented in the literature [11,19,20] include a substantial emphasis on the importance of dynamic simulation for control scheme design validation and performance evaluation.

7.8 Boiler control

Boilers produce steam for power generation and heat; this is referred to as cogeneration. To control boilers, one requires complete combustion without too much excess air. The boiler's water level must be maintained by setting the feed water flow in equal to the steam flow rate out. The boiler must be able to control steam pressure and the temperature of superheated steam as demand fluctuates [12].

7.8.1 Combustion control

The control scheme must burn all the fuel with minimal excess air. There is a risk of too little air, resulting in carbon monoxide production from partial combustion and excess fuel. This excess fuel is not only expensive, it is also dangerous, as it may explode if the air flow is increased.

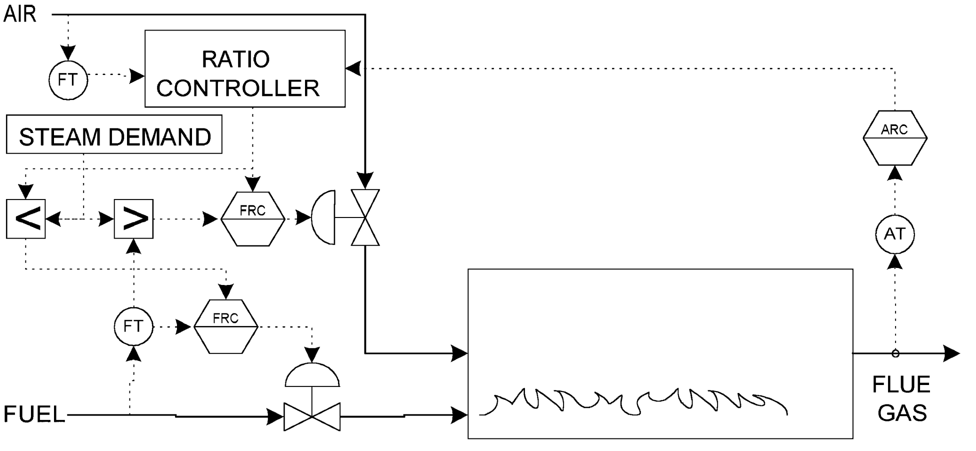

Figure 7.32 Combustion control scheme

If there is too much air, then the production of carbon monoxide is minimized. However, the excess air that is not used in combustion cools the flue gas, resulting in less efficient heat transfer to the boiler.

Optimally, the flue gas should be composed of 0.5 to 2 per cent oxygen, and carbon monoxide production should remain in the parts per million (ppm) [12]. These flue gas composition values depend on many variables, including the quality of fuel used, the boiler's condition and the steam demand in the plant.

For steady-state control the carbon monoxide and oxygen in the flue gas should be controlled.

For unsteady-state operation the control scheme is more complex. A high selector is used to set the set point on the airflow controller, and a low selector is used to control the ratio controller on the fuel side (Figure 7.32) [12]. For an increase in heat demand, the demand signal will be higher than either the fuel or air flow measurements. This will be passed by the high selector (>) to the airflow controller [12]. The increase in air flow will then be transferred through the low selector (<) to the ratio controller on the fuel side [12]. This ensures no excess in fuel, since airflow leads fuel flow for an increase in steam demand.

When steam demand decreases, the demand signal will be passed via the low selector (<) to the fuel flow and the air flow will be lowered by the ratio controller. Air flow will lag fuel flow on a decrease in demand, ensuring no excess fuel will build up in the system [12].

7.8.2 Water drum level control

To understand how to control the level in the water drum for the boiler, one must first understand the shrink–swell phenomena of this process unit. If the demand for steam increases, then the pressure in the boiler decreases. Therefore, more water boils to steam and the vapour bubbles in the evaporator tubes increase in size. This expansion in bubble size and increase in boiled water lifts more liquid into the water drum, increasing the level. As the drum level increases, a controller will add less feedwater to compensate. However, this swell resulting from increased steam demand actually requires more feed water. To correct this error, a shrink–swell compensator can be used on the drum level measurement [12].

The shrink–swell compensator works by detecting that an increase in steam flow has occurred, corresponding to a decrease in pressure. Therefore, a negative correction is applied to the level signal. As the swell subsides, the pressure increases, and the corrective action is cancelled.

A decrease in steam flow would similarly result in a rising pressure. This increased pressure causes the water drum level to shrink, since apparent level decreases due to a smaller bubble size in the evaporating tubes. Once again the shrink–swell compensator would correct the response of the level controller for the water drum.

7.8.3 Water drum pressure control

The pressure control system is used to maintain the energy within the boiler at a constant value. Pressure depends on the demand for steam in the plant and the rate of steam generation in the boiler.

To quantify the demand for steam in the plant, the controller relies on the steam flow and pressure. If the steam demand increases, then the steam flow increases and the pressure decreases, calling for more generation of steam. This increase in firing rate demand increases the amount of fuel burned and maintains the thermal energy of the boiler during an increase in demand for steam.

7.8.4 Steam temperature control

Superheaters are used to raise the steam temperature above the saturation point to get superheated steam (Figure 7.33). As explained in Section 7.8.3, the firing rate of the boiler is controlled by steam demand in the plant. The heat generated in the boiler is first used in the evaporating tubes and then the superheating tubes.

So, if demand for steam is low, then there is a decrease in the boiler's firing rate and less energy is transferred to the superheaters. However, at high steam demand, more fuel is burned to increase the boiler's steam generation, making more energy available for the superheaters. Since the temperature of the superheatered steam must be maintained at a constant value, feed water is added to keep the temperature from getting too high. Although this feed water will decrease the boiler efficiency, it is important for improved temperature control.

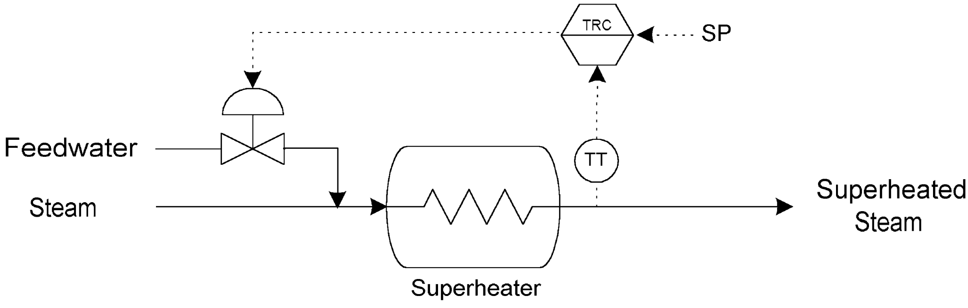

Figure 7.33 Control scheme for superheated steam

Control must always be balanced to ensure safety and maximize benefits while minimizing losses.

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8

Distillation column control

Steady-state simulation and design methods for separation processes, with emphasis on distillation, have been presented in detail in many references, a few of which are listed in the references for this chapter. This chapter will present a discussion of the basic control schemes for distillation columns. Let us start by stating the obvious: the amount of literature on separation processes, particularly distillation, is colossal. Particularly readable books and references are those by Buckley and co-workers [1,2], King [3], Tyreus [4], Seborg *et al.* [5], Shinskey [6], Smith and Corripio [7], Svrcek and Morris [8], and Wilson and Svrcek [9].

8.1 Basic terms

When determining the control system design for a multivariable process, the terms control strategy, control structure, and controller structure are used interchangeably. In this context, the meaning is the selection and pairing of manipulated and controlled variables to form a complete, functional control system. However, the three terms can also have individual meanings. Control strategy can describe how the control loops in a process are configured to meet a given overall objective, such as the purity of a given stream. Control structure, on the other hand, is the selection of controlled and manipulated variables from a set of many choices. Finally, controller structure means the specific pairing of controlled and manipulated variables by way of feedback controllers.

This chapter will describe a methodology for designing a multivariable control system that includes elements of control strategy considerations, control structure selection, and variable pairing. The methodology is largely empirical and based on general principles for distillation control. The methodology for control system design assumes that the process configuration is fixed and that changes are not possible. This is the case in many instances where control engineers are asked to design the control system for process configurations in an existing plant or a plant well into the design phase. The task for the control engineer is to select appropriate variables to be controlled and design controllers that will tie these variables to the control valves (manipulated variables) in such a way that the resulting controller structure meets the desired objectives. The final

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assumption is that the controller structure will be built up around conventional PID controllers, ratio, feedforward, and override control blocks found in all commercial DCS.

8.2 Steady-state and dynamic degrees of freedom

When a process engineer works with a detailed steady-state simulation of a distillation column, a certain number of variables have to be specified in order to converge to a solution. The number of variables that need to be specified, or degrees of freedom, can be determined through the concept of the description rule as stated by King [3]:

> In order to describe a separation process uniquely, the number of independent variables which must be specified is equal to the number which can be set by construction or controlled during operation by independent, external means.

Applying the description rule to a distillation column with a total condenser and two product streams gives two steady-state degrees of freedom. In this case the column would require two specifications, i.e. a composition and a component recovery. The steady-state simulator will then manipulate two variables, such as reboiler and condenser duties, in order to satisfy the specifications and close the steady-state material and energy balances. If a partial condenser is added to the column, then another degree of freedom is added to the steady-state column. Likewise, for each additional side draw added to the column, a new degree of freedom is added, requiring another specification.

When the same two-product distillation column is viewed in dynamics, the number of degrees of freedom increases from two to five. These three new dynamic degrees of freedom correspond to three new manipulated variables needed to control the integrating, inventory variables within the column that are not fixed by the steady-state material and energy balances alone. The inventory variables for this column are condenser level, reboiler level, and the column pressure.

There are restrictions on the control of a distillation column. The overall enthalpy balance limits the heat removed by the condenser and added by the reboiler. The rate of distillate produced may not exceed the feed rate. The number of stages in the column and the reflux ratio must be greater than or equal to the number required for the desired separation [3].

One control valve (or degree of freedom) must be used for each controlled variable. This relationship between controlled variables and degrees of freedom (or control valves or manipulated variables) is known as variable pairing and is an important concept in control system design.

When the five manipulated variables, which correspond to five valve positions as shown in Figure 8.1, are viewed, it can be seen that the two steady-state manipulated variables are a subset of the overall five. However, there is nothing about the heat duties that make them exclusive steady-state manipulators and prevent them from being used for inventory control. In many control schemes, the condenser duty is used for pressure control rather than composition control. For the same reason, any three of the five manipulated variables can be used to control the column inventories.

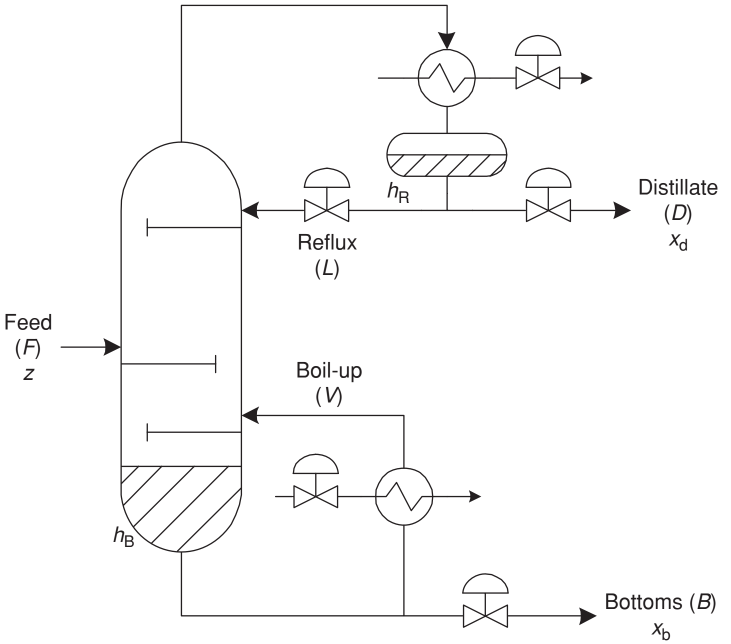

Figure 8.1 Basic distillation column schematic

Although the previous paragraph describes the manipulated variables as control valves, there are many choices available other than just the individual valves. For example, many columns have reflux ratio as a manipulated variable for either inventory or composition control. When ratios and linear combinations of variables are included, the choice of a manipulator for a given loop broadens considerably for a simple two-product column. However, the steady-state and dynamic degrees of freedom remain unchanged as two and three respectively, totalling five.

One must take care in determining the number of steady-state and dynamic degrees of freedom for more complex columns. Tyreus [4] describes the determination of the degrees of freedom for an extractive distillation system and for an azeotropic column with an entrainer. In the case of an extractive distillation system, recycle streams reduce the dynamic degrees of freedom through an increase in the steady-state degrees of freedom if the recycle contains a component that neither enters nor leaves the process. Also, if it is important to control the inventory of a *trapped* component, such as an entrainer for azeotropic distillation, then it is necessary to provide extra control valves to account for the loss of degrees of freedom. The loss comes from the addition of a side stream.

In summary, the total degrees of freedom for actual plant operation equals the number of valves available for control in that section of the plant. To find out how many integrating variables, i.e. pressures and levels, are to be controlled with the available valves, subtract the degrees of freedom required for steady-state control from the total degrees of freedom.

8.3 Control system objectives and design considerations

Defining and understanding the control system objectives should be a collaborative effort between process engineers and control engineers. Left to either of these contributors alone, the objectives can be severely biased. The control engineer might be tempted to make the control system too complex in order for it to do more than is justified based on existing disturbances and possible yield and energy savings. On the other hand, a process engineer might underestimate what process control can achieve and thus make the objectives less demanding. It is crucial to define what the control system should do, as well as to understand what disturbances it has to contend with.

Process understanding is another key, but often overlooked, activity for successful control system design. In practice, more time is spent designing and implementing algorithms and complex controllers than on analysing process data and understanding how a process really works. Modelling and simulation are integral parts in the process understanding step.

Rigorous dynamic simulation is the third important activity in control system design. A flexible dynamic simulator allows for rapid evaluation of different control structures and their responses to various disturbances. In choosing a control scheme, there are several design considerations to take into account. First, it is important to remember that a distillation column performs two basic functions:

1 feed split
2 fractionation.

The feed split is the primary point of separation between the overhead and bottoms product. Fractionation is determined by the number of separation stages in the column and the energy input. Figure 8.2 illustrates these concepts with a mixture of a low-boiling-point component (light shading) and a high-boiling-point component (dark shading). The boiling point of the distillation products is determined by how much of each component is present in each product. As the distillation feed split changes, the line will shift left or right. As the fractionation changes, the slope of the line will

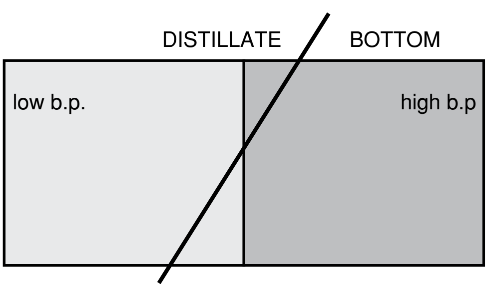

Figure 8.2 Feed split and fractionation

change, with a steeper slope representing better separation. It is important to realize that fractionation increases the purity of both products simultaneously, whereas changing the feed split will make one product more pure and the other less pure.

Once the inventory variables are controlled there are two degrees of freedom left in the case of the column shown in Figure 8.1. One degree of freedom should be used to control the feed split and the last available degree of freedom controls fractionation. Feed split has a much more significant effect on the product compositions than fractionation. Therefore, after the inventory and capacity variables have been paired, the primary controlled variable is normally used to set the feed split and the secondary controlled variable is used to set fractionation.

The following equations describe how the various manipulated variables are related and show that virtually any variable pairing can be used to achieve the desired control objectives. However, some pairings will provide significantly better sensitivity and responsiveness. Better sensitivity means that the control scheme will react with smaller changes, whereas a more responsive control scheme reacts more quickly.

$$\text{Overall material balance:}\quad F = D + B \tag{8.1}$$

$$\text{Component balance:}\quad Fx_{\mathrm{f}} = Dx_{\mathrm{d}} + Bx_{\mathrm{b}} \tag{8.2}$$

where F is feed, B is bottoms, D is distillate, x_i is the concentration of a particular component in the feed, distillate or bottoms, Q_{reb} is the reboiler duty, and Q_{cond} is the condenser duty.

$$\text{Energy balance:}\quad FH_{\mathrm{f}} + Q_{\mathrm{reb}} = DH_{\mathrm{d}} + BH_{\mathrm{b}} + Q_{\mathrm{cond}} \tag{8.3}$$

Where H_{f} is the enthalpy of the feed, H_{b} is the enthalpy of the bottoms, and H_{d} is the enthalpy of the distillate.

Combining Equations 8.1 and 8.2 to eliminate B or D gives

$$D = F(x_{\mathrm{f}} - x_{\mathrm{b}})/(x_{\mathrm{d}} - x_{\mathrm{b}}) \tag{8.4}$$

$$B = F(x_{\mathrm{f}} - x_{\mathrm{d}})/(x_{\mathrm{b}} - x_{\mathrm{d}}) \tag{8.5}$$

The control system must satisfy Equations 8.1 and 8.3 at all times. For particular values of x_{d} and x_{b} (i.e. composition specifications), Equation 8.4 or 8.5 also has to be satisfied. D, B, Q_{reb}, and Q_{cond} can all be fixed or adjusted dynamically by control valves on the flow rate or utility streams. The reflux flow can also be adjusted dynamically and will directly affect the energy balance.

One of the most difficult aspects of distillation column control is the interaction effects between the material and energy balance and composition controls. Depending on the inventory controls, a heat input or removal can alter both the material draws and the compositions. This interaction can work for us or against us, depending on the control strategy.

Another point to consider when choosing a column control scheme is that, typically, the process gains from a high-purity separation are very nonlinear. This can be verified by simply using the component balance equations. For example, Equation 8.4 can be rearranged and differentiated at constant x_b to give

$$\left(\frac{\partial x_d}{\partial D}\right)_{x_b} = -\frac{F\,(x_f - x_b)}{D^2} \tag{8.6}$$

Equation 8.6 shows that changes in the distillate rate D will have a much larger effect on the distillate composition x_d when the distillate rate is relatively low compared with cases when the distillate rate is relatively high.

A final and important consideration to keep in mind is the dead time that may be present in the column. In Chapter 3, dead time was described as being generated by a series of lags (material or energy capacitances). It is easy to see how a distillation column with its multiple stages can generate dead times. The control scheme on a distillation column should be set up to minimize the dead times with respect to the process lags and disturbances.

The steps for determining a suitable controller structure are as follows:

1. Define the objectives of the control system and the nature of the disturbances.

2. Understand the principles of the process in terms of its dynamic behaviour.

3. Propose a control structure consistent with the objectives and process characteristics.

4. Assign controllers and evaluate the proposed control structure with anticipated disturbances through the use of dynamic simulation.

Ultimately, the importance of process control is seen through increased overall process efficiency, allowing the plant engineer to get the most from the process design. This is especially true of distillation control. Most distillation columns are inherently flexible, and a wide range of product yields and compositions can be obtained at varying levels of energy input. A key requirement of any control system is that it relates directly to the process objectives. A control system that does not meet the process objectives or produces results that conflict with the process objectives does not add value to the process.

8.4 Methodology for selection of a controller structure

The economic performance of a distillation system is linked to its steady-state degrees of freedom. In other words, the economic benefits of a column control scheme depend on how well it controls composition, recovery, or yield and not on how well it holds integrating variables such as levels and pressures. The integrating variables (i.e. inventories of level and pressure) must obviously be controlled, but their control performances do not directly translate into profits. However, inventory controls can be the most troublesome of all loops and can preoccupy the operators to the point where the economically important composition and recovery are neglected. This problem has been resolved by designing the level and pressure controls before dealing with the composition controls [2]. However, one must be careful in the selection of the manipulated variables for inventory control as they can significantly impact the control performance of the composition loops.

The following methodology [4] can be employed to define a control structure for the simple distillation column shown in Figure 8.1.

1 Count the control valves in the process to determine the overall degrees of freedom for control.

2 Determine from a steady-state analysis the steady-state degrees of freedom.

3 Subtract the steady-state degrees of freedom from the overall degrees of freedom to determine how many inventory loops can be closed with available control valves.

4 Design pressure and level controls and then test for disturbance rejection.

5 Design composition controls based on the product stream requirements. It is important that the manipulated variable chosen can control the feed split.

6 Design optimizing controls with the remaining manipulative variables.

For the simple distillation column in Figure 8.1 there are five degrees of freedom, which translates into five independent valves from a control point of view. In this 5×5 system, there are 120 possible SISO control combinations of controlled and manipulated variables. Fortunately, most of these combinations are not useable due to various constraints, such as economics. From a steady-state degree of freedom analysis there are only two degrees of freedom, since a total condenser is assumed. If the column had a partial condenser then there would, of course, be three degrees of freedom instead of two. Inventories that must be controlled are the reflux drum level h_R, level in column base or reboiler h_B, and the column pressure (vapour holdup). The remaining two variables are used to control the feed split and the fractionation.

The feed split is simply the amount of feed that leaves as distillate versus the amount that leaves as bottoms. The other variable, fractionation, is the amount of separation that occurs per stage. The overall column fractionation depends on the number of stages, the energy input, and the difficulty of separation. A typical control scheme for this column is shown in Figure 8.3.

The most convenient method of verifying the operability of a proposed multivariable control scheme is through dynamic simulation. However, to use dynamic simulation effectively it is first necessary to define the objectives of the control system, define the

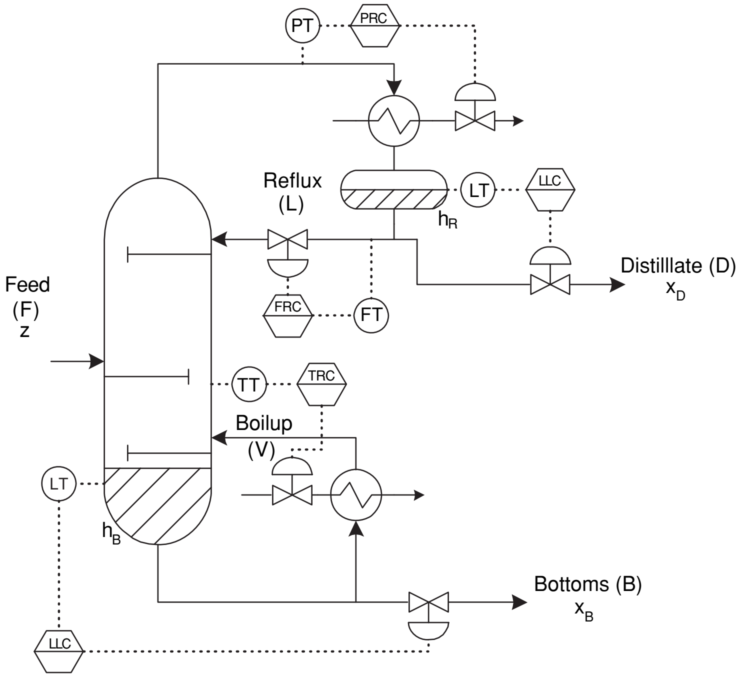

Figure 8.3 Column basic control scheme

nature of the expected disturbances, and develop a basic understanding of the process both in terms of its steady-state and dynamic behaviour.

8.5 Level, pressure, temperature and composition control

Measurement of fractionating column variables must be within certain tolerances of accuracy, speed of response, sensitivity and dependability; they must also be representative of the true operating conditions before successful automatic control can be realized. The instrument equipment selected, the installation design and the location of the measuring points determine these requirements.

This section is concerned with selecting the specific location in a fractionating column of the measuring points that will provide the best automatic control under variable process operating conditions. Specifically, level, temperature, pressure and composition measuring points in conventional fractionating columns are discussed. It should be clearly understood that this discussion, which is general in nature, is intended only to serve as a guide, from which detailed recommendations may be formulated and tested through dynamic simulation.

Locating temperature, pressure, flow and composition measuring points for automatic control systems depends on the control scheme used and the static and dynamic interdependence of these variables. The control scheme utilized is usually determined by the source of energy or process stream to be manipulated to control a particular variable. Therefore is it, important to consider the static measuring sensitivity of the instrument selected to measure the controlled variable. Measuring sensitivity should generally increase with requirements of control precision by the use of narrow-span suppressed-range instruments. In addition, the location of the measuring element with respect to the energy source and the time lag involved for it to sense effects of changes in manipulated variables will determine dynamic measuring lags introduced by changing process conditions. The dynamic measuring lags will determine the quality and stability of the control scheme.

The interaction of temperature, pressure, and composition will differ with location in the column. The selection of a temperature control point in a fractionating column, which is determined assuming that the pressure and composition are constant, may be unsatisfactory when these variables are permitted to vary with changing process conditions, i.e. feed composition changes. The complex effect of all of the sources of disturbances in the form of changing process conditions on the measuring point must be considered for dynamic stability via dynamic simulation.

8.5.1 Level control

Level control was discussed in detail in Section 7.3. Typical hold-up times for condenser accumulators and reboilers are of the order of 5–10 min and 20 min (large enough to hold all the liquid from the trays if dumped) respectively. From a common sense point of view, to assign manipulative variables for level control, simply choose the stream with the most direct impact. For example, in a column with a reflux ratio of 100, there are 101 units of vapour entering the condenser and 100 units of reflux leaving the reflux drum for every unit of distillate leaving. Therefore, the reflux flow or vapour boil-up should be used to control the drum level. If this assignment principle is not followed and distillate flow is selected for level control, it would only take a change of slightly more than 1 per cent in either vapour boil-up or reflux flow to saturate the drum level controller and saturate the distillate valve. The ‘Rule of 10’ can be applied. This rule states that if there is a 10-to-1 or greater difference, say reflux versus product, then the larger stream must be used to control the level.

8.5.2 Pressure control

Pressure control is a primary requirement for all towers because of its direct influence on the separation process. Columns are typically designed to operate at sub-atmospheric, atmospheric, or above atmospheric pressure. Tower pressure control configurations can also be required to vent varying amounts of inerts from the overhead accumulator. The venting of inerts or maintaining the desired operating pressure is often the crux of the control problem.

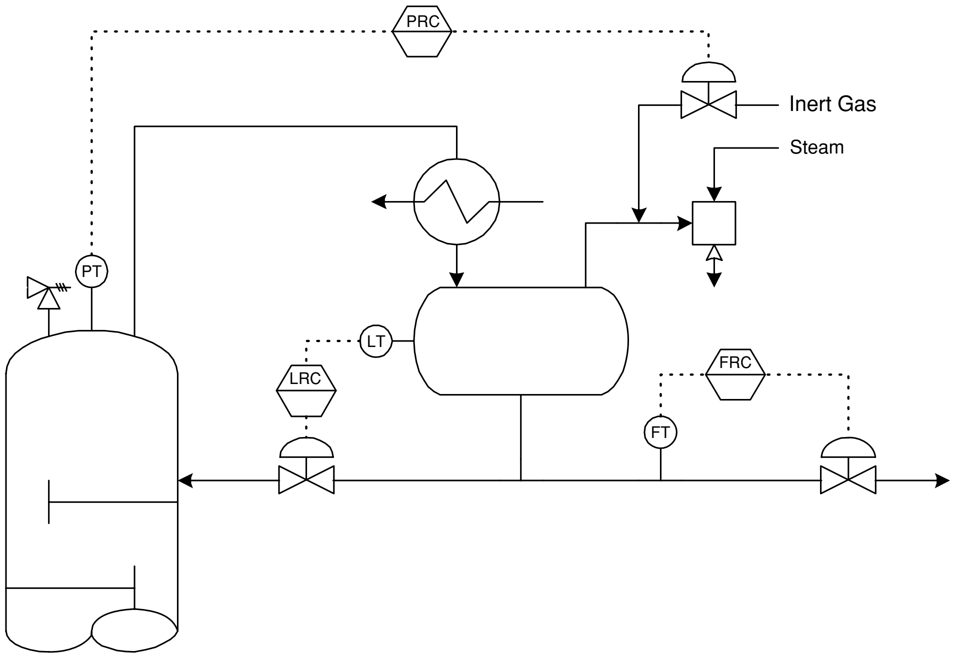

Figure 8.4 Pressure control for sub-atmospheric operations

The same general principle is followed when finding manipulated variables for pressure control as for level control. Column pressure is generated by boil-up and is relieved by condensation and venting. To find an effective variable for pressure control, it is necessary to determine what affects pressure the most. For example, in a column with a total condenser, either the reboiler heat or the condenser cooling is a good candidate for pressure control. On a column with a partial condenser, it is necessary to determine whether removing the vapour stream affects pressure more than condensing the reflux. Sometimes the dominating effect is not obvious. If the vent stream is small, then it might be assumed that the condenser cooling should be manipulated for pressure control. However, if the vent stream contains non-condensables, these will blanket the condenser and affect the condensation significantly. In this situation, the vent flow, although small, is the best choice for pressure control.

Figure 8.4 shows a typical pressure control scheme for sub-atmospheric column operation used for total condensing service. The eductor is not controlled by regulating the motivating steam, because the turndown on the jets is very limited. Rather, the capacity is controlled by regulating the addition of non-condensable gas. This method provides a smooth and rapidly responding control system.

Figure 8.5 shows a typical control scheme used for an atmospheric or above-atmospheric tower in a total condensing service with little or no inerts. In this situation the pressure is controlled by regulating the flow of the coolant, which in turn changes the condensing surface temperature and the vapour condensing rate. The pressure response

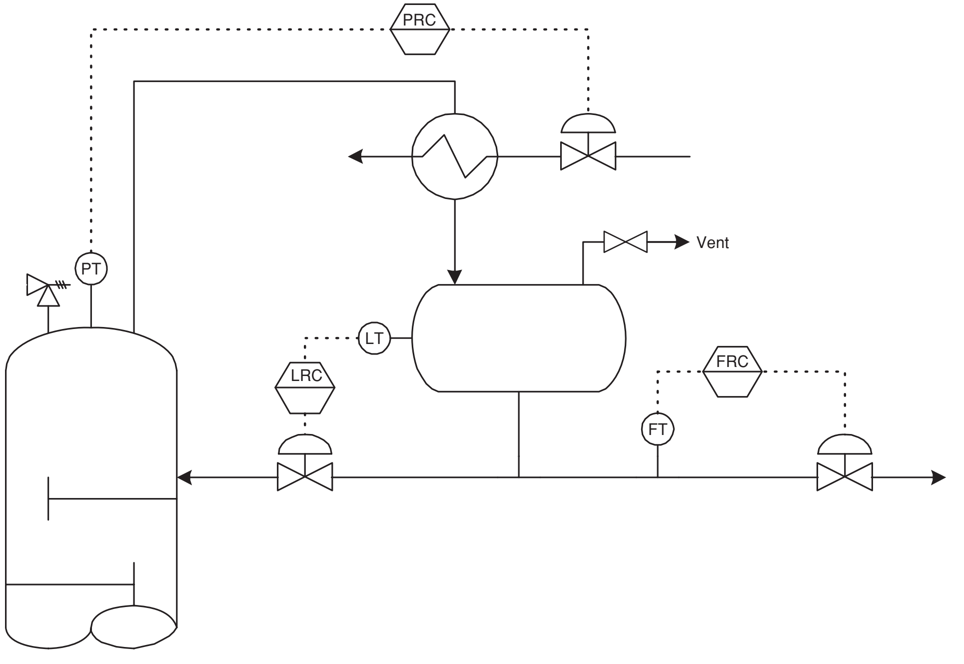

Figure 8.5 Pressure control for above-atmospheric operation

of this scheme to changes in the coolant flow rate is inherently slow in comparison with methods regulating vapour withdrawal directly and/or condenser surface area control.

Figure 8.6 shows a scheme where column pressure is controlled by regulating the flow of the vapour product from the accumulator. The reflux is on flow control. A level controller is required to control the coolant flow in order to maintain accumulator liquid inventory. This method provides a smooth, rapidly responding column pressure control.

In Figure 8.7 the column pressure is controlled by regulating the inert and vapour flow from the accumulator. The condenser coolant is fixed at a constant flow rate and should not be subject to change. A flow controller fixes the reflux rate and a cascade (level to flow) is used to adjust the product overhead rate. This cascade arrangement isolates the overhead product flow from internal column pressure disturbances that could affect the overhead product flow rate. Cascade control is only used if minimization of overhead product flow rate is critical to downstream unit operations.

For a total condensing service the column pressure can be controlled by varying the condenser level or the condenser surface area exposed to the column overhead vapours, as shown in Figure 8.8. The accumulator pressure and reflux temperature can also be controlled by providing a condenser vapour bypass or by controlling the coolant flow rate.

In a total condensing service, when varying quantities of inerts are present in a pressurized tower, it is often necessary to vent or alternatively inject a blanketing inert gas. This is normally accomplished using a split-range control scheme, as shown in

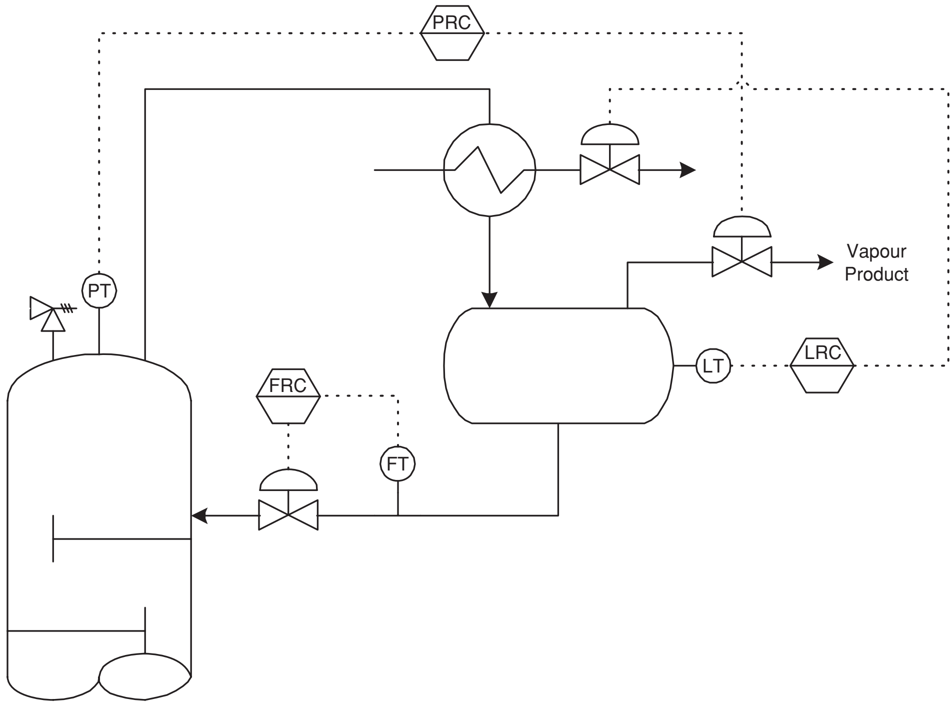

Figure 8.6 Pressure control by control of overhead product vapour flow

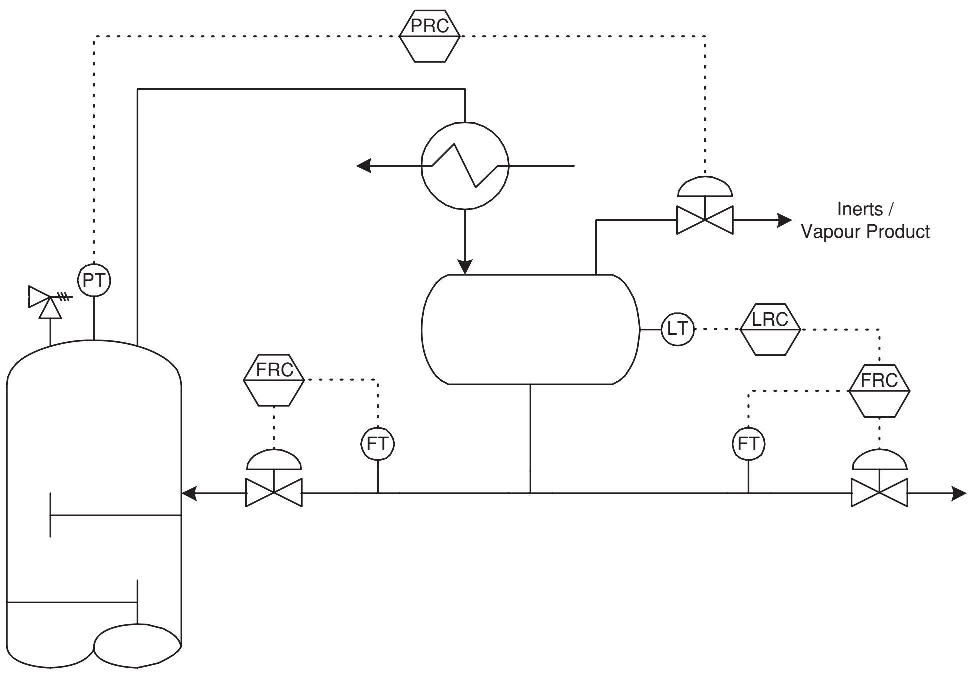

Figure 8.7 Pressure control by venting of inerts

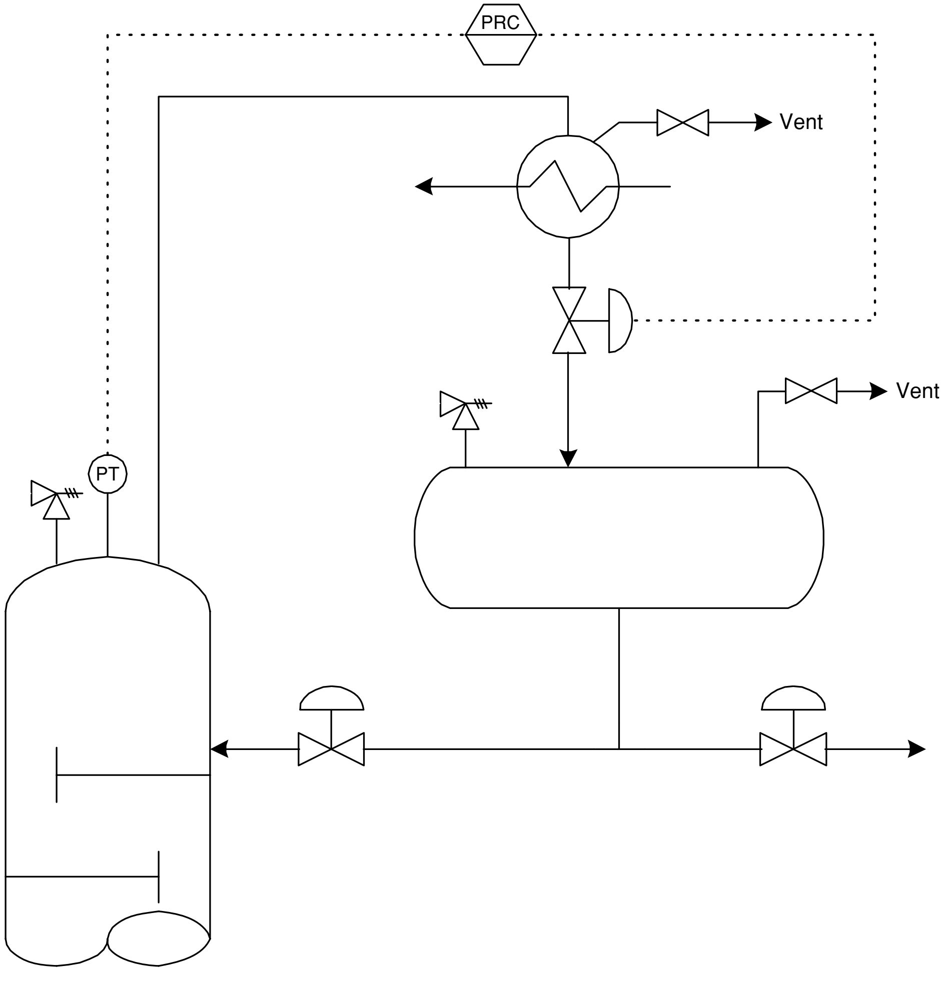

Figure 8.8 Pressure control by condenser level control

Figure 8.9. The column pressure is controlled by either injecting or venting blanketing gas from the accumulator/condenser.

The location of a column pressure control point is not restricted by dynamic considerations. The response time of pressure changes in a column and the dynamic measuring lag has been found to be equally fast for any location in the column when the manipulated energy source used to control pressure is either condenser cooling or reboil heat. Pressure is regulated at a constant value and is rarely used as a variable to control a product specification. Generally, temperature is used to control composition, making pressure compensation necessary to sustain accurate control. Fractionation is affected by changes in the relative volatilities of the components due to variations in pressure. A decrease in pressure may cause the feed to flash, resulting in two-phase feed and column flooding [10].

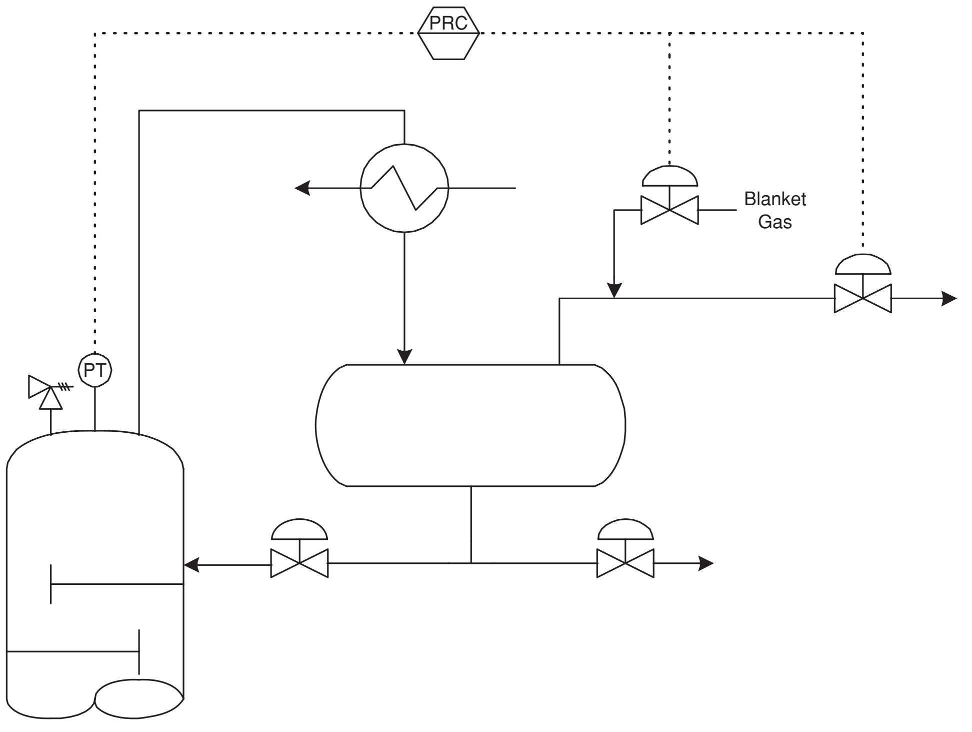

Figure 8.9 Pressure control by regulating condenser surface area with blanket gas

Some of the factors that should be considered in locating the pressure measuring point are:

1 When columns are operating near the relief valve pressure setting, the pressure measurement should be located near the relief valve.

2 Bottom pressure control of an atmospheric column via reboiler heat input will, in effect, control the column differential pressure and, thereby, the column vapour flow and tray vapour loading.

3 Providing temperature on a tray is a good indication of composition; pressure at this tray should be measured and controlled. This concept is explained in greater detail in the next section.

8.5.3 Temperature control

Composition control of products from a column is usually realized via temperature control. Temperature sensors are inexpensive, highly reliable, repeatable, continuous, and fast compared with composition sensors [11]. The measurement lag is particularly important in dynamic considerations. For temperature it is a fraction of a minute, whereas composition measurement by gas chromatography is of the order of 15–30 min. Infrared analysers are available that produce continuous composition estimates, but they are not always cost effective. Periodic checks of product composition by analytical means provide information which is used in setting the temperature control point. The accuracy of the correlation of column temperature to product composition depends on the sensitivity of the controlled temperature to composition changes and pressure variations at the temperature measuring point.

The sensitivity of the temperature measurement to key or major component composition changes for each tray can be determined if tray-by-tray composition changes are large and the other component changes are small. It must be determined which stage exhibits the composition-related temperature response in all disturbance situations. A sizeable temperature response must be present for all process variable changes to which the column will be subjected. Select a range of process disturbances and change these in short step sizes to compare the tray temperature profiles [12]. A temperature measurement in this area will give a good indication of composition provided that the effects of the pressure variations are small. Controlling pressure at the point or tray where temperature is controlled can eliminate pressure variations that have large effects on composition.

The temperature-composition correlations of key components are often affected by changes in the concentration of other components, i.e. column feed composition changes. If the magnitude of these changes can be estimated, then a calculation using equilibrium constants can be made to determine the effect on the temperature composition correlation. Then a control tray can be selected where the effect of non-key component variations is small.

Stable column temperature control, from the tray selected by the foregoing static considerations, depends on the dynamic measuring lag or response of the tray temperature with respect to the manipulated energy source used to control the temperature. Based on experimental tests, the following observations are cited for use as guides:

1 Temperature control is made less stable by thermowell and measuring instrument lag or response times.

2 The speed of response and control stability of tray temperature, when controlled by reboil heat, is the same for all tray locations.

3 The speed of response and control stability of tray temperature, when controlled by reflux, decreases in direct relation with the number of trays below the reflux tray.

4 When pressure is controlled at the temperature control tray, the speed of response of the temperature instrument can vary considerably with tray location, and is normally slower.

8.5.4 Composition control

The composition control loops on a column are the most important steady-state controls. The purpose of composition control is to satisfy the constraints defined by product quality specifications. These constraints must be satisfied at all times, particularly in the face of disturbances. The objective of composition control is then to hold the controlled composition as close as possible to the imposed constraint without violating the constraint. This objective translates to on-aim, minimum variance control.

To achieve good composition control, two things must be examined: process dynamics and disturbance characteristics. Process dynamics includes measurement dynamics, process dynamics and control-valve dynamics. Tight process control is possible if the equivalent dead time in the loop is small compared with the shortest time constant of a disturbance with significant amplitude. To ensure small overall dead time in the loop, it is necessary to find a rapid measurement along with a manipulated variable that gives an immediate and appreciable response. In distillation, a rapid measurement for composition control often translates into a tray temperature. A good manipulated variable is vapour flow, which travels quickly up through the column and usually has a significant gain on tray temperatures and indirectly on composition.

If the feed contains multiple components, fixing the temperature and pressure of a stage in the distillation column may not fix the composition. Therefore, a steady-state model may be used to compare advantages of using an online composition analyser rather than a temperature controller. Factors to consider are yield loss, energy consumption and dead time [11].

In situations where the apparent dead time in the composition loop cannot be kept small compared with significant disturbances, the disturbances themselves must receive the attention. Sometimes the important disturbances can be measured or anticipated, in which case feedforward control is a candidate. In other situations, the control loop structure can be rearranged to influence the way the disturbance affects the composition variable. Several researchers have proposed numerous algorithms for determining the disturbance sensitivity for different control structures. Tyreus [4] states that, in his opinion, direct dynamic simulation of the strategies resulting from assignment of the manipulated variables for pressure and level control gives the best insight into the viability of a proposed composition control scheme.

8.6 Optimizing control

After the inventory and composition controls have been assigned, there are typically a few manipulated variables remaining. These variables can be used for process optimization. Because process optimization should be performed on a plant-wide scale, in-depth discussion of this topic will be delayed until Chapter 10.

8.6.1 Example: benzene column with a rectifying section side stream

To illustrate better how the control strategy design method described is put into practice, consider the case of a liquid side draw benzene column.

Figure 8.10 shows the flowsheet configuration of a column with a rectifying section liquid side draw. The multi-component feed comes from an upstream unit in the process. The benzene liquid side draw is the product stream, and has a purity specification in terms of benzene. The distillation removes *n*-pentane from the feed mixture and the heavies (toluene, naphthalene and biphenyl) are purged from the reboiler. A small overhead purge stream is connected to the condenser for pressure relief.

The control scheme objective of this column is to operate close to the quality constraint of the liquid side draw product. The major disturbances are changes to the overall feed flow rate, as well as individual component feed flow rates.

The column has seven control valves and requires four degrees of freedom for steady-state control. The remaining three dynamic degrees of freedom are used to control the column inventories. Column pressure is controlled by manipulating the condenser duty. However, if there were non-condensables in the column, then the overhead vapour stream would have been a more suitable choice as a manipulated variable. Non-condensables in a column tend to accumulate in the condenser and significantly reduce the dew point of incoming vapours. The low dew point reduces heat transfer because of small temperature driving forces. Because the vent stream is rich in

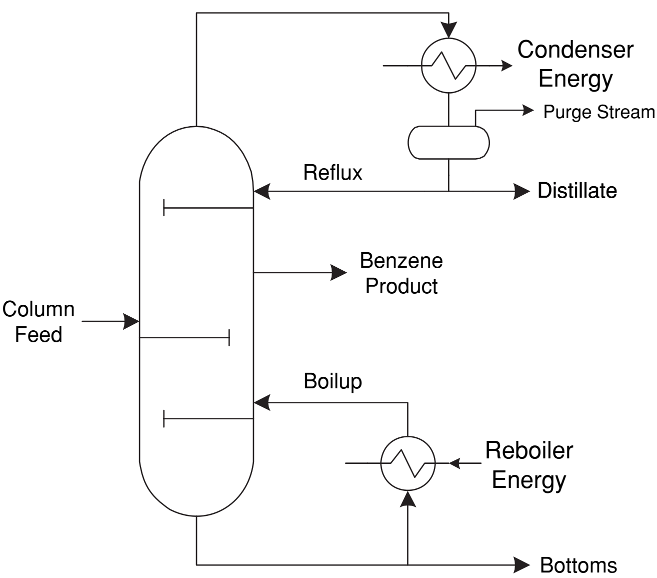

Figure 8.10 Liquid side draw benzene column

non-condensables, vent flow rate is an effective manipulator for removing the non-condensables and thereby quickly increasing heat transfer whenever needed.

Control of the reflux drum is fairly straightforward. Because the reflux ratio is very high, with a steady state value of 145, reflux flow is the only reasonable manipulator for drum level. However, there is a potential loss of one dynamic degree of freedom unless it is ensured that the material balance for the distillate product is satisfied. This can be achieved by ratioing the distillate flow to the reflux flow. The effective manipulator is now the distillate flow and the reflux flow combined instead of just reflux flow. Control of the base level in the column is basically restricted to the use of reboiler steam due to the large vapour boil-up to bottoms ratio.

At this point, the inventories in the system have been placed under control and composition control can be considered. However, first the side stream material balance must be considered. The condenser level control refluxes any disturbances in vapour flow rate back down into the column as liquid. On the other hand, the base level controller sends any disturbances in liquid flow back up the column as vapour. To prevent a build-up of side stream material in the column, a route must be provided for the side stream material to escape. This can be accomplished by ratioing the liquid side draw flow to the reflux flow.

Finally, a temperature controller can be added to provide a method of controlling the composition of the liquid side draw. This controller can have its temperature sensor on the bottom tray of the main tray section and use the bottoms flow rate as a manipulated variable. Temperature sensitivity analysis can be performed using the steady-state model to ascertain the proper location for the temperature sensor. Using the bottoms flow rate allows a method for excess heavies to be removed from the system in the event of a disturbance while retaining the target composition of the liquid side draw. The resulting control scheme for the liquid side draw benzene column is shown in Figure 8.11.

How does this control scheme respond to disturbances? The major control objective is to produce a side stream of essentially pure benzene, approximately 99 per cent. To test this control scheme, the column was subjected to two different types of disturbance. The first disturbance was that of an increase in the total volumetric flow rate of the feed introduced to the column. A strip chart of the feed flow rate, the three product flow rates and the side stream benzene purity is shown in Figure 8.12. A step change in the feed flow rate is introduced, increasing the flow rate from 11.5 m^3h^{-1} to 12.075 m^3h^{-1}. This corresponds to an increase of 5 per cent.

An increase in the overall volumetric flow rate of the liquid feed adds considerable liquid to the column system. Because the feed is primarily benzene, it is expected that the benzene side draw flow rate will increase. The flow rate overshoots and then assumes its new steady-state value. A similar-shaped curve exists for the distillate. This is expected due to the ratio control between the distillate stream and the reflux stream and the ratio control between the benzene side draw and the reflux stream. Throughout the overshoot in flow rates, the benzene purity in the side draw remains relatively constant. What is interesting to note is the response of the bottoms flow rate. Here, an inverse response is exhibited. The flow rate first decreases, then increases, overshoots, and finally assumes its new steady-state value. Why does the bottoms flow rate behave in such a manner?

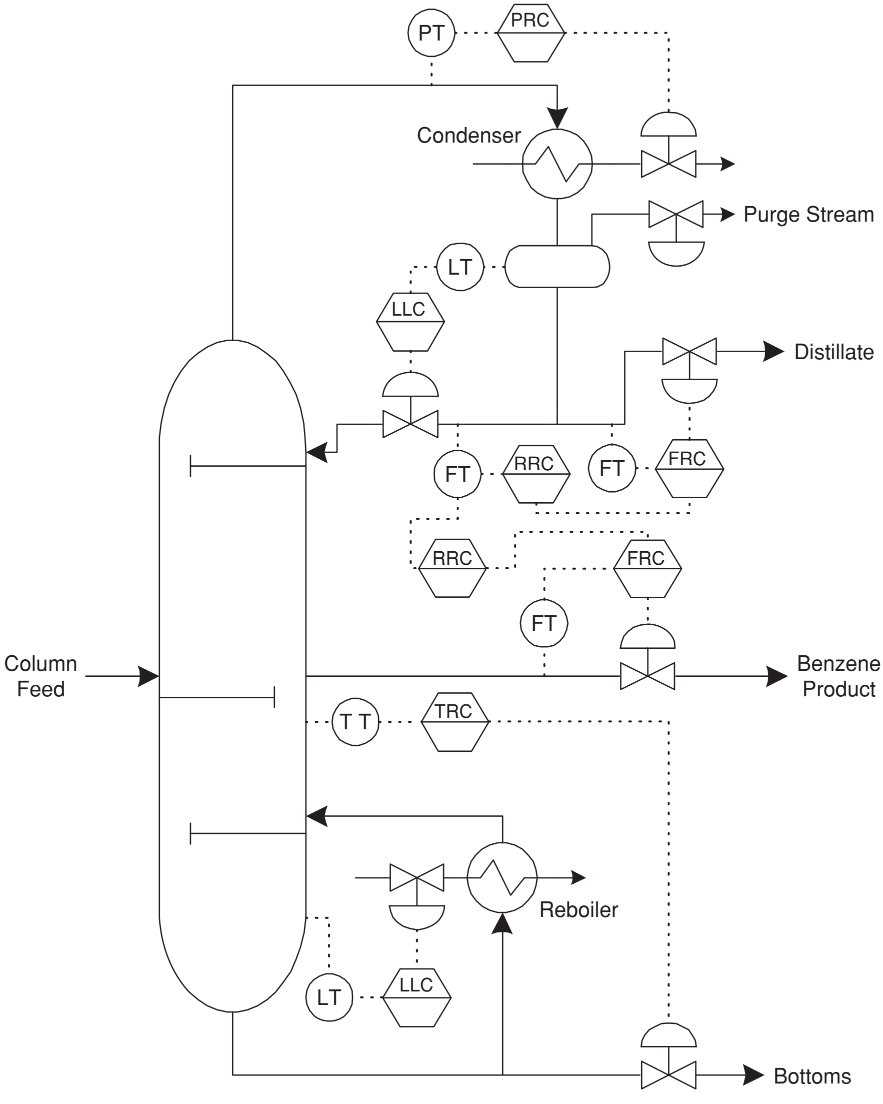

Figure 8.11 Liquid side draw benzene column control scheme

The introduction of more liquid feed means that more liquid benzene is cascading down the trays. As liquid reaches the bottom tray, the bubble point of the liquid on that tray decreases. The temperature controller reduces the valve opening on the bottoms stream to compensate. The reboiler level rises, resulting in more steam being introduced by the action of the level controller. The benzene and pentane are vaporized and move back up the column. As the column adjusts to the increased feed flow rate, the temperature profile in the column rises. The bottoms flow rate is then increased and settles back down to its new steady-state value.

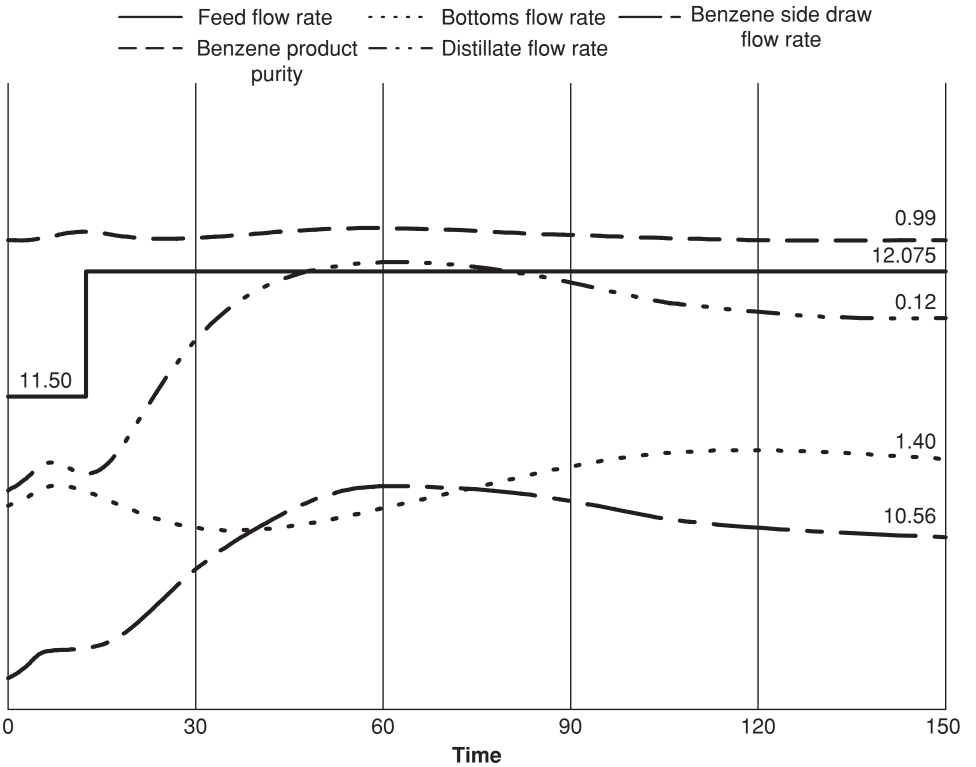

Figure 8.12 System response to a step change in feed flow rate

To test the control structure against changes in the composition of the feed stream, sinusoidal disturbances were introduced to the feed compositional flow rates. Each compositional flow rate was varied ± 10 per cent over periods ranging from 20 to 30 min. This example of a disturbance is a little unrealistic, but it demonstrates how the control structure would respond to compositional upsets. The strip chart of the same four flow rates and the benzene side draw purity is shown in Figure 8.13.

Each of the product flow rates responded in a similar fashion to that for a step change in the feed volumetric flow rate. As the feed flow rate increased, the products increased as well. The inverse response in the bottoms flow rate is not as observable now. The variable of interest is the benzene purity on the side draw. Although the feed composition is varying continuously, the variance in the benzene purity is much less. The affect on the benzene purity for drifting compositions is considerably damped.

8.7 Distillation control scheme design using steady-state models

Steady-state simulation of distillation columns has become routine. The use of these simulations has been restricted to use for heat and material balance and sizing purposes.

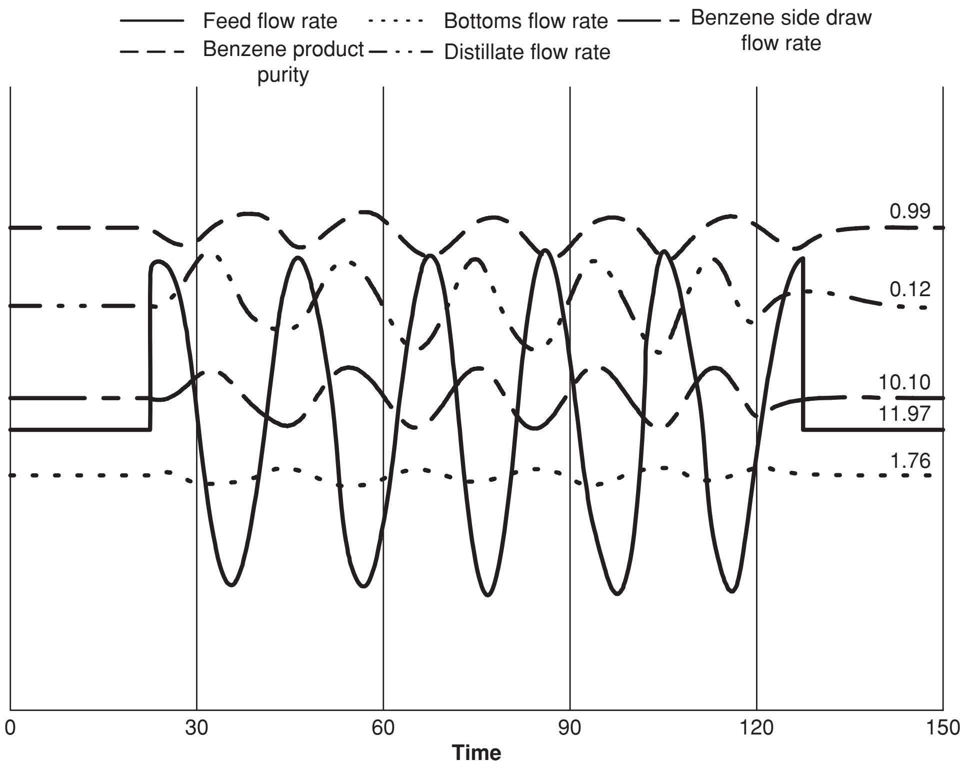

Figure 8.13 System response to sinusoidal disturbances in the component flow rates

Freuhauf and Mahoney [11,13,14] have shown that steady-state calculations [15] can be used to screen candidate control schemes, to provide a means for tray temperature location, and to calculate static gains.

Steady-state models are easily manipulated and are robust. This allows for the efficient generation of a large number of case studies necessary for steady-state design procedures. The obvious disadvantage of this procedure is that nothing is known about the dynamic response, and hence the dynamic disturbance rejection capability of alternative control schemes is also not known. These need to be evaluated using a dynamic simulator.

The basic steady-state design procedure consists of the following five steps:

1 *Develop a design basis.* Here, there is a need to define product composition specifications, disturbance type and size, constraints, and original column design basis.

2 *Select a candidate control scheme.* The literature abounds with alternative control configurations. Consider as an example a typical column that has feed as the disturbance stream. As was pointed out in the previous section, for such a column, only two degrees of freedom remain, i.e. feed split and fractionation. The resulting best feed-split control schemes are shown in Figures 8.14 and 8.15.

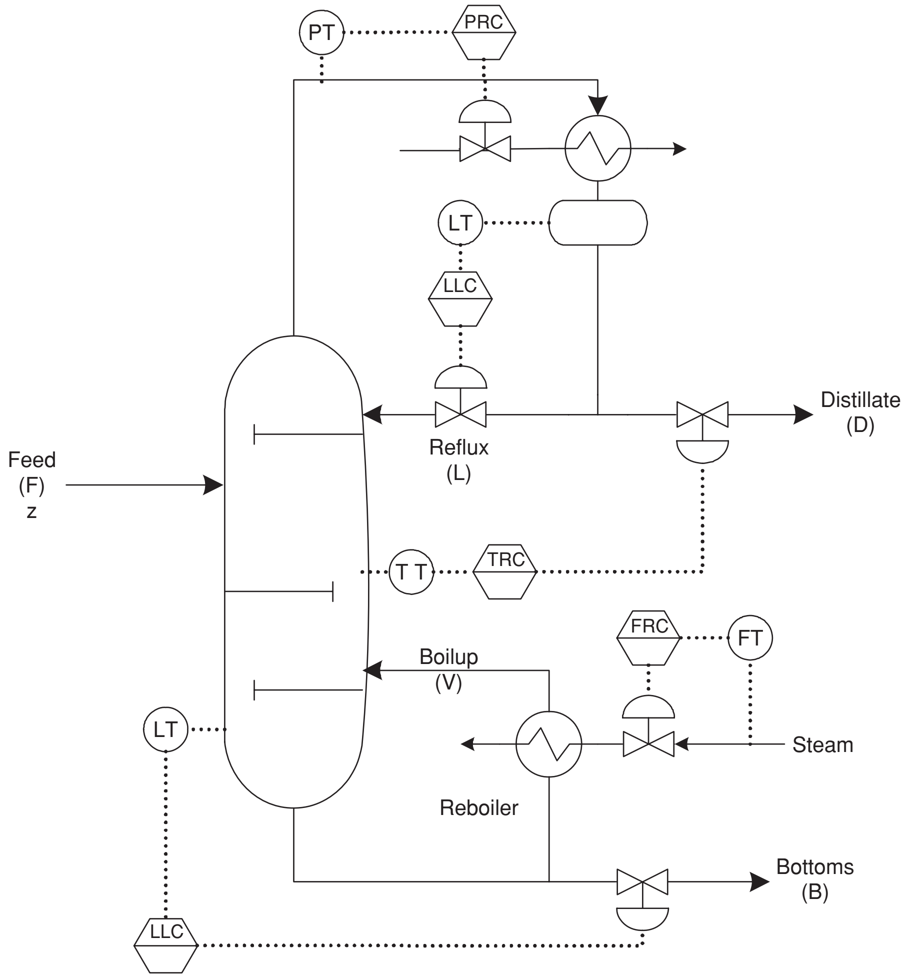

Figure 8.14 Direct feed-split control scheme

In Figure 8.14 we have a direct feed-split control scheme, because the distillate is manipulated directly to control composition. Product compositions are controlled by fixing a column temperature. The temperature controller manipulates the distillate flow. This control scheme is often selected when the heat input is limited or must be fixed.

Figure 8.15 illustrates the second common choice, indirect feed split control, where the distillate flow is increased indirectly by increasing the steam flow. The compositions are controlled by a temperature controller that manipulates the steam flow. This alternative has two advantages. One is that the temperature loop has faster closed-loop response (i.e. shorter natural period) and, therefore, provides better disturbance rejection. The second is that, because the reflux drum level sets the distillate flow, the reflux drum can be used to smooth flow disturbances to other

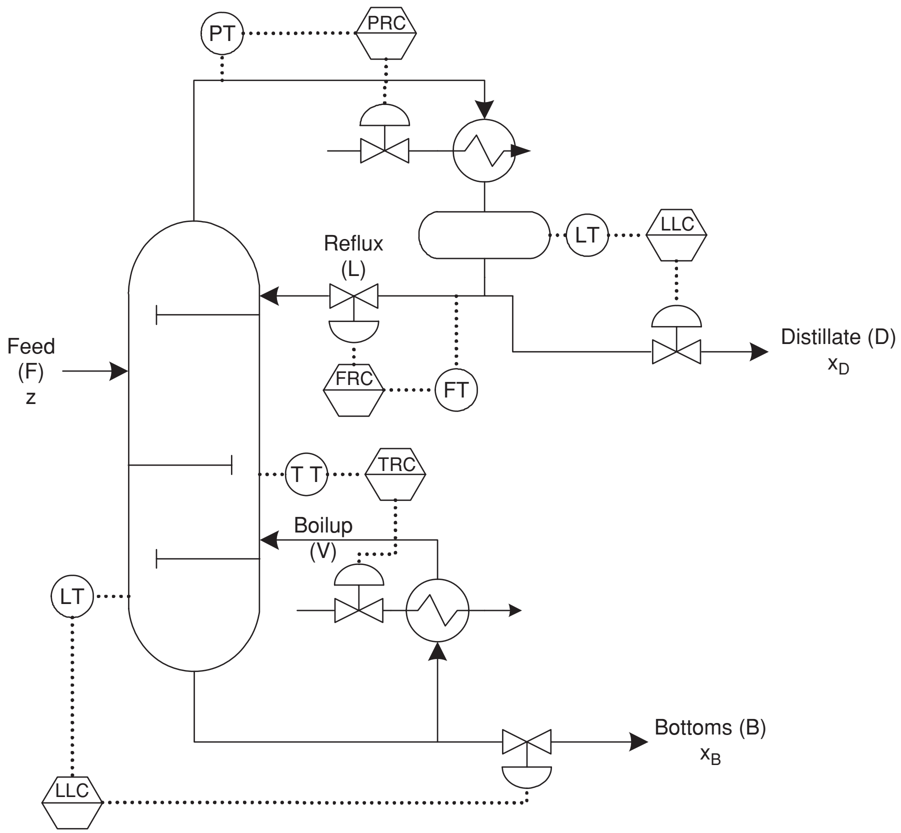

Figure 8.15 Indirect feed-split control scheme

downstream unit operations. To achieve flow smoothing, the level controller must have averaging level controller tuning. (Refer to Chapter 7).

The last part of the control strategy selection process is to select a ratio control alternative that might use less energy than the primary alternative. One example of a ratio control alternative for the scheme illustrated in Figure 8.15 would be a controller that keeps a constant reflux to feed flow ratio. This scheme likely will consume less energy than the non-ratio alternative because, as the feed flow to the column decreases, the amount of reflux will decrease. Less reflux will require less heat input.

3 *Conduct open-loop testing.* The purpose of this step is to use the steady-state model to identify a suitable tray for the temperature sensor for composition control. The procedure consists of using a candidate control scheme such as in Figure 8.14, in this case varying the distillate flow and observing the change in column temperature profile. A good tray is one on which the temperature change is significant and nearly equal when the flow is increased or decreased.

4 *Run closed-loop testing.* In this step, the steady-state model is used to simulate the candidate control scheme and to test its robustness to feed flow and feed composition changes. This step consists of a series of runs (sensitivity studies) aimed at locating a set of operating conditions that meet or exceed the product specifications for all expected disturbances in feed flow and composition.

5 *Confirm the objectives have been met.* If the objectives have been met, then the procedure is complete. If not, then the procedure is repeated with another candidate control scheme.

Case studies that demonstrate the application of this design scheme is available in the literature [13,14,16].

Screening control strategies via steady-state simulation

Steady-state simulation can often be used to evaluate options for base level process control strategies early in the design. The advantages of such an approach are many:

1 It leverages the work done by the process designers by extending the use of the design steady-state simulations to control work.

2 The process control design can parallel the equipment design work and can indicate where the limits inherent in the process design will cause controllability issues. This information can be used to drive changes to the process design, both in expanding or reducing the capacity of specific pieces of equipment.

3 A detailed cost/benefit analysis of expensive sensors, like online analysers, is available early in the project.

For example, Shell Canada has used this approach in three recent grass roots engineering projects and has found it to be very effective [17]. The following steps summarize the approach:

1 Respecify the simulation specifications to reflect candidate base level control loop objectives.

2 Mimic the behaviour of any process analysers or laboratory analyses.

3 Develop an economic profitability function for the process that takes simulation output information and computes the profitability of steady-state operating points.

4 Run multiple steady-state simulations across wide ranges of controller set points to assess the profitability of the candidate base level designs.

5 Assess the profitability of the base level designs against disturbances (typically, but not limited to, feed rate and feed composition).

The approach yields a steady-state control strategy design that is optimized for the particular economics and disturbance structures used, explicitly trading off reduced engineering costs against a potential drop in operational flexibility.

A case study: the Workshop stabilizer

The approach is best demonstrated with an example. Here, we will examine potential control strategies for the stabilizer column described in Workshop Exercise 7. The stabilizer is designed to remove volatile components from potential gasoline blend stocks. The feed is usually a mixture of C3, C4, and C5. In this case, the feed contains 5 per cent propane, 40 per cent isobutane, 40 per cent *n*-butane, and 15 per cent isopentane. The total flow rate is 40 000 bbl day^{-1} at 720 kPa and 30°C.

The stabilizer contains 20 trays and a total condenser. Feed enters at tray 10. The normal column overhead pressure is 700 kPa and there is a 20 kPa pressure difference that is evenly distributed between the condenser and the reboiler. Each tray is 2.0 m in diameter with a 0.10 m weir, which is 1.6 m long.

Respecifying simulation specifications

Figure 8.16 shows the column specifications in the HYSYS simulation file as received from the process design engineer. Note that the designer has included 12 potential specifications for the column, of which only two can be active at any given time. Here, the component fraction C3 in the bottoms and the tray 20 temperature are currently active.

A review of the potential specifications is warranted at this point. The goal is to remove any specifications that do not map onto typical instrumentation measurements. For example, flow meters work in volume units, not in molar units. Because the specifications of boil-up, reflux, distillate, and bottoms product rate given in Figure 8.16 are all molar flow specifications, they should not be used as control objectives. The reflux ratio and component recovery specifications are also expressed in molar units and, therefore, cannot be used. In fact, the only specifications that can be used directly from the design simulation are:

1 The reboiler duty, mimicking a steam or heat medium flow controller.

2 The three temperatures (trays 20, 3 and 1), mimicking three temperature controllers.

3 The component fraction C3 in the bottoms mimics an online analyser controller.

At this point, the control engineer may choose to add additional specifications to mimic other control loops. In this example, adding a specification setting the volumetric

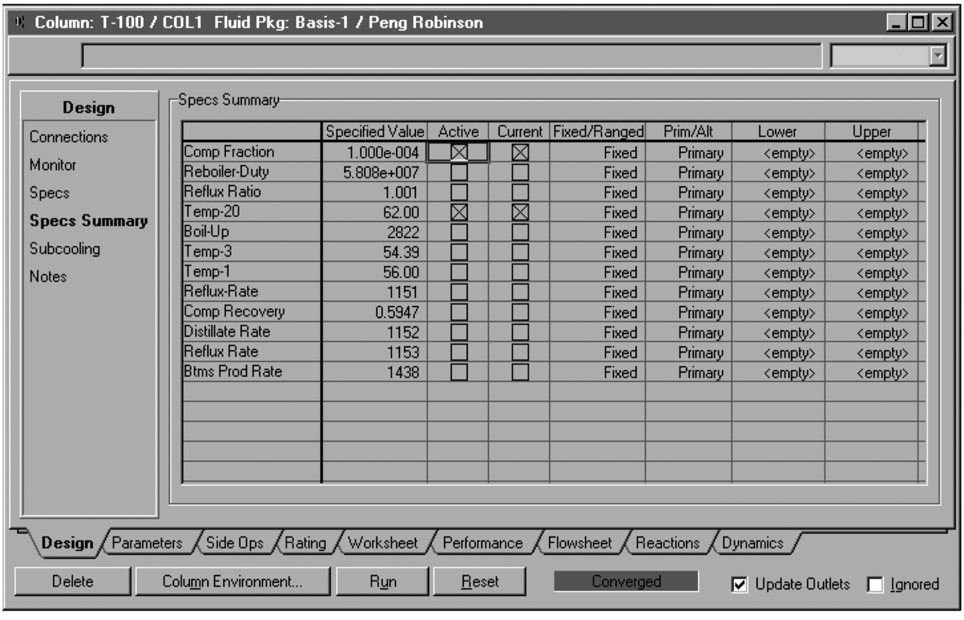

Figure 8.16 Typical column specifications from process design

flow rate of reflux (cubic metres per hour) would allow the control engineer to specify the reflux flow rate.

Figure 8.17 shows the column specification page with the unrealistic column specifications removed and the specifications reassigned to simulate a control strategy where the tray 20 temperature and the reflux volumetric flow are controlled. The simulation converges quickly with these two specifications active, indicating that the two control objectives are relatively decoupled.

Identifying conflicting specifications is relatively straightforward, as the simulation will take a long time to converge if conflicting specifications are used. Figure 8.18 shows that simultaneously specifying two tray temperatures results in convergence problems for the simulation. Therefore, this control structure should not be used.

The steady-state simulation is now prepared for control work. For simplicity, only two control strategies for the stabilizer will be investigated in this study:

1 *Active specifications: tray 20 temperature and reboiler duty*
With these two specifications, the distillate flow rate is being used to control the tray 20 temperature.

2 *Active specifications: tray 20 temperature and reflux volumetric flow*
With these two specifications, the reboiler duty is being used to control the tray 20 temperature.

Our goal is to determine which of two strategies is preferred.

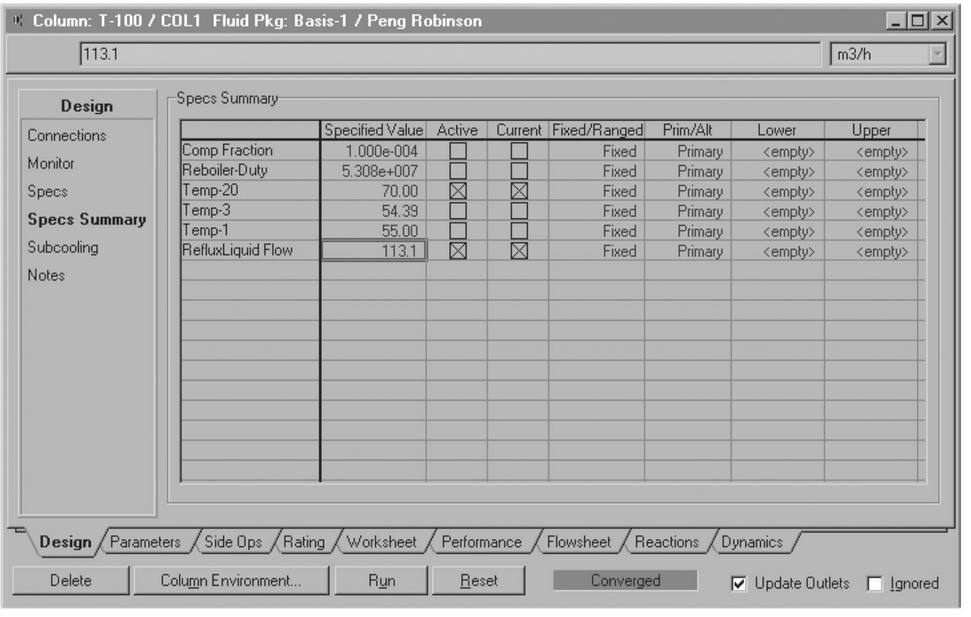

Figure 8.17 Column respecified to simulate one potential control strategy

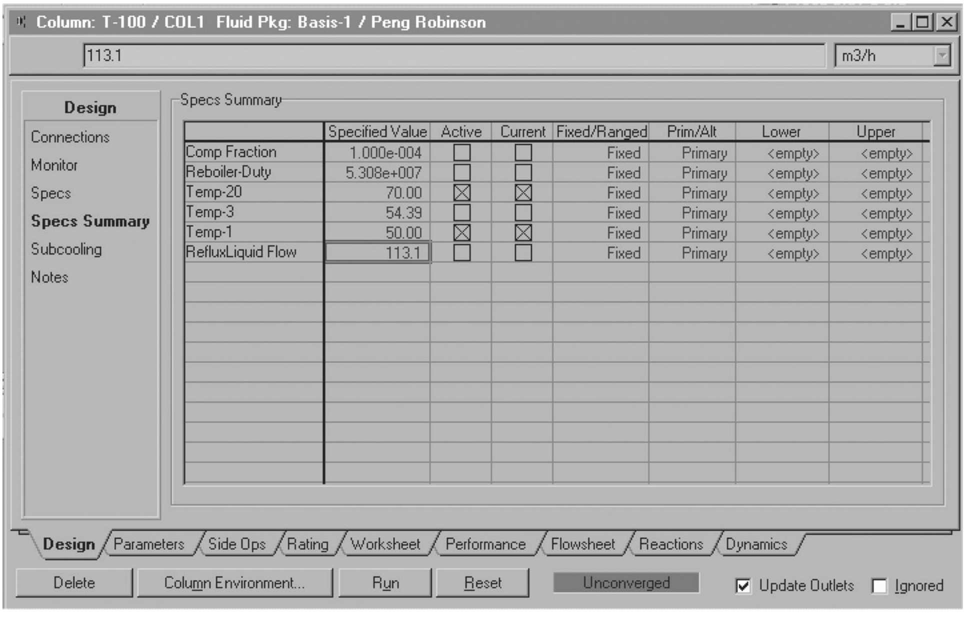

Figure 8.18 Example of conflicting specifications

Mimicking the behaviour of Analysers or laboratory analyses

The process designers used the component fraction C3 in the bottoms to design the stabilizer. Although it is possible to add a C3 analyser to the bottoms stream and control the C3 content to a desired value, it is more likely that the stabilizer will be run to a bottoms stream vapour pressure target, indicating the amount of light material in the bottoms stream. The Reid vapour pressure (RVP) or total vapour pressure (TVP) is a key blending property for gasoline and is typically measured via laboratory analysis.

For this work, the RVP/TVP add-in functionality from Hyprotech (RVP Methods Extension v2.0) was used to create a TVP measurement for the stabilizer bottoms stream.

Developing an economic profitability function

The most straightforward means of evaluating control strategies using the steady-state simulation approach is via an economic profitability function. Working in profit makes assessment of the trade-offs between various control objectives much easier. Typically, the profitability function is given as

$$\text{Profit} = \text{Value of Products} - \text{Value of Feeds} - \text{Operating Costs} \tag{8.7}$$

For this work, the main economic driver is maximizing the production of the bottom stream, which will be valued at regular gasoline prices. The value of the distillate product is significantly less than that of the bottoms given that its final destination is refinery fuel gas. The base value of the bottoms stream is 58.22 $\$\ m^{-3}$ at a TVP of 347.1 kPa. The base value of the distillate stream is 50.00 $\$\ m^{-3}$. Therefore, making more bottoms material at the same TVP will increase the profitability of the operation.

Because changing the volume of the bottoms product will affect its vapour pressure, the profitability function must include an adjustment for changing the bottoms product vapour pressure. The correction for changing vapour pressure is 0.25 $\$\ m^{-3}$/psi TVP. Therefore, if the TVP of the bottoms stream increases, then the profitability of the operation will drop, as this will reduce the amount of inexpensive light components that can be added to the gasoline blend and keep it on the blend vapour pressure constraint.

The value of the feed stream is 53.00 $\$\ m^{-3}$.

Finally, the operating costs must be accounted for. For the stabilizer column, the main operating cost is the cost of steam to reboil the column which is given as $5.6 \times 10^{-6}\$$ kJh^{-1}.

Computing the change in profitability from the base case operation is the most straightforward way of using the profitability function.

Evaluating the candidate strategies

The first step is to screen the profitability of both control strategies across a range of controller set points. Because both candidate strategies include the tray 20 temperature as a specification, it makes sense to screen it first.

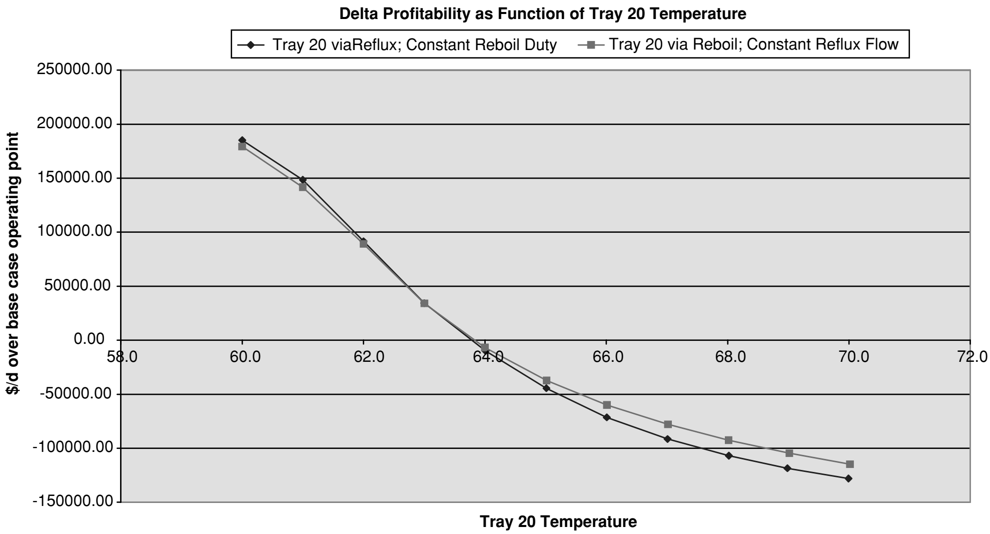

Figure 8.19 Delta profitability curves for candidate control strategies

Figure 8.19 shows several interesting results:

1 The base case design point is not the most profitable operation point. Under both control strategies, more profit can be made if the tray 20 temperature is reduced from its base case value of 63.8°C. A lower tray 20 temperature produces more bottoms stream volume, but at a higher TVP. But because the profitability of the increased volume is higher than the cost of the increased TVP of the bottoms stream, a lower tray 20 temperature set point is preferred. At some point, however, it is likely that a constraint will be hit on how much bottoms stream can be generated. For this work, it will be assumed that the maximum bottoms stream production that can be handled in blending is 200 m^3 h^{-1}.

2 The steam costs are insignificant in relation to the changes in bottoms product volume and composition.

3 There is a small difference in the profitability curves between the two control strategies. The constant reboiler duty strategy is slightly preferred at lower tray 20 temperatures, whereas the constant reflux flow strategy is preferred at high tray 20 temperatures. However, when the bottoms stream flow constraint of 200 m^3/h^{-1} becomes active, the constant reflux flow strategy is slightly preferred over the constant reboiler duty strategy, since the constant reboiler duty strategy must run at a slightly higher tray 20 temperature to make 200 m^3 h^{-1} of bottoms material (62.1°C versus 91.9°C). This shows the importance of determining all the important process constraints affecting the operation of the process under investigation.

Evaluating the candidate strategies under disturbances

While the behaviour of the candidate strategies appears similar to this point, it is important to examine their behaviour in relation to disturbances. To begin, the simulation was respecified to produce 200 m^3 h^{-1} of bottoms product, since this point was found to be the most profitable operating point. Then the main disturbance was introduced: a change in the feed composition to 5 per cent propane, 41 per cent isobutane, 40 per cent *n*-butane, and 14 per cent isopentane. Remember that the base case feed was 5 per cent propane, 40 per cent isobutane, 40 per cent *n*-butane, and 15 per cent isopentane. The profitability of the operation was then evaluated for both strategies.

In this case, there was a large difference in the profitability. The constant reboiler duty strategy produced 187.3 m^3h^{-1} of bottoms product (a drop of \$18 000 per day in profitability), whereas the constant reflux flow strategy only produced 183.3 m^3 h^{-1} for the same disturbance (a drop of \$26 500 per day in profitability). Therefore, the constant reboiler strategy is preferred.

Typically, a wide range of disturbances would be simulated and the control performance evaluated over this range. When constructing these additional case studies, the control engineer should be aware that specific variables used as disturbances might in fact be correlated with each other. For example, the process feed rate and composition to a reactor effluent distillation process may in reality be correlated with each other because the feed rate affects the reaction kinetics via a space–velocity relationship. Whenever possible, using real process data to determine the disturbance cases is preferred.

Therefore, our screening methodology indicates that, for our specific economics and specific disturbances, the constant reboiler duty strategy is preferred, since its profitability is less sensitive to disturbances.

Evaluating sensor strategies

In both strategies evaluated so far, the tray 20 temperature has been standing in for the TVP analysis. Another question could be asked: what is the economic driver to do the TVP analysis online and have the control strategy control it directly? In this case, we take each disturbance case and redo the simulation specifications to control this new measurement. Then the difference in profitability between the strategies without and with the new measurement would be computed.

In our example, the tray 20 temperature control objective would be replaced with the TVP objective. For the feed composition change, the profitability of the tray 20 temperature objective is \$126 729 per day and the profitability of the TVP objective is \$128 611 per day. Therefore, this indicates that, for this disturbance, there is a positive economic driver to control the TVP directly of about \$1880 per day. By performing this same type of analysis around the range of potential disturbances and weighting the benefits as per the likelihood of the disturbance occurring, we can determine a benefit number for the analyser installation. Of course, this benefit must be balanced against the installation and maintenance cost of the TVP analyser.

Example summary

The stabilizer case study has demonstrated how the effects of different control strategies on the profitability of a given process can be generated directly from steady-state simulations. The methodology requires:

1 An accurate economic profitability function for the process.

2 An accurate description of the expected disturbances to the process that will push the process away from its optimal profitability.

It is these two parts that take the most time to develop when using the methodology; the actual simulation runs are only a small part of the work.

8.8 Distillation control scheme design using dynamic models

As detailed above, the steady-state methodology can be used to screen a large number of candidate control schemes quickly and efficiently. However, it is desirable then to evaluate the candidate control schemes using a dynamic simulator to check the dynamic disturbance rejection capabilities of the alternative control schemes. A case study that demonstrates the application of this design scheme is available in the literature [16].

The basic dynamic design procedure consists of the following five steps (which follow upon the steady-state procedure):

1 *Provide information on material hold up.* Material may be held up in the condensers, tray sections, reboilers and other equipment used in the process. These delays will affect a controller's ability to respond in a reasonable amount of time when attempting to smooth out process disturbances. Typical hold-up times for various equipment items are listed in Table 8.1.

2 *Add controls and instrumentation.* Depict cascade controls when necessary, add lags and dead time to process measurement if this is expected.

Table 8.1 Typical equipment hold-up times

Equipment	Hold-up time
Heat exchangers	30 s
Distillation column trays	15–30 s (larger for crude columns)
Distillation column reflux accumulators	5–10 min
Distillation column reboilers	15–20 min (large enough to hold up liquid from trays, i.e. if dumped)
Surge vessels	5–15 min

3 *Implement control structures.* Tune these controllers.

4 *Practice with dynamic simulation.* Test for stability at startup, shutdown, runtime, typical upsets, etc.

5 *Repeat steps 2–4 for each control strategy.* Compare ease of startup and shutdown, disturbance rejection, what would happen if there were a change in the rate of production, complexity, and interaction between the controllers [14].

A case study that demonstrates the application of this design scheme is available in the literature [16].

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9

Using steady-state methods in a multi-loop control scheme

Control scheme selection for a unit operation, a process or a total plant is straightforward provided that each controlled variable is only affected by one manipulated variable. However, interactions are often present between the various control loops in multi-loop control schemes. Interactions occur when a manipulated variable affects a controlled variable of another loop. For instance, in a distillation column a manipulated variable, such as reflux flow rate, may affect several different controlled variables, such as distillate flow rate, distillate composition and/or reboiler duty. Hence, selecting the best control scheme for pairing manipulated and controlled variables is not straightforward. This chapter explores different methods for designing multi-input–multi-output control schemes for processes using steady state methods such as relative gain array (RGA), the Niederlinski index (NI) and singular value decomposition (SVD).

9.1 Variable pairing

When multiple, single-loop control schemes interact, the closure of one loop can change the closed loop gain of one or all the other control loops in the scheme. The SISO control loops may become unstable or respond sluggishly to disturbances, since the overall loop gain has been altered (Equation 2.6). The interaction between two control loops, in block diagram form, is illustrated in Figure 9.1, and is described mathematically as follows:

$$y_1 = a_{11}m_1 + a_{12}m_2 \tag{9.1}$$

$$y_2 = a_{21}m_1 + a_{22}m_2 \tag{9.2}$$

where y_i is the controlled or output variable i, m_j is the manipulated or input variable j, and a_{ij} is the input–output relationship or transfer function between y_i and m_j.

Figure 9.1 shows how a change in m_1 will affect both y_1 and y_2. For a 2×2 interacting system such as this one, there are two possible control configurations. One could pair m_1

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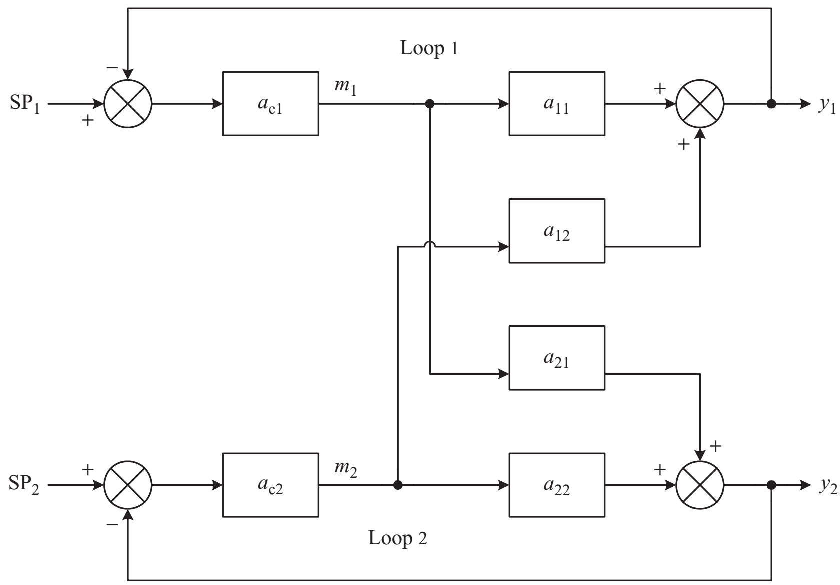

Figure 9.1 Loop interactions for a 2 × 2 system

with y_1 and m_2 with y_2, or m_1 could be paired with y_2 and m_2 with y_1. The best control scheme is the one that has minimal interaction between the two control loops and will remain stable in dynamic situations, rejecting load changes or random disturbances. For a control system containing n different controlled variables and n different manipulated variables, there are $n!$ control configurations possible. The relative gain array (RGA), first introduced by Bristol [1] in 1966, offers a quantitative approach to the analysis of the interactions present between the required control loops, thus providing a method of pairing manipulated and controlled variables.

9.2 The relative gain array

The RGA provides engineers with a quantitative comparison of how one control loop will affect others. Since it compares the affect of a manipulated variable on one controlled variable at a time, the problem is broken into more manageable segments. The disadvantage of the RGA method lies in the fact that it provides no information on the controller stability in dynamic situations, i.e. during process disturbances.

9.2.1 Calculating the relative gain array with experiments

Let us now consider m_1 as a candidate input to pair with y_1. To evaluate this choice against the alternative of using m_2, the system must undergo two experiments.

Experiment 1: step change in m_1 *with all loops open*

First, we will test the direct influence of m_1 on y_1. With no feedback control to affect m_1 (loop 1 is open), and m_2 held constant (loop 2 is open), the response of y_1 can be solely attributed to the change introduced in m_1. When a step change is made in m_1 with all the loops open, the output y_1 will change. As can be seen in Figure 9.1, y_2 will also change, but we will only be monitoring the response of y_1. The change in steady-state value of y_1 is equivalent to the steady-state gain between y_1 and m_1:

$$g_{11} = \left.\frac{\partial y_1}{\partial m_1}\right|_{m_2=\text{constant}} = \frac{\Delta y_1}{\Delta m_1} = \text{gain}(y_1 - m_1)\text{ loop with all loops open} \tag{9.3}$$

Experiment 2: step change in m_1 *with loop 1 open*

Now, loop 2 is closed and the same step change is made in m_1. 'Perfect control' is assumed in loop 2, meaning that m_2 will change in order to keep the controlled variable y_2 at a constant value. Mathematically:

$$g_{11}^* = \left.\frac{\partial y_1}{\partial m_1}\right|_{y_2=\text{constant}} = \frac{\Delta y_1}{\Delta m_1} = \text{gain}\,(y_1 - m_1)\text{ loop with all other loops closed} \tag{9.4}$$

Since Figure 9.1 shows m_1 interacting with y_2 via the a_{21} element, m_2 has to change to keep y_2 constant. Changing m_2 has an effect on y_1 via the a_{12} element, and it is this interaction that is being observed.

Using the results of experiments 1 and 2

The relative gain is a ratio or comparison of the gain of the $(m_j - y_i)$ loop with all loops open to the gain of the $(m_j - y_i)$ loop with all other loops closed and in 'perfect control' (no offset of other loop controlled variables).

$$\lambda_{ij} = \frac{\left(\dfrac{\partial y_i}{\partial m_j}\right)_{\text{all loops open}}}{\left(\dfrac{\partial y_i}{\partial m_j}\right)_{\text{all loops closed and in 'perfect control' except the } m_j \text{ loop}}} = \frac{g_{ij}}{g_{ij}^*}$$

$$= \left(\frac{\text{open-loop gain}}{\text{closed-loop gain}}\right)_{\text{for loop } i \text{ under control of } m_j} \tag{9.5}$$

The RGA or the Bristol array takes into account the relative gains for all combinations of input–output pairs in a multi-loop SISO system so that

$$\text{RGA} = \Lambda = \begin{bmatrix} \lambda_{11} & \lambda_{12} & \cdots & \lambda_{1n} \\ \lambda_{21} & \lambda_{22} & \cdots & \lambda_{2n} \\ \vdots & \vdots & \vdots & \vdots \\ \lambda_{n1} & \lambda_{n2} & \cdots & \lambda_{nn} \end{bmatrix} \tag{9.6}$$

This experimental method can be repeated using m_2 to control y_1. Experiments 1 and 2 must be repeated, where m_2 undergoes the step change and the changes in y_1 are observed. For experiment 2, m_1 will have to change to hold y_2 at a constant value, maintaining perfect control while m_2 is subjected to its step change.

9.2.2 Calculating the relative gain array using the steady-state gain matrix

It is possible to calculate the RGA as previously described by performing experiments 1 and 2 on each control pairing possibility. However, this may not be feasible for an operating plant. An alternative method for calculation of the RGA is feasible if a process model is available. The process model can be used to calculate the steady-state ($n \times n$) gain matrix, which can then be used to calculate the RGA.

The steady-state gain matrix can be calculated if one assumes that the steady-state condition is linear around each of the manipulated variables. It is a calculation that shows how each of the manipulated variables contributes to the overall effect on the controlled variables at steady state. It is also referred to as the 'open-loop gain matrix'.

Process model is available

The steady-state gain matrix G of the process can be derived based on the process model. The steady-state gain matrix is defined as follows:

$$\boldsymbol{y} = G\boldsymbol{m} \tag{9.7}$$

where $\boldsymbol{y}$ is the vector of the output or controlled variables, $\boldsymbol{m}$ is the vector of the input or manipulated variables, and G is the steady-state gain or open-loop matrix of the process.

Therefore, if $n = 3$, the steady-state gain matrix would be derived by first holding m_2 and m_3 constant while taking the partial derivative of y_1 with respect to m_1 to calculate g_{11}, the partial derivative of y_2 with m_1 to calculate g_{21}, and the partial derivative of y_3 with m_1 to calculate g_{31}; then for m_2 one would hold m_1 and m_3 constant while taking the partial derivative of y_1 with respect to m_2 resulting in g_{21}, and so on.

Let the steady-state gain matrix of the process be defined as follows:

$$G = \begin{bmatrix} g_{11} & g_{12} & \cdots & g_{1n} \\ g_{21} & g_{22} & \cdots & g_{2n} \\ \vdots & & & \\ g_{n1} & g_{n2} & \cdots & g_{nn} \end{bmatrix} \tag{9.8}$$

where g_{ij} has been defined in Equation 9.3 as

$$g_{ij} = \frac{\partial y_i}{\partial m_j} = \text{steady-state gain of } (y_i - m_j) \text{ with all loops open}$$

Now, let R be the transpose of the inverse of the steady-state gain matrix G, i.e.

$$R = \left(G^{-1}\right)^{\mathrm{T}} \tag{9.9}$$

The elements of the RGA can be obtained as follows:

$$\lambda_{ij} = g_{ij} r_{ij} \tag{9.10}$$

It is important to note that Equation 9.10 indicates an element by element multiplication of the corresponding elements of the two matrices G and R. This type of multiplication is called the Hadamard product of two matrices [2], and is *not* the normal matrix product.

Process model is not available

If there are no process model equations to be differentiated to obtain g_{ij}, then one can use experimental results to calculate the steady-state gain matrix. Refer to experiment 1 of the RGA calculation (Section 9.2.1), where, as in Equation 9.3:

$$g_{ij} = \frac{\Delta y_i}{\Delta m_j} = \text{steady-state gain of } (y_i - m_j) \text{ with all loops open}$$

Once the steady-state gain matrix is known it can then be manipulated to generate the RGA as described previously.

9.2.3 Interpreting the relative gain array

The RGA is a useful tool if its properties and limitations are recognised. To understand the significance of the RGA, the following points should be understood [3]:

1 The elements of the RGA across any row, or down any column, sum up to unity:

$$\sum_{i=1}^{n} \lambda_{ij} = \sum_{j=1}^{n} \lambda_{ij} = 1 \tag{9.11}$$

So, in the case of a 2×2 system, only one relative gain element needs to be calculated to determine the values of the others.

2 λ_{ij} is dimensionless and unaffected by scaling.

3 If $\lambda_{ij} = 0$, then the manipulated variable m_j has no effect on the output or controlled variable y_i.
Pairing implication: do not pair m_j with y_i.

4 If $\lambda_{ij} = 1$, then this implies that m_j affects y_i without interaction from the other control loops. The gain when all the loops are open is equal to the gain when all the other loops except $(m_j - y_i)$ are closed.
Pairing implication: pair m_j with y_i.

5 If $\lambda_{ij} < 0$, then the system will potentially be unstable when m_j is paired with y_i, or the system may initially respond opposite to what is really happening (i.e. the shrink–swell effect in boilers described in Section 7.8.2). Note, that the interaction from the other control loops is more dominant than the pairing interaction. This may result in the system becoming unstable if the other loops are opened, since the open loop response is opposite in direction to the closed loop response.
Pairing implication: avoid pairing m_j with y_i.

6 If $0 < \lambda_{ij} < 1$, then the other control loops are interacting with the $(m_j - y_i)$ control loop. If $\lambda_{ij} = 0.5$, then the control pair effect is equal to the retaliatory effect of the other loops. $\lambda_{ij} < 0.5$ indicates that the other control loops have a greater influence on the control pair than it has on itself. $\lambda_{ij} > 0.5$ indicates that the control pair $(m_j - y_i)$ has a greater effect than the other loops.
Pairing implication: avoid pairing m_j with y_i when $\lambda_{ij} \leq 0.5$.

7 If $\lambda_{ij} > 1$, then the open-loop gain of the pair $(m_j - y_i)$ is greater than the gain when all other loops are closed. This indicates that the other loops are influencing the pair in the opposite direction. However, the relative gain is still greater than zero, and so the pair $(m_j - y_i)$ is dominant. Note, that a higher value of λ_{ij} indicates more retaliatory effects from other control loops, and may result in the control pair becoming unstable when the other loops are opened.
Pairing implication: avoid pairing m_j with y_i when $\lambda_{ij} \gg 1$.

9.3 Niederlinski index (NI)

The NI [4] is a useful tool to analyse the stability of the control loop pairings determined using the relative gain analysis. If a manipulated variable is to be used to control an output variable, then the loop must not become unstable in dynamic situations. The NI can be used to prove that a 2×2 matrix is stable; however, when $n > 2$ (if there are more than two input–output variables being paired) the NI can only be used to prove that the control loop is definitely not stable. Then, for the steady-state matrix G described in Equation 9.7, where each element in $\hat{G}$ is rational and open-loop stable [3], the system will definitely be unstable if the NI is negative, i.e. if

$$\mathrm{NI} = \left. \frac{|G|}{\prod_{i=1}^{n} g_{ii}} \right|_{\mathrm{SS}} \tag{9.12}$$

is negative.

The NI will detect instability introduced by closing the other control loops. Remember that the NI does not prove the control system is stable when there are $n > 2$ variables; a negative NI only proves that the system is definitely not stable.

The NI should not be used for systems with time delays (dead time). Grosdidier *et. al.* [5] provides a detailed explanation on how to use the index for systems containing dead time. Dynamic simulation should always be used to test the stability of a system if the NI is positive.

Relative gain array pairing rules

There are some basic rules that should be followed to obtain optimal pairing in control loops:

RGA rule 1. Pair the input and output variables that have positive RGA elements that are closest to 1.0.

RGA Rule 2. Any loop pairing is unacceptable if it leads to a control system configuration for which the NI is negative.

9.4 Decoupling control loops

In some process situations, one manipulated variable may dominate more than one controlled variable's response. This situation is best avoided, since it is almost impossible to control such interactions. There are practical ways of dealing with such significant controller interactions, and these include restructuring the pairing of variables, detuning the offending control loops to minimise interactions, opening some loops (manual control) and using linear combinations of manipulated and/or controlled variables [2].

Singular Value Decomposition (SVD) [2,3] is a useful tool to determine whether a system will be prone to control-loop interactions resulting in sensitivity problems. These sensitivity problems typically result from small errors in process gains. This section will present and demonstrate the use of SVD.

9.4.1 Singular value decomposition

Ideally, manipulated variables are coupled to controlled variables on a one-to-one basis, i.e. m_1 controls y_1, m_2 controls y_2, etc., for ease of control. Since interactions do occur often between control loops, it is these controller interactions that need to be decoupled. SVD is a matrix technique useful in determining whether it is structurally impossible to apply decoupling to a system [3]. When the sets of equations in the steady-state gain matrices are nearly singular, the problem is ill conditioned and it may not be possible to decouple control-loop interactions.

SVD can be applied to the steady-state gain matrix. The gain matrix is first decomposed into the product of three matrices, where two are eigenvectors and one is an eigenvalue [2,3,6] matrix:

$$G = U \Sigma V^{\mathrm{T}} \tag{9.13}$$

where U is the matrix of normalised eigenvectors of GG^{T}, V is the matrix of normalised eigenvectors of $G^{\mathrm{T}}G$, and Σ is a diagonal matrix of eigenvalues.

For systems where $n = 2$, analytical expressions have been developed to calculate the three matrices [2]. The matrix of most interest is the eigenvalue matrix Σ.
For the gain matrix:

$$G = \begin{bmatrix} g_{11} & g_{12} \\ g_{21} & g_{22} \end{bmatrix} \tag{9.14}$$

The following values can be defined [2]:

$$b = g_{11}^2 + g_{12}^2 \tag{9.15}$$

$$c = g_{11}g_{21} + g_{12}g_{22} \tag{9.16}$$

$$d = g_{21}^2 + g_{22}^2 \tag{9.17}$$

Then

$$\lambda_1 = s_1^2 = \frac{(b+d) + \sqrt{(b-d)^2 + 4c^2}}{2} \tag{9.18}$$

$$\lambda_2 = s_2^2 = \frac{bd - c^2}{s_1^2} \tag{9.19}$$

and

$$\Sigma = \begin{bmatrix} s_1 & 0 \\ 0 & s_2 \end{bmatrix} \tag{9.20}$$

The values of s_1 and s_2 are always positive and the ratio of the larger s_i to the smaller s_i is called the condition number CN [2]:

$$\mathrm{CN} = \frac{s_1}{s_2} \tag{9.21}$$

For example, if the CN number were equal to 100, then this would indicate that one manipulated variable has 100 times more effect on the system than the other manipulated variable. The higher the CN number, the more difficult it becomes to decouple a control loop interaction. A rule of thumb is that when $CN \geq 50$ the system is nearly singular, and decoupling is not feasible [2].

9.5 Tuning the controllers for multi-loop systems

Since a manipulated variable generally affects more than one controlled variable in a multi-loop system, it may be challenging to tune the system properly. The easiest way to work with a multi-loop system is to treat it as a group of individual control loops. First tune each loop with all the other control loops in manual. Then, close all the control loops and retune the control loops until the system can 'handle' a known disturbance without losing its stability. It is often necessary to loosen the original tuning parameters to minimise interactions between control loops. This entails decreasing the controller gains and increasing the integral times [3]. Dynamic simulation can be used to reduce the time required drastically and to simplify the above controller tuning procedure.

9.6 Practical examples

The techniques presented will now be illustrated in the following two examples, namely temperature control of a mixer outlet and control scheme configuration for a distillation column. The RGA will be used to select which controlled and manipulated variables will be paired, the NI will then be used to demonstrate whether or not the resulting control loops are stable, and SVD will be used to test whether the control loop interactions are overly sensitive to slight errors in process gains.

Example 1: a two stream mixer

Consider a mixer where the hot and cold streams are being used to control the temperature and flow rate of the outlet stream (Figure 9.2). The hot stream has a constant temperature of 50°C and the cold stream has a constant temperature of 5°C. At steady state, the final desired temperature is 35°C, and the final flow rate is 600 kg h^{-1}.

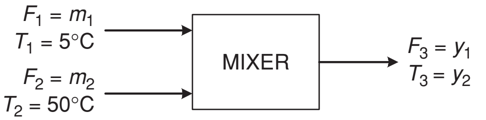

Figure 9.2 Mixer control

The equations, or process model to describe this system, are:

$$y_1 = m_1 + m_2 \tag{9.22}$$

$$y_2 = \frac{T_1 m_1 + T_2 m_2}{y_1} \tag{9.23}$$

where T is in kelvin and the specific heat capacities of the streams are assumed constant.

At steady state the following values are maintained:

$$\begin{aligned} T_1 &= 5^\circ\mathrm{C} = 278.15\,\mathrm{K} \\ m_1 &= 200\,\mathrm{kg\,h^{-1}} \\ T_2 &= 50^\circ\mathrm{C} = 323.15\,\mathrm{K} \\ m_2 &= 400\,\mathrm{kg\,h^{-1}} \end{aligned}$$

resulting in

$$\begin{aligned} y_1 &= 600\,\mathrm{kg\,h^{-1}} \\ y_2 &= 35^\circ\mathrm{C} = 308.15\,\mathrm{K} \end{aligned}$$

Steady-state gain matrix calculation

To calculate the steady-state gain matrix, open-loop gains can be found by differentiating the model with respect to m_i while holding m_j $(j \neq i)$ constant.

$$g_{11} = \left.\frac{\partial y_1}{\partial m_1}\right|_{m_2 = \text{constant}} = \frac{\partial}{\partial m_1}(m_1 + m_2) = 1 \tag{9.24}$$

$$g_{21} = \left.\frac{\partial y_2}{\partial m_1}\right|_{m_2 = \text{constant}} = \frac{\partial}{\partial m_1}\left(\frac{T_1 m_1 + T_2 m_2}{m_1 + m_2}\right) = \left[\frac{(T_1 - T_2)m_2}{(m_1 + m_2)^2}\right] \tag{9.25}$$

$$g_{12} = \left.\frac{\partial y_1}{\partial m_2}\right|_{m_1 = \text{constant}} = \frac{\partial}{\partial m_2}(m_1 + m_2) = 1 \tag{9.26}$$

$$g_{22} = \left.\frac{\partial y_2}{\partial m_2}\right|_{m_1 = \text{constant}} = \left[\frac{(T_2 - T_1)m_1}{(m_1 + m_2)^2}\right] \tag{9.27}$$

Equations 9.24–9.27 are open-loop gains, and can be evaluated by experiment 1 in Section 9.2.1 when a mathematical model is not available.

Equations 9.24–9.27 are now solved using the steady-state values listed previously. For example:

$$g_{22} = \frac{(323.15 - 278.15) \times 200}{(200 + 400)^2} = 0.025 \tag{9.28}$$

The steady-state gain matrix is

$$\begin{bmatrix} y_1 \\ y_2 \end{bmatrix} = \begin{bmatrix} 1 & 1 \\ -0.05 & 0.025 \end{bmatrix} \begin{bmatrix} m_1 \\ m_2 \end{bmatrix} \tag{9.29}$$

which is of the form

$$\boldsymbol{y} = G\boldsymbol{m}$$

as described in Equation 9.7.

Relative gain array calculation

To calculate the RGA, first the inverse of the steady-state matrix must be found:

$$G^{-1} = \begin{bmatrix} \frac{1}{3} & -13\frac{1}{3} \\ \frac{2}{3} & 13\frac{1}{3} \end{bmatrix} \tag{9.30}$$

The transpose of G^{-1}:

$$R = (G^{-1})^{\mathrm{T}} = \begin{bmatrix} \frac{1}{3} & \frac{2}{3} \\ -13\frac{1}{3} & 13\frac{1}{3} \end{bmatrix} \tag{9.31}$$

Now, the Hadamard product of the two matrices must be calculated where $\lambda_{ij} = g_{ij}r_{ij}$. For example:

$$\lambda_{21} = -0.05 \times -13\tfrac{1}{3} = \tfrac{2}{3} \tag{9.32}$$

So, the resulting RGA is

$$\Lambda = \begin{bmatrix} \frac{1}{3} & \frac{2}{3} \\ \frac{2}{3} & \frac{1}{3} \end{bmatrix} \tag{9.33}$$

The RGA for this example could also be calculated by taking the ratio of the open-loop gain to the closed-loop gain (using the mathematical process models or experiments 1 and 2 from Section 9.2.1). For example:

$$\left.\frac{\partial y_1/\partial m_1\Big|^{m_2\,=\,\text{constant}}}{\partial y_1/\partial m_1\Big|_{y_2\,=\,\text{constant}}} = \frac{1}{1+\partial m_2/\partial m_1|_{y_2=\text{constant}}} \right. \tag{9.34}$$

$\partial m_2/\partial m_1\big|_{y_2=\text{constant}}$ will now be evaluated by substituting Equation 9.22 into Equation 9.23 and solving for m_1:

$$m_1 = \frac{T_2 - y_2}{y_2 - T_1} m_2 \tag{9.35}$$

Substituting Equation 9.35 into Equation 9.22:

$$y_1 = \frac{T_2 - y_2}{y_2 - T_1} m_2 + m_2 \tag{9.36}$$

Now this equation must be differentiated at constant y_2 and set equal to the denominator in Equation 9.34.

$$\left.\frac{\partial y_1}{\partial m_1}\right|_{y_2=\text{constant}} = \frac{T_2 - y_2}{y_2 - T_1}\frac{\partial m_2}{\partial m_1} + \frac{\partial m_2}{\partial m_1} = 1 + \frac{\partial m_2}{\partial m_1} \tag{9.37}$$

Evaluating to solve for $\partial m_2/\partial m_1$ at steady state:

$$\frac{\partial m_2}{\partial m_1} = \frac{y_2 - T_1}{T_2 - y_2} = \frac{308.15 - 278.15}{323.15 - 308.15} = 2 \tag{9.38}$$

Substituting Equation 9.38 back into Equation 9.34 to solve for λ_{11} results again in a value of 1/3. This result, of course, matches that using the 'steady-state gain matrix', which only required that the open-loop gains be calculated. This RGA shown in Equation 9.33 indicates that the best control scheme is to use m_2 to control y_1 and m_1 to control y_2. In Section 9.2.3, point 6 states that one should avoid pairing m_j with y_i when $\lambda_{ij} <= 0.5$.

The steady state gain matrix for pairing $y_1 - m_2$ and $y_2 - m_1$ is:

$$\begin{bmatrix} y_1 \\ y_2 \end{bmatrix} = \begin{bmatrix} 1 & 1 \\ 0.025 & -0.05 \end{bmatrix} \begin{bmatrix} m_2 \\ m_1 \end{bmatrix}$$

The resulting RGA for this $y_1 - m_2$ and $y_2 - m_1$ pairing is now

$$\Lambda = \begin{bmatrix} \frac{2}{3} & \frac{1}{3} \\ \frac{1}{3} & \frac{2}{3} \end{bmatrix}$$

The rest of Example 1 will be evaluated using m_2 to control y_1 and $m_1$to control y_2.

Niederlinski index

Calculating the NI to determine whether or not the mixer control scheme will produce a stable system results in a value of 1.5 as follows:

$$\mathrm{NI} = \left.\frac{|G|}{\prod_{i=1}^{n} g_{ii}}\right|_{\mathrm{ss}} = \frac{(1 \times -0.05) - (1 \times 0.025)}{(1 \times -0.05)} = 1.5$$

Hence, this pairing will result in a stable system, since NI > 0 and $n = 2$.

Singular value decomposition

From the RGA, it can be seen that interaction does exist between the two loops, since $0 < \lambda < 1$. Whether or not this interaction will cause sensitivity problems may be determined from SVD. The SVD is calculated from the 'G' as follows:

$$G = \begin{bmatrix} 1 & 1 \\ 0.025 & -0.05 \end{bmatrix}$$

where

$$\begin{aligned} b &= g_{11}^2 + g_{12}^2 = 1^2 + 1^2 = 2 \\ c &= g_{11}g_{21} + g_{12}g_{22} = (1 \times 0.025) + (1 \times -0.05) = -0.025 \\ d &= g_{21}^2 + g_{22}^2 = (0.025)^2 + (-0.05)^2 = 0.003\,125 \end{aligned}$$

resulting in:

$$\begin{aligned} \lambda_1 = \mathrm{s}_1^2 &= \frac{(b+d) + \sqrt{(b-d)^2 + 4c^2}}{2} \\ &= \frac{(2 + 0.003\,125) + \sqrt{(2 - 0.003\,125)^2 + 4(-0.025)^2}}{2} = 2 \\ \lambda_2 = \mathrm{s}_2^2 &= \frac{bd - c^2}{s_1^2} = \frac{2 \times 0.003\,125 - (-0.025)^2}{2} = 0.002\,81 \end{aligned}$$

and

$$\Sigma = \begin{bmatrix} \mathrm{s}_1 & 0 \\ 0 & \mathrm{s}_2 \end{bmatrix} = \begin{bmatrix} 1.414 & 0 \\ 0 & 0.053 \end{bmatrix}$$

$$\mathrm{CN} = \frac{\mathrm{s}_1}{\mathrm{s}_2} = \frac{1.414}{0.053} \approx 27$$

The condition number CN is less than 50, hence this system will not be prone to sensitivity problems [2].

9.6.1 Example 2: a conventional distillation column

In this example, the RGA analysis is used to find the appropriate pairing for a conventional distillation column. There are typically two control schemes for distillation columns: single and dual composition control. The single composition control scheme maintains the composition of one of the products at a desired value, whereas in dual composition control both products are regulated.

Once the column pressure is set (typically using coolant flow rate in the condenser), then the following variables can be used as manipulated variables:

1. reboiler duty Q_R
2. reflux flow L
3. distillate flow D
4. bottom product flow B.

The reasons why feed flow rate and reflux ratio are not considered as manipulated variables are as follows:

- The feed stream of the column is usually downstream of other units; or, restated, its characteristics are usually set based on the operating condition of upstream units.
- Using reflux ratio as one of the manipulated variables results in an upset to the column whenever the distillate flow rate changes.

The variables usually considered as the process outputs for a distillation column are liquid levels at the base of the column and reflux drum and product compositions in dual composition control. Since there are four inputs that can be used to control four outputs, there are 4! different combinations. These combinations are shown in Table 9.1.

A preliminary screening of these 24 alternatives based on the dynamic response of the manipulated variable to the measured variable results in three viable alternative pairings: cases 4, 10, and 18, shown in bold in Table 9.1. The reasons for discarding other pairings are as follows:

- Cases 1, 3, 5, 7, 9, 11, 13, 15, 19, 20, 23 and 24 are discarded since they involve control of base level by reflux flow or distillate flow.
- Cases 6, 8, 14 and 19 are discarded since they involve manipulating flow rate of bottom product or reboiler heat to control the liquid level in the reflux drum.

Table 9.1 Pairings in dual composition control

Case	Reflux drum	Column base	Top composition	Bottom composition
1	D	L	B	Q_R
2	D	Q_R	B	L
3	L	D	B	Q_R
4	$\boldsymbol{L}$	$\boldsymbol{B}$	$\boldsymbol{D}$	$\boldsymbol{Q_R}$
5	B	L	D	Q_R
6	B	Q_R	D	L
7	Q_R	D	B	L
8	Q_R	B	D	L
9	D	L	Q_R	B
10	$\boldsymbol{D}$	$\boldsymbol{Q_R}$	$\boldsymbol{L}$	$\boldsymbol{B}$
11	L	D	Q_R	B
12	L	B	Q_R	D
13	B	L	Q_R	D
14	B	Q_R	L	D
15	Q_R	D	L	B
16	Q_R	B	L	D
17	D	B	Q_R	L
18	$\boldsymbol{D}$	$\boldsymbol{B}$	$\boldsymbol{L}$	$\boldsymbol{Q_R}$
19	B	D	L	Q_R
20	B	D	Q_R	L
21	L	Q_R	D	B
22	L	Q_R	B	D
23	Q_R	L	B	D
24	Q_R	L	D	B

- Cases 21 and 22 are discarded since they do not regulate the material balance.
- Cases 2, 12 and 17 are discarded since each involves the control of one (or both) product composition(s) at the end of the column using a manipulated variable at the other end of the column.

Base case steady-state solution

The best pairing among these three alternatives, i.e. 4, 10, and 18, will be found through RGA analysis of a water–ethanol distillation column. A common approach is to use a process simulation software package to determine the necessary gains for the RGA analysis and NI. For this example we have used HYSYS. Process™ [7]. The condenser and reboiler levels will be assumed to be under perfect control. For the water–ethanol system the NRTL activity model with the ideal gas vapour model was selected. The column feed stream is shown in Table 9.2.

Table 9.2 Characteristics of the column feed

Conditions and composition	
Temperature (°C)	20.0
Pressure (kPa)	101.3
Molar flow rate of water (kmol h^{-1})	60.00
Molar flow rate of ethanol (kmol h^{-1})	40.00

The distillation column has 20 stages and a total condenser. A steady-state solution for the distillation column can be performed using the information in Tables 9.3 and 9.4.

The steady-state solutions for the column yield the results shown in Table 9.5.

Relative gain array calculation

For this exercise, the steady-state values for the compositions will be used as the desired set points for the controllers and the best control pairings must be determined for maintaining the distillate and bottoms product compositions. The RGA will be calculated for each pairing of the three feasible alternatives: cases 4, 10, and 18 from Table 9.1.

Pairing comparison for cases 4, 10, and 18 At this point, the steady-state gain between the distillate flow rate and the distillate composition, g_{11}, will be calculated using the steady-state distillation column model. Perform a step input in the distillate flow rate from 30 to 40 kmol h^{-1}. Change one of the column specifications from reflux ratio to reboiler duty, specifying a reboiler duty equal to the base case steady-state solution of 4.2×10^6 kJ h^{-1}. The new specifications for the column should be the same as those given in Table 9.6.

Run the column and determine the new mole fractions for ethanol in the distillate and bottoms. The results should be very close to those shown in Table 9.7.

The open-loop gain is then calculated as follows:

$$g_{11} = \frac{\Delta x_{\mathrm{D}}}{\Delta D} = \frac{0.7890 - 0.8165}{40 - 30} = -2.75 \times 10^{-3} \tag{9.39}$$

The closed-loop gain g_{11}^* can be calculated from the steady-state solution by closing the bottoms composition control loop, i.e. making the desired bottoms composition a steady-state specification. The closed-loop specifications are given in Table 9.8.

Run the column and determine the new mole fractions for ethanol in the distillate and bottoms. The results should be very close to those shown in Table 9.9.

Table 9.3 Distillation column data

Column characteristics		Column pressure	
No. of stages	20	Condenser pressure (kPa)	95
Feed stage	10	Condenser ΔP	0
Condenser type	Total	Reboiler pressure (kPa)	105

Table 9.4 Distillation column specifications for base case steady-state

Specification	Value
Reflux ratio	2.0
Distillate rate (kmol h^{-1})	30

Table 9.5 Base case steady-state solution

Mole fraction of ethanol in distillate	0.8165
Mole fraction of ethanol in bottoms	0.2214
Reboiler duty (kJ h^{-1})	4.2×10^6

Table 9.6 Specifications for case 4 open loop

Specification	Value
Reboiler duty (kJ h^{-1})	4.2×10^6
Distillate rate (kmol h^{-1})	30

Table 9.7 Steady-state solution for case 4 open loop

Mole fraction of ethanol in distillate	0.7890
Mole fraction of ethanol in bottoms	0.1407
Reboiler duty (kJ h^{-1})	4.2×10^6

Table 9.8 Specifications for case 4 closed loop

Specification	Value
Mole fraction of ethanol in bottoms	0.2214
Distillate rate (kmol h^{-1})	30

Table 9.9 Steady-state solution for case 4 closed loop

Mole fraction of ethanol in distillate	0.6680
Mole fraction of ethanol in bottoms	0.2214
Reboiler duty (kJ h^{-1})	2.6×10^6

Now, the closed-loop gain is calculated as follows:

$$g_{11}^* = \frac{\Delta x_D}{\Delta D} = \frac{0.6680 - 0.8165}{40 - 30} = -1.48 \times 10^{-2} \tag{9.40}$$

At this point, the RGA matrix can be calculated because this is a 2×2 system.

$$\lambda_{11} = \frac{g_{11}}{g_{11}^*} = \frac{-2.75 \times 10^{-3}}{-1.48 \times 10^{-2}} = 0.185 \tag{9.41}$$

The RGA matrix is then

$$\begin{bmatrix} \lambda_{11} & 1 - \lambda_{11} \\ 1 - \lambda_{11} & \lambda_{11} \end{bmatrix} = \begin{bmatrix} 0.185 & 0.815 \\ 0.815 & 0.185 \end{bmatrix} \tag{9.42}$$

The step changes to calculate λ_{11} for cases 10 and 18 are shown in Table 9.10. The resulting RGA matrix for case 10 is

$$\begin{bmatrix} \lambda_{11} & 1 - \lambda_{11} \\ 1 - \lambda_{11} & \lambda_{11} \end{bmatrix} = \begin{bmatrix} 0.949 & 0.051 \\ 0.051 & 0.949 \end{bmatrix}$$

and for case 18 is

$$\begin{bmatrix} \lambda_{11} & 1 - \lambda_{11} \\ 1 - \lambda_{11} & \lambda_{11} \end{bmatrix} = \begin{bmatrix} 2.859 & -1.859 \\ -1.859 & 2.859 \end{bmatrix}$$

Table 9.10 Open- and closed-loop results with the corresponding relative gain

Case		Steady-state specifications	New value	New distillate x_{EtOH}	New bottoms x_{EtOH}	Steady-state gain	Relative gain λ_{11}
10		$L = 60$	$L = 70$				0.9487
	Open loop	$B = 70$		0.8239	0.2180	$g_{11} = 7.4 \times 10^{-4}$	
	Closed loop	$Bx_{EtOH} = 0.2214$	$L = 70$	0.8243	0.2214	$g_{11}^* = 7.8 \times 10^{-4}$	
18		$L = 60$	$L = 70$				2.8590
	Open loop	Reboiler duty = 4.2×10^6		0.8388	0.2880	$g_{11} = 2.23 \times 10^{-3}$	
	Closed loop	$Bx_{EtOH} = 0.2214$	$L = 70$	0.8243	0.2214	$g_{11}^* = 7.8 \times 10^{-4}$	

Using the three RGA matrices calculated using the steady-state model of the distillation column, it can be concluded that case 10 would result in the best pairing of measured variables to manipulated variables. The RGA matrix associated with case 10 has elements that approach unity, indicating very little interaction. Case 10 uses the distillate flow rate D to control the top composition and the reboiler duty Q_R to control the bottom composition.

Niederlinski index

The NI [4] can be calculated from the full steady-state gain matrix. Using the steady-state model of the distillation column, the remaining elements for the steady-state gain matrix can be calculated for case 10. The resulting matrix is

$$G = \begin{bmatrix} 7.4 \times 10^{-4} & -3.4 \times 10^{-4} \\ 1.7 \times 10^{-3} & 7.0 \times 10^{-3} \end{bmatrix}$$

The NI can then be calculated from

$$\mathrm{NI} = \frac{|G|}{\prod_{i=1}^{n} g_{ii}} = \frac{\left|\left(7.4 \times 10^{-4}\right)\left(7.0 \times 10^{-3}\right) - \left(-3.4 \times 10^{-4}\right)\left(1.7 \times 10^{-3}\right)\right|}{\left(7.4 \times 10^{-4}\right)\left(7.0 \times 10^{-3}\right)} = 1.11$$

Because NI > 0, the control pairing cannot be ruled out because it is definitely unstable. In this 2×2 case, the NI indicates that the system is stable. However, as mentioned earlier, a positive index value for higher-order systems would indicate only that the system is not definitely unstable; in other words, a positive index value does not indicate stability for higher-order systems: the system may or may not be unstable. Therefore, one should also test the selected control scheme extensively via dynamic simulation before adoption.

Singular value decomposition

The SVD may now be calculated for this example from the steady-state gain matrix.

$$b = g_{11}^2 + g_{12}^2 = (7.4 \times 10^{-4})^2 + (-3.4 \times 10^{-4})^2 = 6.63 \times 10^{-7}$$

$$c = g_{11}g_{21} + g_{12}g_{22} = (7.4 \times 10^{-4}) \times (1.7 \times 10^{-3}) + (-3.4 \times 10^{-4}) \times (7.0 \times 10^{-3})$$
$$= -1.12 \times 10^{-6}$$

$$d = g_{21}^2 + g_{22}^2 = (1.7 \times 10^{-3})^2 + (7.0 \times 10^{-3})^2 = 5.19 \times 10^{-5}$$

resulting in

$$\lambda_1 = s_1^2 = \frac{(b+d) + \sqrt{(b-d)^2 + 4c^2}}{2}$$

$$s_1^2 = \frac{(6.63 \times 10^{-7} + 5.19 \times 10^{-5}) + \sqrt{(6.63 \times 10^{-7} - 5.19 \times 10^{-5})^2 + 4(-1.12 \times 10^{-6})^2}}{2}$$

$$s_1^2 = 5.19 \times 10^{-5}$$

$$\lambda_2 = s_2^2 = \frac{bd - c^2}{s_1^2} = \frac{(6.63 \times 10^{-7})(5.19 \times 10^{-5}) - (-1.12 \times 10^{-6})^2}{5.19 \times 10^{-5}} = 6.39 \times 10^{-7}$$

and

$$\Sigma = \begin{bmatrix} s_1 & 0 \\ 0 & s_2 \end{bmatrix} = \begin{bmatrix} 7.20 \times 10^{-3} & 0 \\ 0 & 7.99 \times 10^{-4} \end{bmatrix}$$

$$\mathrm{CN} = \frac{s_1}{s_2} = \frac{7.20 \times 10^{-3}}{7.99 \times 10^{-4}} \approx 9$$

The condition number CN is less than 50; therefore, this system is not prone to sensitivity problems (therefore, a small error in process gain will not cause a large error in the controller's reactions) [2].

9.7 Summary

In this chapter, guidelines for pairing input and output variables in a multi-input–multi-output control system have been presented using the relative gain analysis. The resulting pairings can then be tested to determine whether they are definitely unstable with the NI. SVD may be used to determine whether the control loops are overly sensitive to errors in process gain, as well as whether the control loops may be decoupled. An in-depth discussion on these subjects is presented in the references by McAvoy [2] and Ogunnaike [3]. Dynamic simulation is a powerful tool to be used to test the viability of a control scheme during various process disturbances.

9.8 References

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10
Plant-wide control

The fundamental questions in plant-wide control are whether the feed rates can simply be set for a process and left unattended and whether the process is meeting the desired purity and quality specifications [1,2]. What happens when common disturbances occur such as feed composition changes, production rate changes, product mix or purity specification changes, ambient temperature changes, or measurement sensors either fail or are in error? This chapter covers some of the most common problem areas encountered when designing a plant-wide control scheme.

10.1 Short-term versus long-term control focus

When applying a plant-wide control scheme, it is important to be aware of the propagation of variation and the transformation that each control system performs. Management of that variation is the key to good plant-wide operation and control. A healthy variation management strategy should have both a short-term and a long-term focus. The short-term focus is to use control strategies to transform the variation to less harmful locations in the plant. The long-term focus should concentrate on improvements that reduce or eliminate either the variations or the problems caused by variations.

To illustrate better the idea of short- and long-term control focus, consider an acid recovery plant [3,4]. An example of a short-term focused control scheme for the plant is shown in Figure 10.1. In this system, acid feeds of varying concentrations are pumped to four storage tanks. Tank A contains high-concentration acid that varies greatly in concentration. Tank B is fed with slightly less-concentrated acid, and a feed that varies noisily. Tank C is fed by streams that are similar to Tank B but vary to a lesser amplitude. Tank D is fed by a stream that has a much gentler, but increasing, variance.

The acid feed is then sent from the tanks to a separation system. This separation system removes water and other impurities to produce the final anhydrous-grade product.

The control scheme shown in Figure 10.1 attempts to minimize variance but, unfortunately, passes along much of the disturbance to the extraction column and the azeotropic column. Consequently, the desired product, which is the bottoms of the azeotropic column, varies significantly in quality.

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Figure 10.1 Acid recovery plant control scheme [3] (Stahl, *Competing Globally Through Customer Value* 1991. Reproduced with permission of Greenwood Publishing Group, Inc., Wesport, CT)

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Figure 10.2 Revised acid recovery plant control scheme [3] (Stahl, *Competing Globally Through Customer Value* 1991. Reproduced with permission of Greenwood Publishing Group, Inc., Wesport, CT)

The long-term focus strategy for the plant involves redesigning the feed inventory system to filter out the high-frequency variations. Figure 10.2 shows the same acid recovery system with a different configuration that helps achieve this long-term focus. All the high-concentration feeds are collected in Tank A, and Tank B gathers the low-concentration feeds. The feed from Tank A is then sent to Tank B at a constant rate, thus eliminating some of the problems in variation seen in the short-term focus scheme.

The other major change to the control scheme is in the control systems used for the feed inventory. Level controllers are used on Tanks B and C. These level controllers use the capacitance of the tanks to attenuate the fluctuations in feed flow. The feeds to the extraction column and azeotropic column are considerably dampened, resulting in a much more consistent end product.

10.2 Cascaded units

The dynamics and control of continuous process units that operate as a cascade of units, either in parallel or in series, have been studied extensively for many years [5–7]. A wealth of knowledge is available to help design effective control systems for a large number of unit operations when these units are run independently [6,8]. This knowledge can be directly applied to the plant-wide control problem if a number of process units are linked together as a sequence of units. Each downstream unit simply sees the disturbances coming from its upstream neighbour.

The design procedure was proposed almost three decades ago [5] and has since been widely used in industry. The first step of the procedure is to lay out a logical and consistent ‘material balance’ control structure that handles the inventory controls, i.e. levels and pressures. This hydraulic structure provides gradual and smooth flow rate changes from unit to unit. Thus, flow rate disturbances are filtered so that they are attenuated and not amplified as they work their way down through the cascade of units. Slow-acting, proportional-only level controllers provide the simplest and most effective way to achieve this flow smoothing.

Then, product quality control loops are closed on each of the individual units. These loops typically use fast PI controllers to hold product streams as close as possible to specification values. Since these loops are considerably faster than the slow inventory loops, interaction between the two is generally not a problem. Also, since the manipulated variables used to hold product qualities are often streams that are internal to each individual unit, changes in these manipulated variables have little effect on the downstream process. The manipulated variables frequently are utility streams that are provided by the plant utility system, i.e. cooling water, steam, refrigerant, etc. Thus, the boiler house will be disturbed, but the other process units in the plant will not see disturbances coming from upstream process units. Of course, this is only true when the plant utilities systems have effective control systems that can respond quickly to the many disturbances that they see coming in from units all over the plant.

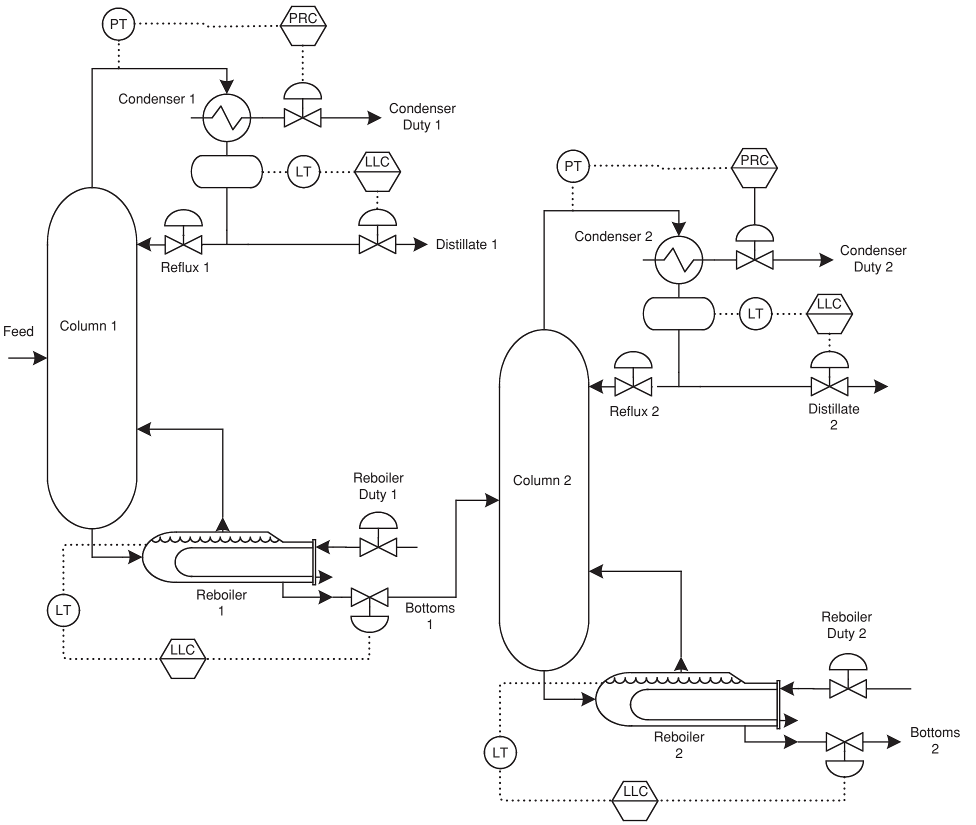

Figure 10.3 Cascade system with two distillation columns in series

As an example of a cascade system, consider a sequence of distillation columns in which the bottoms of the first column feeds the downstream column, shown in Figure 10.3.

Figure 10.3 shows the column with the inventory loops closed. Now that the inventory loops have been closed, the product quality loops can be chosen. Each column has two degrees of freedom remaining, i.e. reflux and vapour boil-up, so some combination of two variables can be controlled in each column, i.e. two compositions, two temperatures, or one temperature and one flow. Vapour boil-up changes require changes in steam flow to the reboiler and also in cooling water flow indirectly through the pressure controller. Both vapour boil-up and reflux changes affect the two liquid levels and, therefore, the distillate and bottoms flow rates, but proportional level controllers usually provide effective filtering of these disturbances. Based on these guidelines and the information provided in Chapter 8, the product quality loops can be closed with the specifics of the loops depending on the control objectives.

Since the propagation of the disturbances in such a system is sequential down the flow path, the use of feedforward control on each unit can also help to improve product quality control [7].

It should be noted that the inventory controls can be in the direction of the flow, i.e. products come off due to level control, or in the opposite direction, i.e. feed is brought in on level control. The same design procedure applies.

10.3 Recycle streams

If recycle streams exist in the plant, then the procedure for designing an effective plant-wide control scheme becomes more complicated. Processes with recycle streams are quite common, but their dynamics are often poorly understood.

The typical approach in the past for plants with recycle streams has been to install large surge tanks. This isolates the unit interactions from one another and permits the use of conventional cascade process design procedures. However, this practice can be very expensive in terms of capital costs and working capital investment. In addition, and increasingly more important, the large inventories of chemicals can greatly increase safety and environmental hazards if dangerous or environmentally unfriendly chemicals are involved.

To demonstrate the principles of plant-wide control for a recycle system, consider the ethylene glycol plant shown in Figure 10.4. Equivalent amounts of water and ethylene oxide are fed to a reactor, as dictated by the reaction stoichiometry, to produce ethylene glycol. The liquid product stream is sent to a distillation column to separate unreacted water and ethylene oxide from the ethylene glycol. The unreacted feed is sent back through a recycle loop to the reactor.

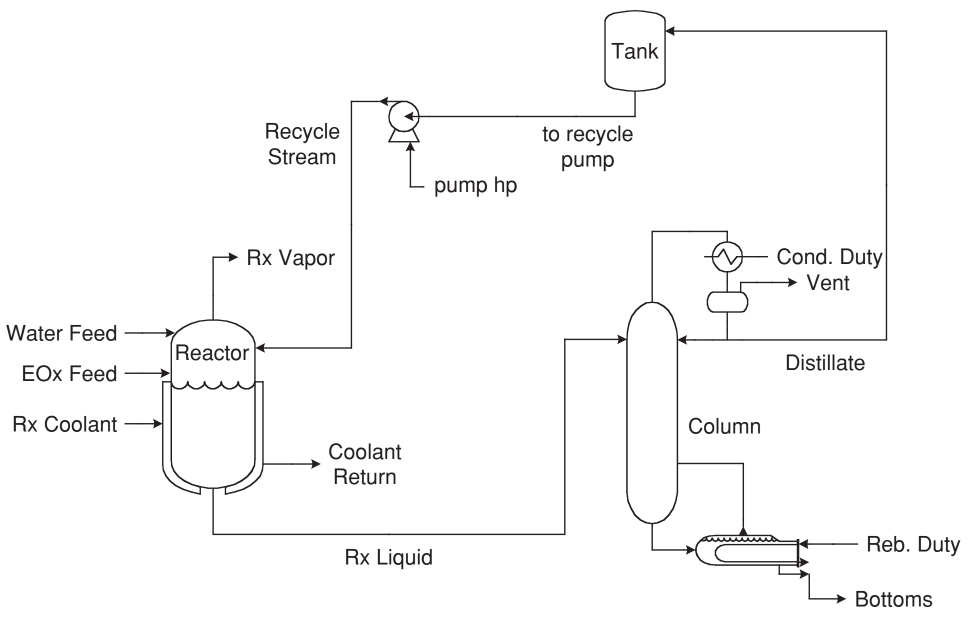

Figure 10.4 Ethylene glycol plant

The reactor control problem is a problem of heat management. The reactor is modelled as a continuously stirred tank reactor (CSTR) with a cooling jacket. As such, the reactor temperature can be measured and controlled by adjusting the rate of cooling flow through the jacket until a desired reactor operating temperature is reached.

The problem of distillation control was addressed in Chapter 8. The issue now is how to control the reactor liquid level, the recycle tank liquid level, the recycle flow rate, the ethylene oxide feed flow rate, and the water feed flow rate.

The biggest danger in the operation of the whole plant is the 'snowball effect' in the recycle [9]. This effect occurs when material accumulates within the recycle loops and cannot be removed. As a result, the plant shuts down. A comparison of two plant-wide control schemes will be made to demonstrate their respective advantages and disadvantages.

The first control scheme involves controlling the level of the reactor by manipulating the flow rate of the reactor effluent. The flow rates of the reactor feed streams are controlled through a ratio controller to meet the required feed ratio. Finally, manipulating the flow rate of the stream to the recycle pump controls the level of the recycle tank. This control scheme is shown in Figure 10.5.

To test the weakness or robustness of this first control scheme, a measurement error is introduced to the flow controller manipulating the water feed flow rate. The water feed flow controller receives a signal that is too low. It adjusts the flow to meet the current set point, when in fact it is supplying excess water. The ethylene oxide flow

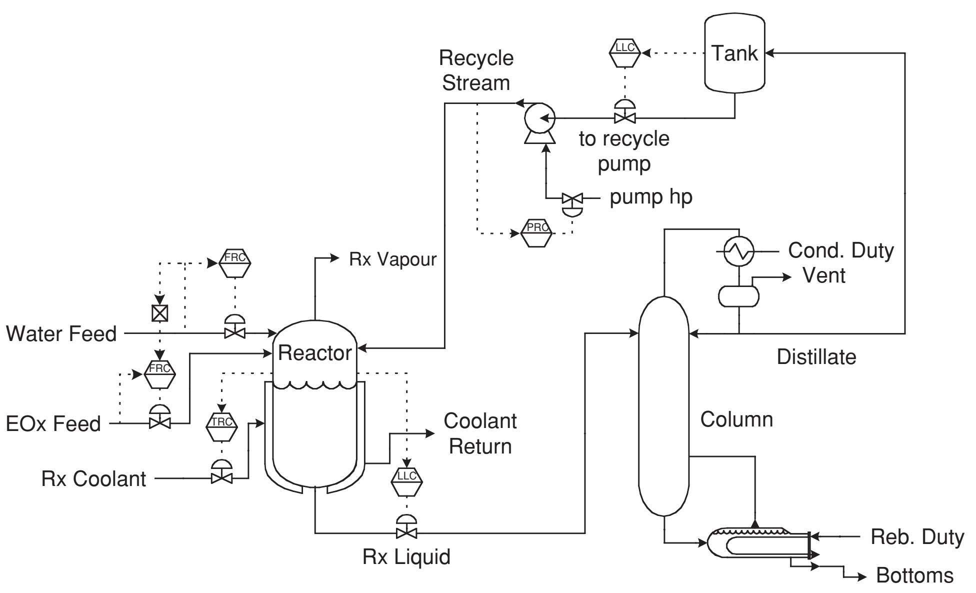

Figure 10.5 Ethylene glycol plant control scheme 1

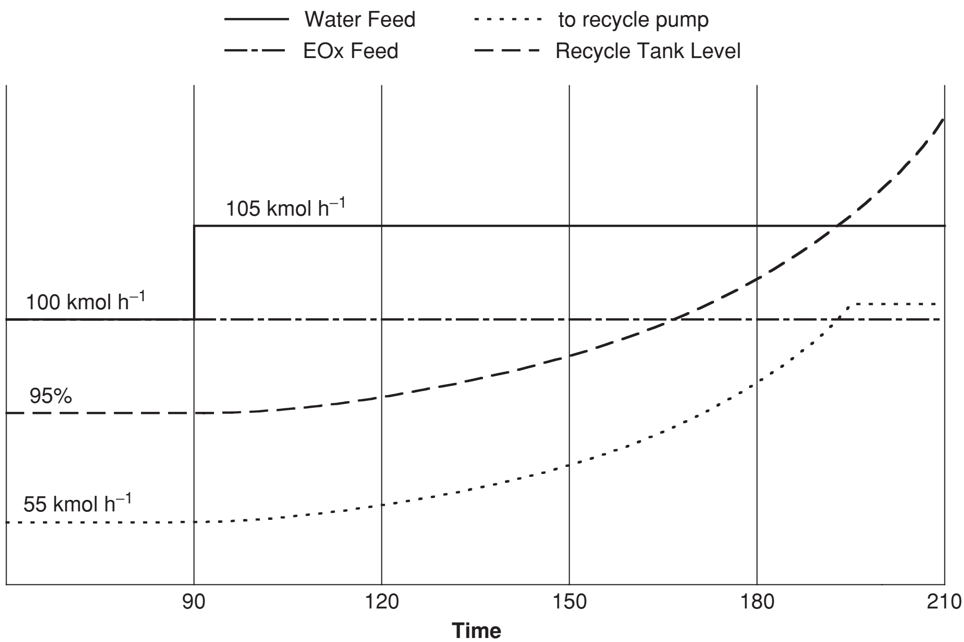

Figure 10.6 Control scheme 1 response to a measurement error

controller moves to match the water feed flow rate through a cascaded ratio controller. The ratio is 1:1 to supply equal amounts of water and ethylene oxide to the reactor.

When excess water is added to the system, the level of the reactor increases. The level controller increases the liquid flow rate, leaving the reactor to compensate. Assuming the distillation column separates the ternary mixture almost perfectly, the unreacted ethylene oxide and the excess water are driven overhead into the distillate stream. This stream feeds the recycle tank, and thus increases the level. The flow rate of the stream to the recycle pump is increased to compensate. This increased flow is recycled to the reactor and increases the level. The cycle begins again, which results in accumulation of water in the system. The recycle stream ‘snowballs’.

To illustrate this concept of snowballing better, a strip chart was recorded for the appropriate variables in the plant using dynamic simulation (see Figure 10.6). The ethylene glycol plant was set with a recycle tank level of 95 per cent and with a valve size on the stream to the recycle pump which results in almost saturated flow, i.e. the valve is almost fully open. The excess water increases the level in the recycle tank, thus opening the valve on the stream to the recycle pump even further until it saturates. The recycle tank then continues to increase, past the 100 per cent point where the tank begins to overflow. The plant must shut down.

If a positive measurement error is supplied, then the flow sensor transmits a flow that is too large. Since the ethylene oxide controller is set up so that the set point is in a 1:1 ratio to the water feed flow rate due to the reaction stoichiometry, too much ethylene oxide enters the system. If this is the case, then the excess ethylene oxide reacts with the surplus water in the recycle loop, thus consuming the water and producing ethylene glycol. This reduces the material inventory within the plant until there is only ethylene oxide remaining.

Although there are a number of different ways to control this plant, it is helpful to keep in mind two fundamental rules of plant wide control. These are affectionately known as 'Luyben's rules', referring to the original author [10]:

1 Only flow control a feed if it is sure to be fully consumed in the reaction.

2 Always put a flow controller on one stream in the recycle path on flow control.

With these rules in mind, a new control scheme can be proposed, illustrated in Figure 10.7. This time the stream to the recycle pump is under flow control. In order to control the level of the recycle tank, it is necessary to manipulate the water feed rate. However, with the water feed introduced to the reactor, a considerable amount of dead time is unnecessarily introduced to the system. To overcome this dead time, the water feed is introduced directly into the recycle tank and is used to control the liquid percent case level.

In order to introduce a similar transmitter error used in the previous control scheme, the recycle tank level controller output is cascaded to provide a set point for a flow controller on the water feed stream. The water feed flow controller uses the measured flow, complete with an error. The ethylene oxide stream is controlled using a composition controller that manipulates the ethylene oxide flow rate to meet a specified composition in the liquid stream leaving the reactor. The liquid level of the reactor will be controlled using the same controller manipulating the flow rate of the liquid leaving the reactor. With these controllers in place, the process flow schematic has been modified. The updated schematic is shown in Figure 10.7.

This control scheme is more robust in the event of a disturbance. The flow rate of the stream to the recycle pump is controlled, preventing any increases in the recycle flow rate. The recycle tank level is controlled by the water feed flow rate. The reactor is still under level control by manipulating the flow rate of the liquid stream leaving the reactor. The ethylene oxide feed flow rate is also manipulated by a composition controller, which measures the exit composition of ethylene oxide from the reactor. This control scheme does not allow for excess ethylene oxide or for excess water in the system, and hence this system cannot snowball.

To demonstrate this system's robustness, the same measurement error can be introduced to the water feed flow rate. The strip chart in Figure 10.8 shows an introduction of -5 kmol h^{-1} error into the sensor transmitting a flow measurement to the water feed flow controller.

With an increase in the amount of water fed to the recycle tank, the level controller adjusts the set point of the water feed flow controller and reduces the amount of water being introduced to the plant. The recycle tank level attains a new operating level. Remember, there will always be offset from the set point due to using only proportional

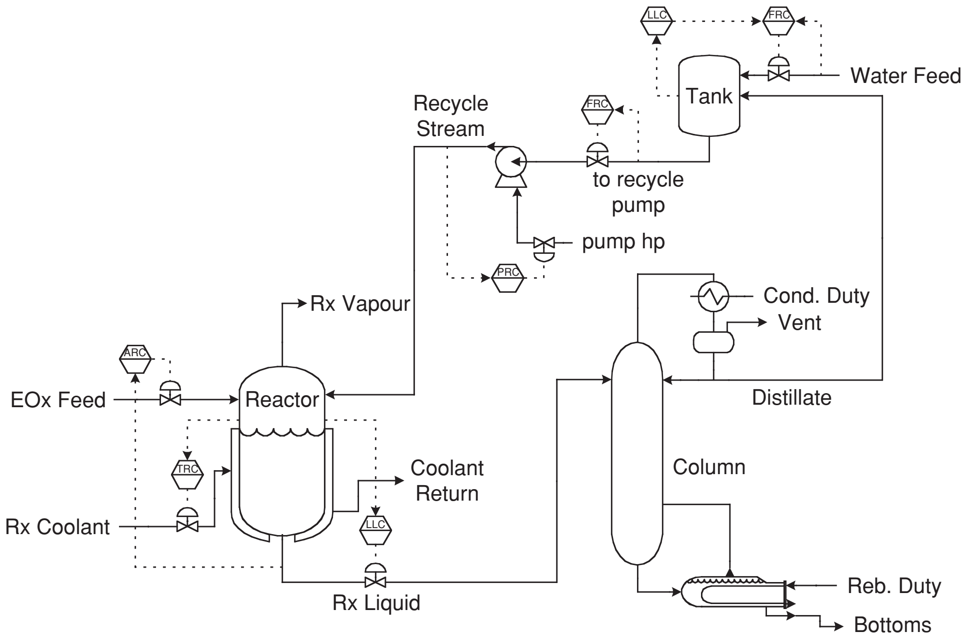

Figure 10.7 Ethylene glycol plant control scheme 2

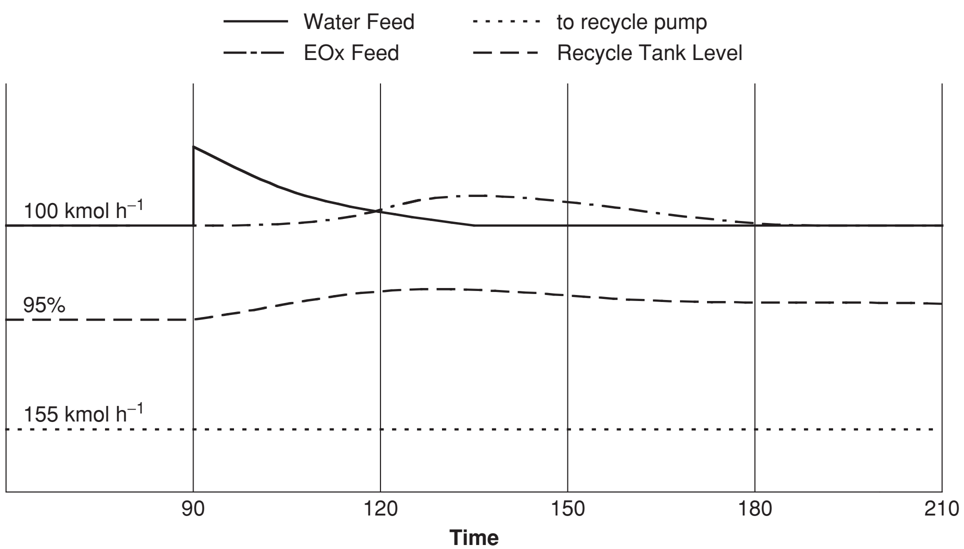

Figure 10.8 Control scheme 2 response to measurement error

control for the level control (refer to Chapter 7). More importantly, notice that there is no accumulation within the system.

10.4 General considerations for plant-wide control

When considering plant-wide control, a number of questions must be answered:

- What are the primary objectives of the plant?
- Where are the production bottlenecks and constraints?
- Where should the production rate be set?
- Where are the bulk inventories and how should they be controlled?
- Will additional inventory improve the operation and control?
- Will changes in the process design improve the operation and control?
- Where should recycle streams be placed?
- How are the component inventories controlled in these recycle systems?
- Will small changes in a feed cause a very large change in the recycle rate around the system (‘snowball effect’)?
- What are the primary sources of variation?
- What can be done to reduce or eliminate variation at the source?
- How does variation propagate through a plant-wide system?
- What can be done to transfer the variation to less harmful locations?
- How much of the plant-wide operation should be automated and how much should be left for the operator?

Plant-wide design, operation, and control is a fast-developing area for research. As such, it cannot be summarized simply in one chapter. For a more in-depth discussion of this topic, refer to the series of papers authored by Luyben and co-workers [9,11–14], and most recently a book entitled *Plant Wide Process Control* by Luyben *et al*. [10]. A practical article providing guidelines to ensure smooth plant operation is given by Lieberman [15], and a book on *Plantwide Process Control* is presented by Erickson and Hedrick [16].

10.5 References

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11. Luyben, W. L. Dynamics and control of recycle streams. 2. Comparison of alternative process designs. *Ind. Eng. Chem. Res.*, 1993, **32**: 476–86.
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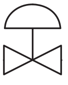
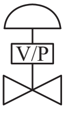
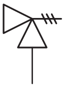
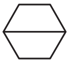
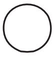
14. Luyben, M. L., Tyreus, B. D. and Luyben, W. L. Plant wide control design procedure. *AIChE J.*, 1997, **43**(12): 3161–74.
15. Lieberman, N. Instrumenting a plant to run smoothly. *Chem. Eng.*, 1977, **84**(19): 140–54.
16. Erickson, K. T. and Hedrick, J. L. *Plantwide Process Control*. Wiley, New York, 1999.

Appendix 1: P&ID symbols

Symbol	Description
	Control valve
	Valve
	Control valve with valve positioner
	Check valve
	Pressure relief valve
	Controller
	Transmitter/sensor
	Transmitter/sensor
	Controller summer
	Summer or multiplier
	Divider

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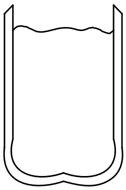
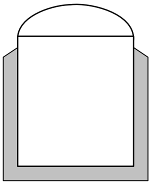
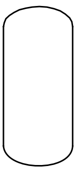
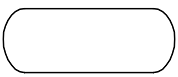
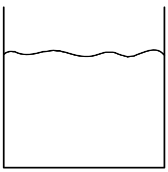
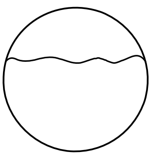
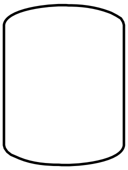
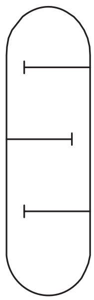
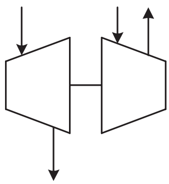
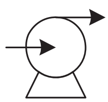
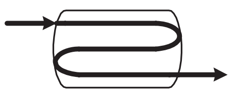
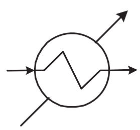
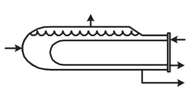
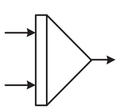
Symbol	Description
$\boxed{<}$	Selector
	Reactor with cooling jacket
	Reactor with cooling jacket
	Knock out drum
	Reflux drum
	Tank
	Horizontal tank
	Vertical tank

Symbol	Description
	Distillation column tray section
	Compressor
	Pump
	Heat exchanger
	Heat exchanger
	Kettle reboiler
	Stream mixer

Appendix 2: Glossary of terms

The following list has been made with reference to Murrill [1].

Actuator	Portion of the valve that may be pneumatic or motor driven, used to open and close automatic valves.
Amplitude	The difference between the average value of a sinusoidal variation and the maximum or minimum value.
Amplitude ratio	The ratio of the amplitude of a system's response to its forcing function's amplitude when the forcing function is a continuous sinusoid; a form of dynamic gain.
Analog controller	A controller that operates on continuous signals such as voltages, pressures, or currents; this is a common older type of controller, as distinguished from a digital controller.
Attenuation	A decrease in the strength of a signal by a system component.
Automatic controller	A device which operates to correct or limit the deviation of a variable from a reference value.
Automation	The use of automatic control devices in a process so that human supervision is minimized.
Capacitance	The amount of energy or material which must be added to a closed system to cause a unit change in potential; the partial derivative of content with potential.
Cascade	A series of stages in which the output of one is the input to the next.
Cascade control	Automatic control involving 'cascading' of controllers; that is, having one controller's output as the input to the next controller instead of manipulating a process variable directly.

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Comparator	The portion of the control element which determines the difference between the set point and the measured feedback variable.
Compensator	A component added to a system to improve the characteristics of its response.
Control element	The portion of the control system which relates the error between the desired value and the manipulated variable.
Controlled variable	That quantity or condition of the controlled system which is controlled.
Controller	A device that receives the set point and feedback signals, computes the difference, and uses adjustable parameters to produce an output signal to eliminate the difference between the set point and feedback signals.
Critical gain	A value of system gain beyond which closed-loop operation is unstable.
Cycling	Periodic changes in the controlled variable. Also known as *oscillation*.
Damping ratio	Also known as *damping coefficient* and *damping factor* η which characterizes the nature of damping of the transient response.
Dead band	The largest range of values of the input variable to which a component does not respond.
Dead time	An interval of time between an input to a component and the beginning of response to the input.
Derivative action	A controller mode in which there is a continuous linear relationship between the derivative of the error signal and the controller output signal. Also known as *rate action*.
Derivative time	The time difference by which the output of a proportional-plus-derivative controller leads the controller input when the input changes linearly with time.
Desired value	The value of the controlled variable which is desired. Also known as *set point*.
Deviation	The difference at any instant between the value of the controlled variable and the set point.
Digital controller	A controller which operates on signals, usually electronic, which have only discrete numerical values.

Distance–velocity lag	The effect resulting from a signal being transmitted over an appreciable distance at a finite velocity; a kind of dead time. Also known as *transportation lag* and *time delay*.
Disturbance	An input signal other than the set point which directly affects the controlled variable. Also known as *load*.
Error	In measurement, the difference between the value found and the true value; in control, the set point minus the measured value of the feedback variable.
Feedback	The signal to the controller representing the condition of the controlled variable; a control system in which corrective action is based on such signals.
Feedback elements	The portion of the control loop which establishes the primary feedback variable in terms of the controlled variable.
Feedforward	A control mechanism in which corrective action is based on measurements of inputs to the process. A form of *predictive control*.
Final control element	The controlling means or element which directly changes the manipulated variable.
First-order system	A system whose dynamic behaviour is described by a first-order linear differential equation.
Frequency response	The amplitude ratio and the difference in phase of a system or element's output with respect to a sinusoidal input. The frequency of the input and the system or element's differential equation determine the frequency response.
Gain	The proportionality constant in a transfer function.
Impulse	A sharp increase or decrease in a variable immediately followed by a return to its original value.
Input	A variable that is dependent only on conditions outside the system.
Input elements	The portion of the control system which provides a reference input to a comparator in response to the set point.
Integral action	A controller mode in which there is a continuous linear relation between the integral of the error signal and the output signal of the controller.
Integral time	For a step input it is the time required for the output of a proportional-plus-integral controller to change an amount equal to the proportional response alone.

Interacting	Two or more consecutive stages whose combined transfer function is not the product of the transfer functions of the preceding stages if they would appear alone.
Lag	The retardation of one condition with respect to another.
Linear	A relationship showing output proportional to input; a system whose behaviour is adequately described by such equations; a system that follows the principle of superposition.
Linearize	To substitute an approximate linear function for a nonlinear function.
Load or load variable	Any outside input to a control system except the set point. Also known as *disturbance*.
Load change	A change in input conditions such that a change in manipulated variable is necessary to maintain the controlled variable at the set point.
Loop	A series of stages forming a closed path.
Lumping	An assumption that the effects of two or more aspects of a system can be considered together as a single quantity; an assumption that a parameter distributed over space may be considered at a single point in space.
Manipulated variable	The process variable that is changed by the controller to eliminate error.
Mode	The classification of a controller by the manner in which the manipulated variable responds to the error signal.
Model	A conceptual approximation of a physical system that is usually mathematical in nature.
Natural frequency	The frequency of oscillation that a system would have if there were no damping.
Noise	Accidental and unwanted fluctuations in a variable that tend to conceal it.
Nonlinear	An equation that contains a term not conforming to linearity; a system whose behaviour is not described by linear equations.
Offset	The steady-state deviation in the controlled variable caused by a change in the load variable.
On–off control	A system of regulation in which the manipulated variable has only two possible values, high and low, maximum and minimum, or on and off. Also known as *two-position control*.

Output	The variable that is chosen to describe the condition of a system; the dependent variable in the dynamic equation.
Oscillation	See *cycling*.
Overdamped	Said of a system of second or higher order whose transient response has no tendency to oscillate or overshoot.
Overshoot	In a step response, the difference between the final steady-state value and the value of the first maximum (or minimum if the response is downward); often expressed as a fraction; only defined for underdamped systems.
Parameter	A constant coefficient in an equation that is determined by the properties of the system.
Period	The amount of time between consecutively recurring conditions, the reciprocal of frequency.
Predictive control	A control scheme that predicts the effect of a load change and takes corrective action before the controlled variable is affected, e.g. *feedforward control*.
Primary element	That portion of the measuring means which first senses a change in the controlled variable. Also known as a *sensor/transmitter*.
Process	The system being controlled.
Proportional action	A controller mode in which there is a continuous linear relation between the value of the error signal and the value of the controller output.
Proportional band	The range of the controlled variable that corresponds to the full range of the final control element.
Proportional sensitivity	A proportional action; the steady-state ratio of the controller output to the error signal.
Rangeability	The ratio of maximum flow to minimum controllable flow in a final control element.
Rate action	See *derivative action*.
Reset rate	The inverse of integral time; usually expressed as repeats per unit time.
Resistance	The potential required to produce change; the partial derivative of driving force with flow rate.
Response	A system's output due to a change in its input.

Response time	The time required for an output to increase from one specified percentage of its final value to another, based on a step input.
Self-regulation	The inherent characteristic of a system that produces a steady state without the aid of automatic control.
Sensor/transmitter	See *primary element*.
Set point	See *desired value*.
Settling time	The time required for the absolute value of the difference between the output of a component or system and its final value to become and remain less than a specified amount.
Signal	Information in transmission.
Stable	A system whose response to a bounded input is also bounded.
Steady-state	The condition when all properties are constant with time, the transient response having died out.
Steady-state error	A control error at steady state.
Time constant	The time required for the output of a first-order system to change 63.2 per cent of the amount of total response to a step forcing function.
Transducer	Any device that transmits, amplifies or changes a signal.
Transfer function	A mathematical relationship that describes the ratio of an output of a system to the input to the system.
Transient response	That part of a system's response that approaches zero as time proceeds.
Transportation lag	See *distance–velocity lag*.
Two-position control	See *on–off control*.
Undamped	Oscillatory transient response of constant amplitude.
Underdamped	Oscillatory transient response of diminishing amplitude.
Valve cage	A cage that surrounds the valve plug in a cage-guided valve that guides the valve plug towards the valve seat.
Valve plug	The part of the valve that restricts flow through the valve.
Valve positioner	A device that precisely controls the control valve stem position by adjusting the instrument air pressure to the control valve.
Valve seat	The part of the valve that the valve plug rests upon when the valve is fully closed.

Valve stem A connecting rod between the diaphragm in a valve actuator and the valve plug that allows air pressure on the diaphragm to control the valve plug position in the valve and hence flow through the valve.

A2.1 Reference

1. Murrill, P.W. *Automatic Control of Processes*. International Textbook Company, Scranton, PA, 1967, pp. 451–59.

Workshops

A Real-Time Approach to Process Control, Second Edition W. Y. Svrcek, D. P. Mahoney and B. R. Young
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Do or do not, there is no try.
Yoda

Workshop 1 Learning through doing

Course philosophy: 'Learning through doing'

In conjunction with this workshop, you should review Chapter 1 in the book.

This course consists of a set of learning modules or workshops, each of which is intended to enhance your understanding of process control and simulation theory and application through hands-on experience using the latest simulation technology and actual laboratory experiments.

Key learning objectives

1 Develop an understanding of the general organization of the course and what is expected.

2 Understand how to proceed to the course.

3 Develop a working knowledge of the HYSYS steady-state dynamic simulation package.

4 Understand the fundamentals of steady-state and dynamic process simulation.

Course coverage

This course will deal with fundamental and underlying principles of automatic process control and simulation. The course covers the theory associated with SISO loops and how these are configured into multi-loop schemes to control complex unit operations and entire plants. It will not discuss in detail the hardware or individual measurement techniques, such as flow, temperature, pressure, etc., except as the measurement affects the control loop. In the course, note that Chapter 2 provides a simplified summary of control loop hardware.

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Prerequisites

No elaborate prerequisites are required; however, an understanding of unit operation modelling is assumed. It is inevitable that modelling involving differential equations will, by necessity, be involved in parts of the theory and workshops. Quite often, mathematics is a barrier that prevents a clear understanding of control concepts and implementation of process control theory. It is anticipated that the 'real-time approach' will remove, or at least minimize, these barriers.

Study material

The course notes and the HYSYS process simulator are the only materials required for the course. The notes are independent, unique, stand-alone, and specifically designed for the tutorial/workshop approach.

The references that are provided at the end of each chapter detail additional selected study and reference reading. The available literature on the subject of process dynamics and control is massive. Additional literature is readily available from instrumentation vendors, such as Honeywell, Foxboro, Fisher, etc. This vibrant area of chemical engineering is represented by the Instrument Society of America (ISA), PO Box 12277, Research Triangle Park, NC 27709-2277, USA.

Organization

The course consists of eight workshops and three laboratory sessions. Each workshop has associated with it a specific portion of the book that provides the necessary theoretical background. During the workshop, a specified assignment using the dynamic process simulator will be completed. The results achieved during each workshop, along with an explanation written in Microsoft Word, will be submitted in the form of a diskette at the end of each tutorial session.

Total course objectives

1. Understand the basic theoretical concepts of feedback and SISO loops.

2. Understand the components of a control loop and how they interact.

3. Understand process control terminology.

4. Have an appreciation of process dynamics.

5. Know how to develop the fundamental models for first-order plus dead time processes.
6. Know how to tune controllers.
7. Know how and where to implement such techniques as cascade, feedforward, ratio, dead time, and multi-loop control.
8. Appreciate the use of process simulation in the development and validation of control strategies.
9. Develop an understanding of the unit operation control schemes.
10. Understand what is meant by plant-wide control and be able to implement a plant-wide control strategy.
11. Familiarize yourself with the appropriate simulation software.

Workshop 2 Feedback control loop concepts

What we have to learn to do, we learn by doing.
Aristotle

Introduction

Prior to attempting this workshop, you should review Chapter 3 in the book.

Process systems respond to various disturbances (or stimuli) in many different ways. However, certain types of response are characteristic of specific types of process. The characteristic response of a process can be described as its personality. Process control engineers have developed a range of terms and concepts to describe different process personalities, and they use this knowledge to develop effective control systems.

Two of the most common personalities are those for first and second-order systems. First-order systems may also be called first-order processes or first-order lags and can be mathematically modeled through the use of a first-order differential equation. Shown in Figure W2.1 is the typical step response of a first-order process. The time constant τ was discussed in Chapter 3 and is related to the speed of the process response; the slower the process, the larger the value of τ.

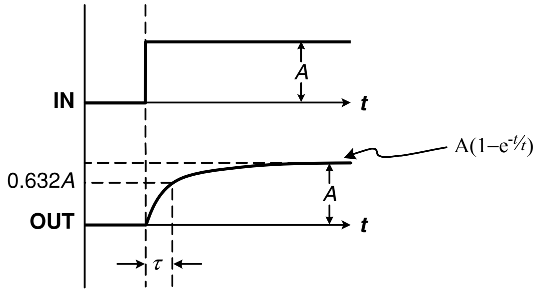

Figure W2.1 First-order process response to a step change

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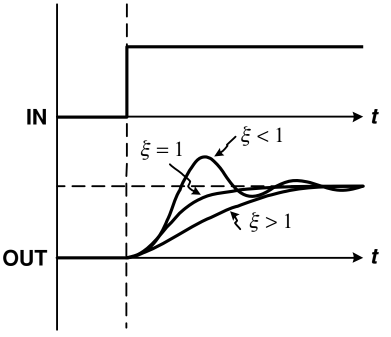

Figure W2.2 First-order process response to a step change

Unlike first-order processes, second-order processes can have several different types of response. Second-order processes are more complex than first-order processes, and hence the mathematical models used to describe these processes are also more complex. There are three types of second-order system to consider. The key parameter in determining the type of system is the damping coefficient ξ. When $\xi < 1$, the system is underdamped and has an oscillatory response, as shown is Figure W2.2. An underdamped system overshoots the final value and the degree of overshoot is dependent upon the value of ξ. The smaller the value, the greater the overshoot. If $\xi = 1$, then the system is deemed critically damped and has no oscillation. A critically damped system provides the fastest approach to the final value without the overshoot that is found in an underdamped system. Finally, the system is overdamped if $\xi > 1$. An overdamped system is similar to a critically damped system, in that the response never overshoots the final value. However, the approach for an overdamped system is much slower and varies depending upon the value of ξ. The larger the damping coefficient is, the slower the response.

There are two main differences between first- and second-order responses. The first difference is obviously that a second-order response can oscillate, whereas a first-order response cannot. The second difference is the steepness of the slope for the two responses. For a first-order response, the steepest part of the slope is at the beginning, whereas for the second-order response the steepest part of the slope occurs later in the response.

First- and second-order systems are not the only two types of system that exist. There are higher-order systems, such as third- or fourth-order systems. However, these higher-order systems will not be discussed.

Key learning objectives

1 Understand the components of the loop and how these components interact.

2 Become familiar with the terminology associated with process control.

3 Be able to explain the components in a SISO block diagram.

4 Be able to develop the underlying mathematical models and relationships for each component of the feedback control (FBC) loop.

5 Understand the effect of capacitance.

6 Understand the effect of resistance.

7 Understand the concept of a response and the metaphor of process personality.

8 Understand the effect of self-regulation on process response.

9 Recognize the open-loop response of second-order processes.

10 Recognize the open-loop response of capacity-dominated processes, with and without dead time.

Tasks

1 Level response

Capacity-dominated process behaviour can best be studied using a very common process element, namely the surge tank or separator. The ODE for a single tank has been presented in the notes in Chapter 3 and is implemented in HYSYS as the TANK unit operation.

Build a simulation in HYSYS using water with a flow of 20 kmol h^{-1} at 15°C and 1 atm as the only component and the Peng–Robinson equation of state as the fluid property package. Use the default tank volume of 2 m^3 and specify liquid flow control on the Liquid Valve page of the tank unit operation. Calculate the flow out of the tank using Equation W2.1, which describes a linear valve, and then Equation W2.2, which describes a nonlinear valve. In both cases, the outlet flow rate is a function of the liquid head on the tank only.

$$\text{Flow} = K \times \text{head} \tag{W2.1}$$

$$\text{Flow} = K \times \sqrt{\text{head}} \tag{W2.2}$$

In Equations W2.1 and W2.2, SI units are flow in m^3/h and head in metres. Use the spreadsheet unit operation to incorporate Equations W2.1 and W2.2 into the simulation. Import the tank liquid level into the spreadsheet, assume a tank geometry, and use the resulting value of head to calculate the liquid outlet flow. A K value of 0.4 should work well with these units. Then export the calculated flow back to the worksheet as the outlet flow rate. This last action can only be performed in the dynamic mode, otherwise the material balance will be violated.

System identification is the term used to define a procedure to characterize the process response. In this case, system identification can be accomplished by adjusting the feed rate to the tank in steps, up and down, and then observing the tank level response on a strip chart. This is termed step response testing.

Set up strip charts, recording the important variables, to study the open-loop response of capacity dominated processes consisting of:

1 a single tank

2 two tanks in series

3 two tanks in series with a pipe segment between them

 [*Hint: use the purge flow reactor (PFR) unit operation to simulate a pipe segment and calculate a volume that should give a dead time of around 10 min*]

4 three tanks in series (use only linear valves).

- What is the open-loop response of the tank level to a step change in the feed rate in a process with a single tank only?
- What effect does the addition of a second tank have on the level response in Tank 1? And in Tank 2?
- What effect does the addition of a third tank have on the level response in Tank 1? And in Tank 2? What is the open-loop response of Tank 3?
- What effect does the volume of the tank have on the personality of the response?
- How does the valve type affect the process personality?
- What effect does the addition of the pipe segment between the two tanks have on the response of the level in Tank 1 and Tank 2? Did the PFR properly simulate dead time in this situation? Why or why not?

[*Note: the pressure–flow solver is relatively fragile and it is important to build up each case carefully from steady state, defining an appropriate pressure–flow profile and specifications for each sub-task.*]

2 Temperature response

The next exercise in this workshop requires that you set up a mixing tank to heat water directly using live steam, as illustrated in Figure W2.3. Use the default tank volume

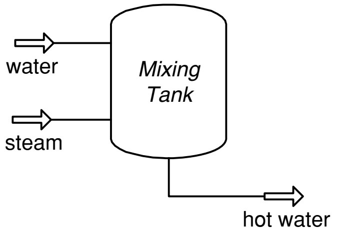

Figure W2.3 Mixing tank process

and set the default tank control to liquid level with a 50 per cent set point. The feed water stream to the tank enters with a flow of 100 kmol h^{-1} at 15°C and 1 atm.

Perform a series of steady-state runs to determine the amount of steam required to raise the temperature of the feed water stream to about 200°F (about 80°C). Then, switch to the dynamic mode of operation and perform step response testing by varying the inlet flow rate and feed temperature to determine the process response. Remember to use the strip charts to observe the important process variables.

Add a pipe segment to the system on the outlet of the tank. Simulate the pipe segment using a PFR, as in the previous exercise. Calculate a PFR volume to approximately 10 min of dead time. Make sure that the number of segments in the PFR operation is set to at least 50 instead of the default of 20 (under 'Reactions'). Repeat your analysis of step disturbances, noting the relationship between the tank temperature and the temperature at the outlet of the pipe segment. Repeat the analysis again, but with a different volume for the PFR, and note any differences in the response.

- What type of response does the process produce?
- How does the pipe segment affect the response?
- What effect does the pipe segment volume have on the response? Did the PFR properly simulate dead time in this situation? Why or why not?

Present your findings on diskette in a short report using MS-Word. Also include on the diskette a copy of the HYSYS files which you used to generate your findings.

Workshop 3 Process capacity and dead time

Knowledge is a treasure but practice is the key to it.
Thomas Fuller

Introduction

Prior to attempting this workshop, you should review Chapter 3 in the book.

This workshop will illustrate the effect on the process response of the three key process dynamic parameters: process gain, process time constant and process dead time. You will also explore the impact that capacitance or 'lag' has on these process parameters.

Key learning objectives

1 Process gain is the key process parameter affecting the extent (magnitude) of the response of a process or process element.

2 The time constant determines the personality of the response for a process or process element.

3 The time constant is the key dynamic parameter that determines the ability of a process to reject, or attenuate, disturbances.

4 The period and the amplitude of the disturbance will determine the amount of attenuation/rejection.

5 Capacitance is good for disturbance rejection, but the down side is that it results in very slow and long response times.

6 Dead time has no effect on the filtering capability of the process.

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7 Dead time has no redeeming features and can make the control loop unstable.

8 Tight process control can only be achieved if the loop dead time is small compared with the smallest time constant of a disturbance of significant amplitude.

Tasks

1 System identification

The process used for this workshop is shown in Figure W3.1. Build a simulation of this system using the Wilson activity model as the fluid package. The feed is a 50/50 mixture of water and methanol (100 kmol at 30°C and 200 kPa) which is heated in a steam heater to about 70°C. The hot mixture is then stored in a surge tank for future use. Note that you do not need to enter any further information about vessel volumes, etc. Simply use the default values for hold-ups for dynamic runs.

This very simple example has many analogies in process plants. Whenever material or energy enters a piece of plant equipment that can accumulate some of the material or energy, the process has capacitance. Virtually all process equipment has the potential to store mass or energy and hence create a capacity dominated process.

System identification is the term used to define a procedure to characterize the process response. In this case, system identification can be accomplished by setting the default level controller set point at 50 per cent (under 'Liquid Valve'), adjusting the steam flow to the heater in steps, up and down, and then observing the temperature response on a strip chart. This is termed step response testing and is the same as was done in the previous workshop.

Figure W3.2 illustrates a typical step-test response for a first-order system. The relevant process parameters of gain K_{p} and time constant τ for this first-order process are shown and can be calculated as follows:

$$K_{\mathrm{p}} = \frac{\text{Tank Temperature Change/Temperature Transmitter Span}}{\text{Steam Rate Change/Steam Valve Span}} \tag{W3.1}$$

$$\tau = \text{Time taken for tank temperature to reach 63.2\% of its final value}$$

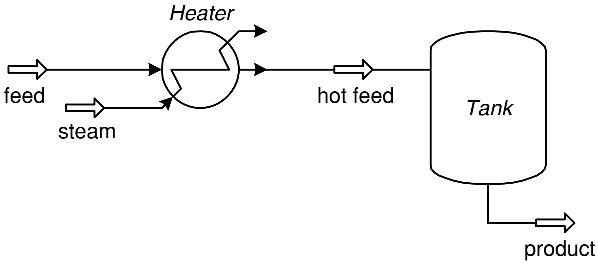

Figure W3.1 Capacity-dominated process

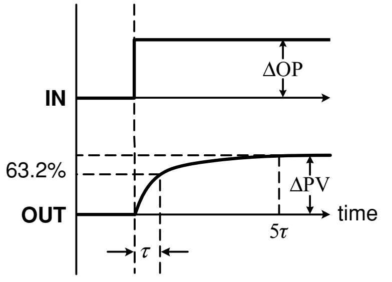

Figure W3.2 System step response

In order to check how linear the process is, it is necessary to determine whether the gain is the same regardless of the steam rate and to see whether the magnitude of the gain is unchanged for increases and decreases in steam. Do this by using the step testing method described above and Equation W3.1.

2 Capacitance

Now we will examine how the gain, time constant, and dead time vary for different tank levels or different amounts of capacitance in the process shown in Figure W3.1.

- Make step changes in the steam rate for three different tank level set points of 5, 50, and 95 per cent. Calculate the gain, time constant, and dead time for the three different process capacities. Present your results in Table W3.1 and then plot the time constant and gain in Figure W3.3.

3 Attenuation

As has been indicated in the notes, the object of both the process and control system is to reject, or at least minimize, the effect of disturbances. In order to quantify disturbance rejection, the term attenuation has been borrowed from the electrical and mechanical engineers and is defined as

$$\text{Attenuation} = 1 - \frac{\text{Disturbance Amplitude Out}}{\text{Disturbance Amplitude In}}$$

Table W3.1 Summary of process parameters

Tank level (%)	Process gain	Time constant (min)	Dead time (min)
5			
50			
95			

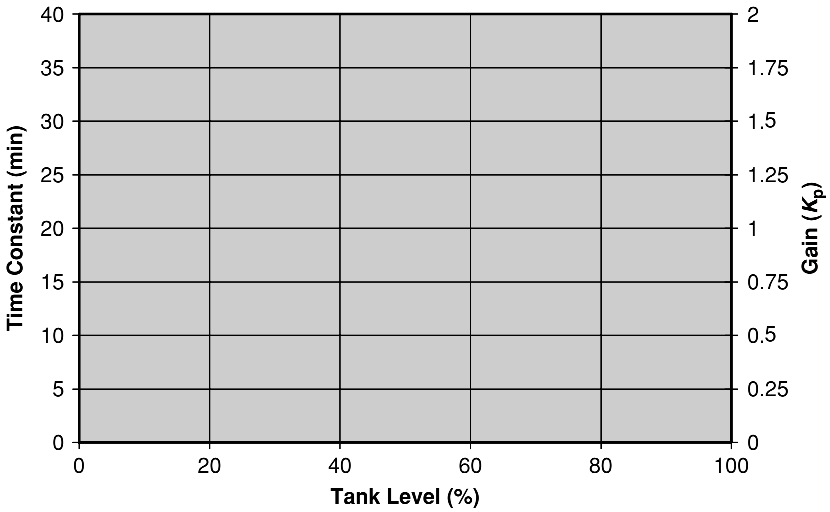

Figure W3.3 Process gain and time constant versus tank level

For example, if the tank temperature varies with an amplitude of 5°C and the input temperature disturbance has an amplitude of 25°C, then the attenuation is $(1 - 5/25) = 0.8$ or 80 per cent.

In HYSYS, a transfer function block is used to generate a sinusoidal feed temperature. From the main simulator builder, or through the PFD, select the transfer function block. On the 'Connections' page attach the output of the transfer function block to the feed 'Object' and select the process variable (Select PV) as temperature. Move to the parameters page and set the nominal (User Input PV) temperature to 30°C and the span of the PV output to vary between 0 and 100°C. To select the actual wave type, move to the page labelled 'Lead/2nd Order' and select 'Sine Wave'. Under 'Sine Wave Parameter', specify an amplitude of 25 and a period of 10 min to start with. It is important to note that the amplitude is entered as a percentage of the PV span, i.e. 25 per cent of 100°C gives an amplitude of 25°C. The disturbance period will be varied depending on the dynamic test being run. Finally, open the faceplate to complete the set up.

- Complete Tables W3.2–W3.4 for each tank level and then plot the results with a curve for each level on Figure W3.4.

Note: you will need your results from this section of the workshop for later workshops, so remember to save a copy of the results for yourself.

4 Dead time

In this section of the workshop the dynamic characteristics of processes with capacitance and appreciable dead time will be studied. The process you will work with is the simple

Table W3.2 Attenuation for level at 5 per cent

Disturbance period (min)	Frequency (min^{-1})	Product ΔT (°C)	Attenuation
5			
10			
20			
30			
50			

Table W3.3 Attenuation for level at 50 per cent

Disturbance period (min)	Frequency (min^{-1})	Product ΔT (°C)	Attenuation
10			
20			
30			
40			
100			

Table W3.4 Attenuation for level at 95 per cent

Disturbance period (min)	Frequency (min^{-1})	Product ΔT (°C)	Attenuation
10			
20			
30			
40			
100			

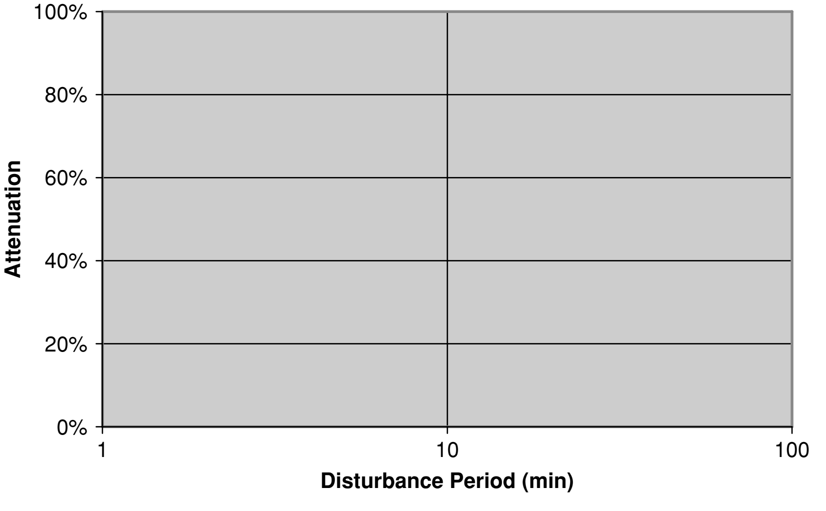

Figure W3.4 Attenuation of feed temperature disturbance first-order process

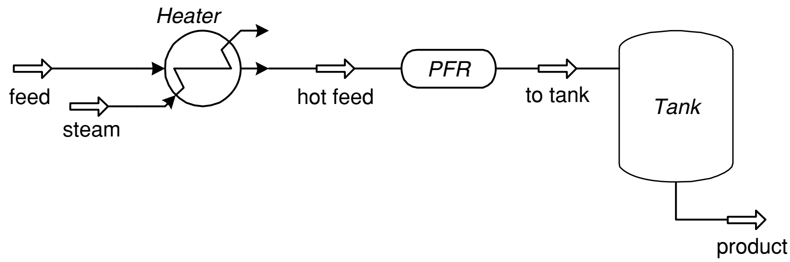

Figure W3.5 A process containing dead time

feed heater and storage tank shown in Figure W3.1, except that additional equipment will be added between the heater and the surge tank. The additional equipment will be a PFR with a volume of 0.3 m^3 and length of 2 m, which gives a dead time of 6 min. Figure W3.5 shows an example of what this process should look like. The objective is to see how dead time affects the temperature response of the warm solution leaving the surge tank.

Again, this simple example is illustrative of many real plant situations. Even the time it takes for fluid to move through pipes between connecting items of equipment is an example of dead time. A sensor located a distance from a vessel such as a reactor introduces process dead time. The time it takes for a process analyser to sample a process stream and measure a particular property or the time it takes for a manual sample to be taken to the laboratory for analysis are also both examples of dead time. This is the time during which there is no knowledge of what is happening in the process.

- Using the built in tank level controller with a set point of 50 per cent, increase and decrease the steam rate to the heater and record the tank temperature response. What is the dead time for the process shown in Figure W3.5? Is there a difference between the dead time predicted and the actual dead time from the simulation? If the answer is yes, why is there a difference between the two values?

- Vary the tank level between 5 and 95 per cent. How does this affect the process dead time?

An important indication of the effect of dead time on a process is the dead time to time constant ratio (t_{DT}/τ). If this ratio is less than 0.3, then the dead time has little or no effect on the process response. However, if the ratio is greater than 0.3, then the process becomes dead-time-dominated and thus is virtually uncontrollable.

- Repeat the step response testing done above to determine the time constant of the new system and use these results to calculate the dead time/time constant ratio. You will need to stop the sine-wave feed-temperature input and use a constant feed temperature. Record your results in Table W3.5.

Table W3.5 Dead time/time constant ratio

Tank level	Dead time (min)	Time constant (min)	Dead time/time constant
5			
50			
95			

- To test the hypothesis that dead time has no effect on the open-loop process attenuation for a capacity-dominated process, again perform a frequency response test using the sine wave input. Run the 40 min (frequency: 0.025 min^{-1}), 25°C amplitude disturbance through the process of Figure W3.5 with the tank level set to 50 per cent. Has the attenuation changed from the value you calculated earlier?

Present your findings on diskette in a short report using MS-Word. Also include on the diskette a copy of the HYSYS files which you used to generate your findings.

Workshop 4 Feedback control

Nothing comes from doing nothing.
Shakespeare

Introduction

Prior to attempting this workshop, you should review Chapters 3 and 4 in the book.

The previous workshop introduced the concepts of capacitance and attenuation. These are 'natural' characteristics of a system, as are dead time and the process time constant. Now that we have a basic understanding of the way processes behave, we can apply this knowledge to control the process response.

Once the process personality is understood, we can manipulate process flows to maintain a desired variable at constant conditions, which are called set points. This is known as feedback control, where the value of a variable is 'fed back' to a controller which manipulates another variable according to the difference between the controlled variable and its set point.

Key learning objectives

1 Feedback control is easiest and most successful for low-capacity processes without dead time.

2 Dead time reduces the ultimate gain of a process.

3 A large time constant decreases the responsiveness of a process and reduces the achievable control performance.

4 Proportional-only control suffers from offset, which can be eliminated through integral action.

5 Derivative action can only be used where there is no significant process noise and relatively little dead time.

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6 Buffer tanks and surge drums can help smooth out changes in flow and, thereby, isolate equipment from upstream disturbances. This can only result from loose level tuning, where the primary interest is flow smoothing not level control.

7 Proportional-only level control with a controller gain K_c of 2.0 is generally sufficient. When the manipulated variable is the outlet flow, this implies that the valve is fully shut at 25 per cent level and fully open at 75 per cent level. $K_c < 1.0$ won't hold the level between 0 and 100 per cent.

8 For averaging level control:

- attenuation decreases as the gain increases;
- adding integral action to an integrating process (level control) can become a disturbance generator rather than a disturbance smoother if not properly tuned;
- level loop tuning is always dependent on the system characteristics;
- if the hold-up is too small to get good flow smoothing, then reduce K_c and add integral action to ensure that the level stays in the tank (K_c $\mathrm{T_i} = 4.0$);
- if the hold-up time is long, then you do not need any integral action;
- if the level is cycling, then increase K_c and decrease T_i (this is the opposite of other loops).

Tasks

1 Low capacity, no dead time process

In the previous workshop you should have built the system shown in Figure W4.1. Check that you still have the Wilson property package specified. The only components required are water and methanol. The inlet temperature should be 30°C, and the heater outlet should be fixed at 70°C. Initially, we wish to analyse a process without dead time; so, you will need to delete the PFR that you added in the previous workshop. Retain the strip charts that you set up previously. If they have been deleted, rebuild them to contain the following variables: feed temperature, tank outlet temperature, steam heat flow, and feed molar flow. Select suitable ranges for each variable.

The results from the previous workshop indicated that processes with low capacitance and relatively long disturbance period have the lowest attenuation and are in most need of process control. Hence, for this portion of the workshop you will need to use a low-capacity process, so adjust the tank level accordingly. Your simulation should still

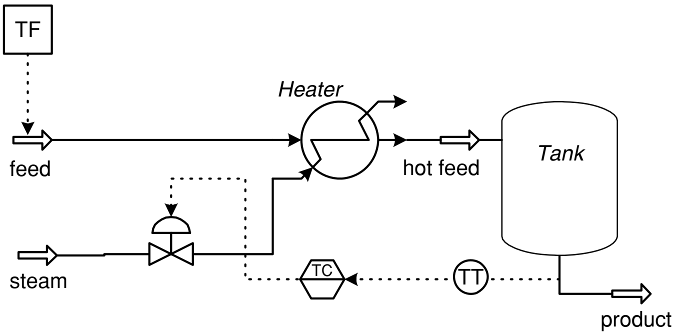

Figure W4.1 Low capacity, no dead time process

contain a transfer block operation, which adds noise to the system. Set the feed water temperature disturbance period to 30 min and its amplitude to 25°C.

Now add a controller for the tank outlet temperature. The controller should manipulate the steam rate to the heater between 0 and 1×10^6 kJ h^{-1} (direct Q). The PV range should be 0 to 100°C, and the set point should be 70°C to match the steady-state conditions. Set the controller gain K_c to 1.0, but leave the integral time and derivative time blank. Make sure that you correctly specify whether your controllers are direct acting or reverse acting so that the controller will open/close the valve when required, and not vice versa. Finally, set the controller to automatic.

The previous workshop demonstrated that capacity-dominated processes have significant disturbance rejection (attenuation) properties without requiring any form of process control. This is called open-loop attenuation. Controllers can usually increase the attenuation of process systems. When operated in 'automatic', the additional attenuation is called closed-loop attenuation. When in 'manual', the system behaves as it would without the controller present.

- Vary the gain to achieve the maximum closed-loop disturbance attenuation. How effective is the controller in rejecting disturbance not already rejected by the natural attenuation of the process?

- What happens when the controller gain gets very high? Is there a limit to how much you can increase the gain?

2 Process with dead time

Capacity-dominated processes are relatively easy to control. However, the presence of dead time makes the control problem more difficult. We can demonstrate this by adding a PFR to the system shown in Figure W4.1 in order to simulate dead time (Figure W4.2). The PFR should have a length of 2.0 m, a total volume of 0.5 m^3 (dead

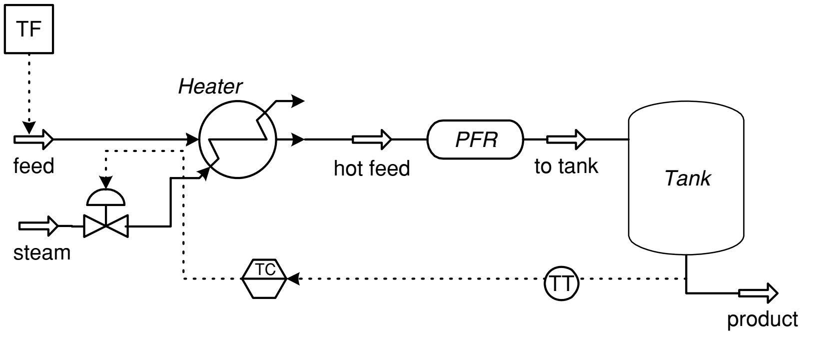

Figure W4.2 High-capacity process with dead time

time. = 10 min) and a pressure drop of 0 kPa. Remember that you want to work with a capacity-dominated process, so ensure that the tank level is set accordingly.

- What is the maximum attenuation for the process with dead time?
- Is there an optimum/maximum gain that maximizes process attenuation?
- Fix the controller gain at $K_c \approx 10$ and calculate attenuation for dead times of 2, 5, 10, and 20 min. Vary the size of the PFR to change the amount of dead time in the system. Record the results in Table W4.1.

3 Proportional-only control

We have found that feedback control can provide good attenuation of process disturbances provided that the dead time is not too great. The ability to provide attenuation to a process is sometimes called disturbance rejection. However, disturbance rejection is only one of the requirements of an effective controller. The other main requirement is that we should be able to change the set point whenever we want to and have the controller manipulate the process so that the controlled variable continues to match the set point. This is sometimes referred to as controller performance. These two parameters,

Table W4.1 Process attenuation with dead time

Dead time (min)	Product ΔT (°C)	Attenuation
2		
5		
10		
20		

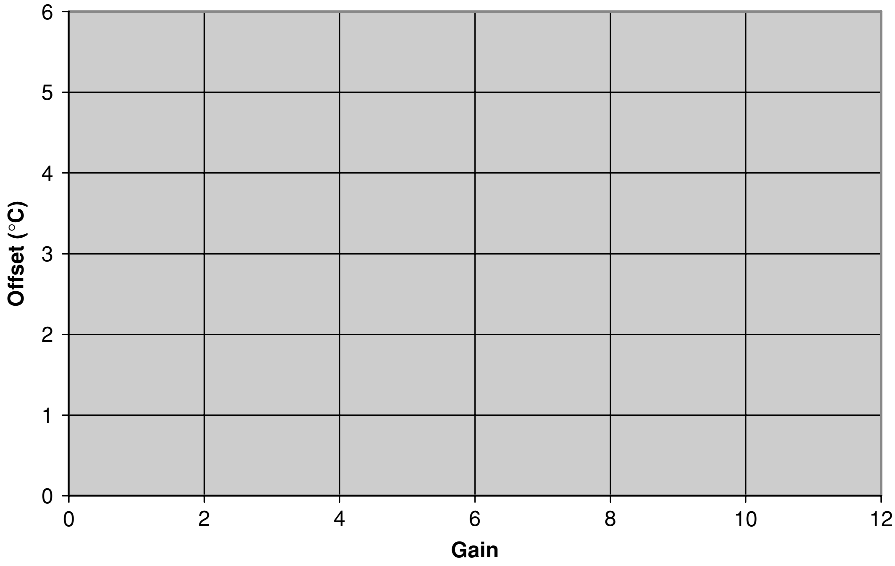

Figure W4.3 Proportional-only controller offset

disturbance rejection and controller performance, are used to assess the effectiveness of a controller.

Eliminate dead time from the system by deleting the PFR or setting its volume to a very low value, and pause the feed temperature disturbance to remove the noise from the process. Starting with a gain of 1.0, interactively change the tank outlet temperature set point to 80°C.

- Where does the tank outlet temperature stabilize? The difference between this value and the set point is called offset.

- Plot the relationship between offset and the controller gain on Figure W4.3.

- Can you explain the relationship shown on Figure W4.3 in terms of the proportional-only controller equation given in Equation W4.1?

$$\text{Output} = \text{Gain} \times (\text{SP} - \text{PV}) + \text{Bias} \tag{W4.1}$$

4 PI and PID control

The feedback controller we have employed to this point has only contained one term, i.e. controller gain. As suggested above, this type of controller is called a proportional-only controller, which suffers from the problem of offset. Offset can be reduced with high

Table W4.2 PI controller optimization

Test	Gain	Integral time (min)	Offset	Time to steady state (min)

values of gain, but sometimes this makes the controller unstable, particularly when there is dead time in the system. We can eliminate offset by introducing another term to the controller equation, i.e. integral time. The control equation for a proportional-integral (PI) controller is

$$\text{Output} = \text{Gain} \times \left[(\text{SP} - \text{PV}) + \frac{1}{T_\text{i}} \int_0^T (\text{SP} - \text{PV})\, \text{d}t \right] \qquad \text{(W4.2)}$$

- How does PI control eliminate offset? Use Equation W4.2 to help explain.

  Add integral time to your controller, starting with $T_\text{i} = 1.0$, and check that it eliminates offset for set-point changes, both with and without dead time in the system. Interactively change the feed rate to determine how effective the controller is at disturbance rejection, i.e. step response testing.

- Summarize your results for PI control in Table W4.2. Record the details of each type of step test performed, i.e. 30–40 kmol h^{-1}, under the 'Test' column.

Integral action can be slow, since it relies on the integral of the error being large, where the error is the difference between the set point and the process variable. Proportional action usually provides the 'muscle' for the controller. However, too much proportional action creates instability. In some circumstances, PI controllers are not sufficiently fast, making a third controller action necessary. This term is called derivative time and can sometimes be introduced to speed up the response time of the controller. The control equation for a proportional–integral–derivative (PID) controller is

$$\text{Output} = \text{Gain} \times \left((\text{SP} - \text{PV}) + \frac{1}{T_\text{i}} \int_0^T (\text{SP} - \text{PV})\, \text{d}t + T_\text{d} \frac{\text{d(SP} - \text{PV)}}{\text{d}t} \right) \qquad \text{(W4.3)}$$

Derivative time increases the controller response when the controlled variable is moving away from its set point most quickly, i.e. straight after a disturbance has affected

Table W4.3 PID controller optimization

Test	Gain	Integral time (min)	Derivative time (min)	Offset	Time to steady state (min)

the system. Apart from increasing the responsiveness of the controller, derivative action also reduces oscillation. Derivative action can be very effective under some circumstances but very damaging under others. For example, if the system is essentially stable but there is a small amount of process noise (usually very high frequency), then the derivative action will interpret the noise as being the start of a large disturbance and will make large changes in the manipulated variable which are clearly not required.

Add derivative action to your system, starting with $T_d = 1.0$, to determine whether or not it improves the controller effectiveness for this example.

- Optimize controller performance by varying the three controller parameters. Consider the responsiveness to process disturbances and the ability to track a set point. Record your results in Table W4.3. Record the type of step change test performed under the 'Test' column.

5 Averaging level control

Surge drums and intermediate product tanks are critical parts of any process system. Their principal purpose is to provide hold-up and capacitance to smooth out flow disturbances so that they do not carry through to downstream process units. This function must be recognized, and it is frequently overlooked in many operating plants. A consequence of this function is that the level in surge drums and intermediate tanks should **not** be tightly controlled. Tight level control will transmit flow disturbances to downstream units and negate the effectiveness of the surge volume. The level controller must only control the level between the low limit (when the tank/drum approaches empty and thereby risks damaging the outlet pump) and the high limit (when the tank overflows). Intentional loose level control is called averaging level control.

One exception to the rule of averaging level control for surge drums is in the case of distillation column hold-ups. Averaging level control should not be used to control the reflux drum level or the reboiler sump level. Tight level control is required for these vessels to maintain the integrity of the column material balance so that changes in the reflux rate and reboiler duty will have the desired effect on product compositions and yields without introducing additional lag to the system.

Table W4.4 Averaging level control

Disturbance period (min)	$K_c = 1.0$ $T_i = 50$ min	$K_c = 2.0$ $T_i = 25$ min	$K_c = 0.5$ $T_i = 100$ min
5			
10			
20			
30			

In order to understand better how averaging level control works, build a simple system consisting only of a 2 m^3 surge drum. The feed to the surge drum should be 250 kmol h^{-1} of water at 25°C and 100 kPa. Add a level controller and enter a set point of 50 per cent. Finally, add a feed disturbance using a transfer function unit operation set up to vary the feed rate sinusoidally with an amplitude of 25 kmol h^{-1} and a period of 4 min. The control valve range should be 0–500 kmol h^{-1}. Remember that the amplitude is entered as a per centage of the PV span for the transfer function operation.

- Test the following combinations of PI control for the surge drum level controller for the range of disturbance periods given in Table W4.4:

  1 $K_c = 1.0$, $T_i = 50$ min

  2 $K_c = 2.0$, $T_i = 25$ min

  3 $K_c = 0.5$, $T_i = 100$ min

- Which combination of gain and level control provides the best disturbance attenuation?

- Are there any problems with using a very low gain?

- Are there any problems with using a proportional-only level controller?

Present your findings on diskette in a short report using MS-Word. Also include on the disk a copy of the HYSYS files which you used to generate your findings.

Workshop 5 Controller tuning for capacity and dead time processes

A little experience often upsets a lot of theory.
Samuel Parks Cadman

Introduction

Prior to attempting this workshop, you should review Chapter 5 in the book.

This workshop will illustrate that HYSYS may be used to determine the appropriate parameters for a PI controller that is controlling a capacitive process with significant dead time. You will learn that controller tuning is determined by the desired load or set-point response as well as the type of process and the values of the process parameters, which include process gain, time constant, and dead time. A review of the three tuning techniques that are used in this workshop is provided below.

Process reaction curve tuning technique

In the process reaction curve method, a process reaction curve is generated in response to a disturbance. This process curve is then used to calculate the controller gain, integral time and derivative time. The method is performed in open loop so that no control action occurs and the process response can be isolated.

To generate a process reaction curve, the process is allowed to reach steady state or as close to steady state as possible. Then, in open loop, so that there is no control action, a small step disturbance is introduced and the reaction of the process variable is recorded. Figure W5.1 shows a typical process reaction curve for the process variable (PV) generated using the above method for a generic self-regulating process. The term self-regulating refers to a process where the controlled variable eventually returns to a stable value or levels out without external intervention.

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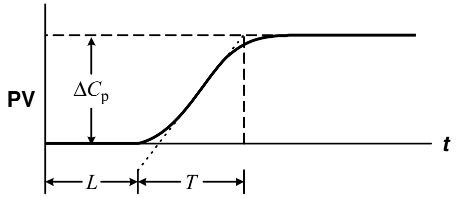

Figure W5.1 Process reaction curve

The process parameters that may be obtained from this process reaction curve are as follows:

L	(min) lag time
T	(min) time constant estimate
P	(%) initial step disturbance
$\Delta C_p(\%)$	change in PV in response to step disturbance, (change in PV)/(PV span) $\times 100$
$N = \dfrac{\Delta C_p}{T}$	(% min^{-1}) reaction rate
$R = \dfrac{L}{T} = \dfrac{NL}{\Delta C_p}$	(dimensionless) lag ratio

The Ziegler–Nichols process reaction curve tuning method for a PI controller is as follows:

1 Determine a *reasonable* value for the step valve change P. This value is arbitrarily chosen, but typically 5 per cent is reasonable.

2 With the controller in *manual* mode, manually move the valve 'P' per cent.

3 Wait until the PV lines out to the new steady-state value.

4 Determine N and R from the process reaction curve.

5 Perform the following calculations:

controller gain $K_c = 0.9P/NL$

controller integral time $T_i = 3.33\,L$

6 Implement these recommendations for the controller settings in the controller.

7 Close the control loop by placing the controller in *automatic* mode.

8 Test thoroughly, fine-tuning the parameters to obtain the QDR.

Auto-tune variation tuning technique

The auto-tune variation or ATV technique of Åström is one of a number of techniques used to determine two important system constants called the ultimate period and the ultimate gain. Tuning values for proportional, integral and derivative controller parameters may be determined from these two constants. All methods for determining the ultimate period and ultimate gain involve disturbing the system and using the disturbance response to extract the values of these constants.

In the case of the ATV technique, a small limit-cycle disturbance is set up between the manipulated variable (controller output) and the controlled variable (process variable). Figure W5.2 shows the typical ATV response plot with critical parameters defined. It is important to note that the ATV technique is applicable only to processes with dead time. The ultimate period will just equal the sampling period if the dead time is not significant.

The general ATV tuning method for a PI controller is as follows:

1 Determine a *reasonable* value for the valve change h. This value is arbitrarily chosen, but typically 0.05 is reasonable, i.e. 5 per cent.

2 With the controller in the off position, manually move the valve '$+h$' units.

3 Wait until the process variable PV starts to move and then move the valve '$-2h$' units.

4 When the process variable crosses the set point, move the valve '$+2h$' units.

5 Repeat until a limit cycle is established, as illustrated in Figure W5.2.

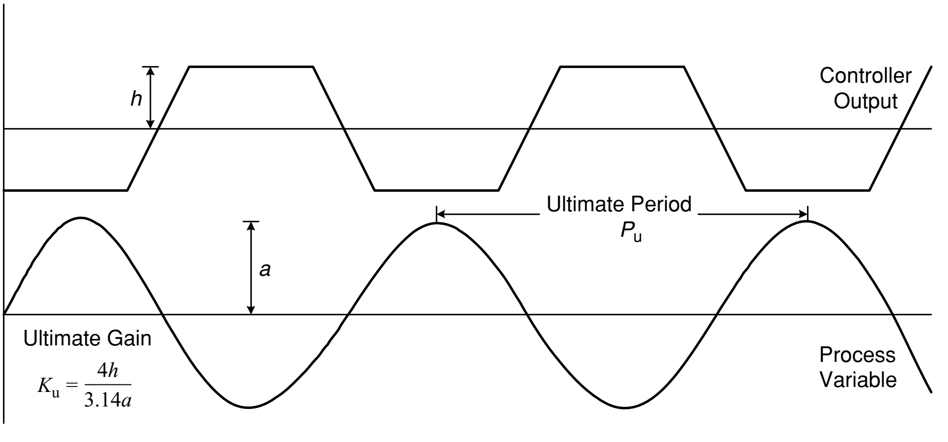

Figure W5.2 ATV critical parameters

6 Record the value of the amplitude a by picking it off the response graph.

7 Perform the following calculations:

$$\begin{aligned} \text{ultimate period } P_u &= \text{ period taken from the limit cycle} \\ \text{ultimate gain } K_u &= 4h/3.14a \\ \text{controller gain } K_c &= K_u/3.2 \\ \text{controller integral time } T_i &= 2.2P_u. \end{aligned}$$

Ziegler–Nichols closed-loop tuning technique

The closed-loop technique of Ziegler and Nichols is another technique that is commonly used to determine the two important system constants, i.e. ultimate period and ultimate gain. Historically speaking, it was one of the first tuning techniques to be widely adopted.

In Ziegler–Nichols closed-loop tuning, as for the ATV technique, tuning values for proportional, integral and derivative controller parameters may be determined from the ultimate period and ultimate gain. However, Ziegler–Nichols closed-loop tuning is done by disturbing the closed-loop system and using the disturbance response to extract the values of these constants.

The Ziegler–Nichols closed-loop tuning method for a PI controller is as follows:

1 Attach a proportional-only controller with a low gain (no integral or derivative action).

2 Place the controller in automatic.

3 Increase proportional gain until a constant-amplitude limit cycle occurs.

4 Perform the following calculations:

$$\begin{aligned} \text{ultimate period } P_u &= \text{ period taken from limit cycle} \\ \text{ultimate gain } K_u &= \text{ controller gain that produces the limit cycle} \\ \text{controller gain } K_c &= K_u/2.2 \\ \text{controller integral time } T_i &= P_u/1.2. \end{aligned}$$

Key learning objectives

1 Controller tuning is determined by the desired controller response.

2 Controller tuning is determined by the type of process.

3 Controller tuning is affected by the value of the process gain.

4 Controller tuning is affected by the value of the time constant.

5 Controller tuning is affected by the value of the dead time.

6 The ATV tuning technique is a powerful method for many loops.

7 The Ziegler–Nichols closed-loop technique is also useful, but more aggressive than ATV.

8 The Ziegler–Nichols process reaction curve technique is also useful, as it provides estimates for the key process parameters.

9 HYSYS can be used to find appropriate tuning parameters for a PI controller

Tasks

1 Tuning Controllers

The process used for this workshop is shown in Figure W5.3. A 50/50 feed mixture of water and ethanol ($T = 5°\mathrm{C}$, $P = 200$ kPa, $F = 100$ kmol h^{-1}) is heated in a steam heater to approximately 70°C. The hot stream passes through a dead-time leg before being stored in a tank for future use. Use a PFR unit operation to simulate the dead time with a volume of 3 m^3 and a length of 2 m. This was the process you worked on in the latter part of Workshop 3.

Set the tank level to 50 per cent with no incoming disturbances. With the temperature controller in manual, adjust the steam valve to get a tank temperature of approximately 70°C. Bring up the temperature controller faceplate.

First use the Ziegler–Nichols process reaction curve technique to determine the controller settings at 50 per cent tank level. Determine the controller settings at two more tank levels (5 and 95 per cent).

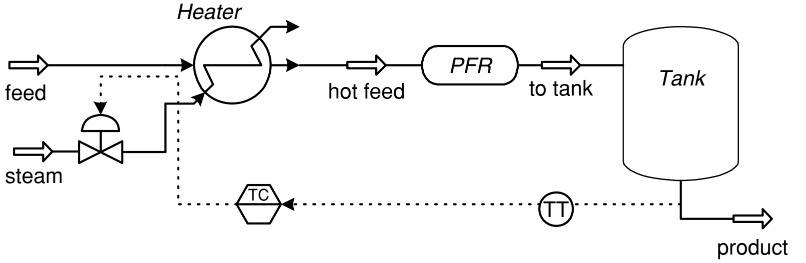

Figure W5.3 Illustrative capacity plus dead time process

Second, use the ATV technique to determine the controller settings as follows. Set the mode to auto-tune. The controller will bring the process into a limit cycle.

- Determine the period of this limit cycle in minutes. Use this limit cycle to determine the amplitude of the temperature cycle of the stream exiting the tank and make this dimensionless by dividing by the temperature transmitter span.

- Now determine the fractional amplitude of the controller output h.

- Calculate the ultimate gain and use this with the ultimate period to compute the controller settings.

- Determine the controller settings at two more tank levels (5 and 95 per cent).

Now use the Ziegler–Nichols closed-loop tuning technique to determine the controller settings at the three tank levels.

- Compare the results of using both the ATV and Ziegler–Nichols tuning techniques.

2 Controller contributions to attenuation

We have seen in Workshop 3 that the process itself is able to attenuate with no control, i.e. open loop. We have just tuned our feedback controller for various levels of capacitance and can now determine what the process plus control (closed loop) is able to attenuate. By subtracting the open loop attenuation from the total attenuation we can determine what the controller itself contributes to the overall process attenuation.

- Determine the total closed-loop attenuation of the tank operating at the 50 per cent level for sinusoidal disturbances of periods 10, 20, 30, 40, and 100 min with an amplitude of 25°C.

- Compute the controller contribution to attenuation for these disturbances.

- At the 5 per cent level determine the controller attenuation for sinusoidal disturbances of periods 5, 10, 20, 30, and 50 min and amplitude 25°C.

- At the 95 per cent level determine the controller attenuation for sinusoidal disturbances of periods 10, 20, 30, 40, and 100 min and amplitude 25°C.

- Plot attenuation versus the logarithm of the disturbance period. Compare the curves using their dead time to time constant ratios that you calculated in Workshop 3.

Present your findings on diskette in a short report using MS-Word. Also include on the disk a copy of the HYSYS files which you used to generate your findings.

Workshop 6 Topics in advanced control

Theory without experience is sterile, practice without theory is blind.
George Jay Anyon

Introduction

Prior to attempting this workshop, you should review Chapters 6 and 7 in the book.

This workshop will show how the response of feedback control loops can be improved through the use of other control methods. These other methods include measuring common disturbances and taking action before they affect the controlled variable (feedforward control) and using a faster responding loop to decrease the response time of a system with a large time constant (cascade control). You will determine what conditions are necessary for feedforward or cascade control to be useful and identify which parameters reduce the effectiveness of these control methods.

Key learning objectives

Feedforward control

1 Feedforward controllers can respond faster than feedback controllers can, since they react to process disturbances immediately without waiting for them to affect the process.

2 Feedforward control can only compensate for disturbances that are measured. Its effectiveness is reduced if unmeasured disturbances are significant.

3 Feedforward control is less effective for nonlinear processes, where nonlinearities exist between the disturbance measurement and controlled variable.

Cascade control

4 Cascade control can significantly improve the control performance if a secondary variable can be found in the system that directly affects the primary loop and is faster responding than the primary loop.

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5 The inner control loop helps to reject disturbances to the primary control variable.

6 The ultimate period of the inner (slave) loop should be at least four times smaller than the outer (master) loop for cascade control to be effective.

7 The most frequently used slave loop is a flow loop, but other types of fast-responding loop, such as pressure loops, can also be used.

Ratio control

8 Ratio control is a type of simple feedforward control that is most effective for low-frequency disturbances.

Tasks

1 Basic process configuration

Build the system shown in Figure W6.1. The feed is pure water with a temperature of 20°C, atmospheric pressure, and a flow rate of 1.5 $m^3\,h^{-1}$. The outlet temperature of the heater should be set to 55°C with steam as the heating medium. Assume that the pressure drop is negligible, i.e. set equal to zero. Incorporate dead time into the process by adding a PFR with a length of 1.0 m, a total volume of 0.2 m^3 and a pressure drop of 0 kPa. Finally, add a tank with a volume of 1.2 m^3 and set the liquid level set point at 50 per cent. Alternatively, you might be able to modify the simulation you used for Workshop 4.

Use strip charts to view your results, monitoring the following variables: feed temperature, product temperature, steam heat flow, and product molar flow. Select suitable ranges for each variable and iconize the strip-chart view for later use.

Add two controllers to the system to set up feedback control of the process:

1 The first controller should manipulate the steam rate to the heater between 0 and 5×10^5 kJ h^{-1} (direct Q) to control the product temperature between 0 and 100°C. The set point should be 55°C initially, with tuning constants of $K_c = 0.5$ and $T_i = 10$.

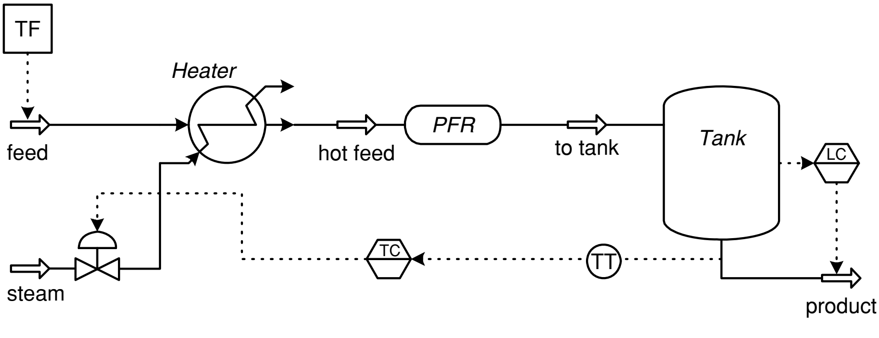

Figure W6.1 Simple heating system

2 The second controller should manipulate the product flow from the tank between 0 and 200 kmol h^{-1} to control the tank level between 0 and 100 per cent. The set point should be 50 per cent with tuning constants of $K_c = 10$ and $T_i = 10$.

Make sure that you correctly specify whether your controllers are direct acting or reverse acting. Set both controllers to automatic.

Two types of feed disturbance will be tested in this workshop: step disturbances and sinusoidal disturbances. Make step disturbances in the feed by changing the feed temperature by 10°C. Add sinusoidal feed disturbances to the system using the Transfer Block unit operation. The transfer block PV target is the feed temperature. Create sine-wave noise with an amplitude of 10°C and a period of 10 min. Remember that the amplitude is entered as a per centage of the PV span, i.e. $PV_{max} - PV_{min}$. The feed temperature should still oscillate around a mean of 20°C.

2 Determine base-line control performance

First make step disturbances to the feed temperature; observe the responses and record values of suitable metrics, such as suggested in Chapter 5.

The heater–tank system, shown in Figure W6.1, has a large capacitance, which should provide good attenuation of process disturbances and help reject high-frequency process noise. However, it will make the system slow to respond to set-point changes or permanent disturbances, i.e. feed rate changes. The significant dead time in the system will compound any control problems and make it difficult to achieve tight control of the system when only FBC is used.

- Vary the period of the disturbance to the feed temperature and fill in Table W6.1 to demonstrate this characteristic of the system.

- Should the magnitude of the feed temperature disturbance affect the attenuation?

- From your results, identify any deficiencies of the feedback control system simulated above. If necessary, vary the temperature controller tuning constants to try to improve the performance of the control loop.

Table W6.1 Base-line control performance

Disturbance period (min)	Frequency (min^{-1})	Product ΔT (°C)	Attenuation
5			
10			
20			
40			
60			

Hint: How long does it take the controller to respond to a change in the feed temperature? Can the warm water temperature be stabilized by manipulating the tuning constants? How does the controller respond to changes in the feed rate, i.e. step response testing?

3 Feedforward control

Feedforward control can be used to combat the control problems associated with processes containing significant dead time. This is achieved by measuring process disturbances and compensating for them before they affect the controlled variables. Ideal feedforward control is realized if pre-emptive control action is taken to cancel out the effect of measured disturbances completely before they enter the process. Sometimes the ideal feedforward controller is not realizable, because disturbances affect the system more quickly than the manipulated variable. However, feedforward control can still be useful in these scenarios when teamed with feedback control, because the feedforward control reduces the duty on the master controller and improves the overall system response. Clearly, no action can be taken if the disturbances are not sensed or measured.

Build a feedforward controller for the heater–tank system, using the built-in spreadsheet function in HYSYS, to compensate for changes in the feed temperature before they become apparent in the warm-water temperature. This feedforward control will be combined with the feedback control to see whether process response can be improved.

Set up the following titles in cells A1–A6:

A1 Actual Feed Temp

A2 Nominal Feed Temp

A3 Temp Difference

A4 Process Gain 2

A5 Steam Valve Span

A6 Process Gain 1

A7 Feedforward Duty

Complete the spreadsheet as follows:

B1 Drag and drop the feed temperature from the process feed stream.

B2 Input the nominal feed temperature, which is equivalent to the feed temperature set point.

[*Hint: refer to the Transfer Function operation.*]

B3 + B2 − B1

B4 Input the value of the process gain between the product temperature and the feed temperature.

[*Hint: how much does the product temperature rise for a 1°C step increase in the feed temperature?*]

B5 Drag and drop the span of the heater duty valve.

B6 Input the value of the process gain between the product temperature and the heater duty.

[*Hint: how much does the warm-water temperature rise if the heater duty changes from 0 to 100 per cent?*]

B7 The feedforward duty can be calculated from Equation W6.1. Incorporate this equation into the spreadsheet.

$$\Delta Q = \frac{\text{Nominal Feed Temp} - \text{Actual Feed Temp}}{\text{Process Gain 1}} \times \frac{\text{Steam Valve Span}}{\text{Process Gain 2}} \qquad \text{(W6.1)}$$

ΔQ represents the changes in heater duty required to produce a 1°C change in the product temperature. Equation W6.1 is not necessarily exact at all values of the feed temperature because of the process nonlinearity. An exact expression is not necessary for successful feedforward control; even if the calculated duty is incorrect by 50 per cent, the controller will still perform better than with no feedforward action.

Create a new energy stream named ‘FF Duty’. Export the result in cell B7 to the new energy stream named ‘FF Duty’. Add a mixer to the flowsheet and combine the new ‘FF Duty’ with the original heater duty stream, called ‘master signal’. Attach the output to the heater duty energy stream, called ‘steam’. Your process should be similar to the one shown in Figure W6.2.

- Test the feedforward controller for the same range of feed temperature disturbances that you analysed for FBC in the previous task. Record the sinusoidal disturbance results in Table W6.2.
- How effective is the feedforward controller?
- What are its major deficiencies?

[*Hint: test the effectiveness of the feedforward controller for changes in the feed rate.*]

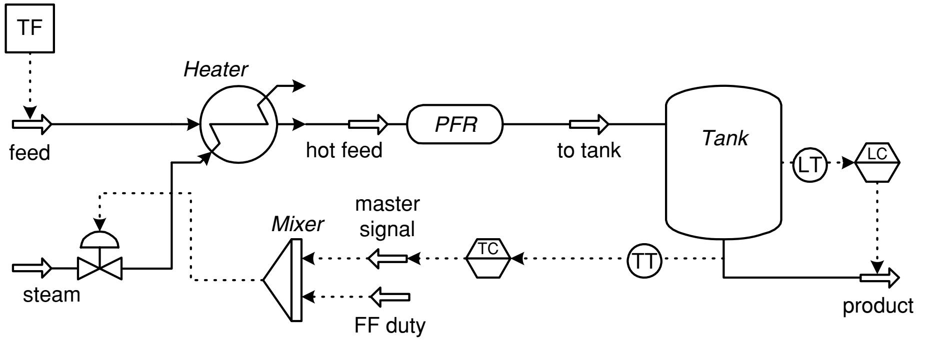

Figure W6.2 Feedforward control system

- Briefly comment on any implementation issues that might be relevant with feedforward control.

  [*Hint: how is the feedforward gain calculated? How can dynamics be incorporated into the feedforward controller? How important is tuning of the feedforward controller?*]

4 Cascade control

Cascade control is an alternative way to manage processes that contain large time constants and/or significant dead time. It is not necessary to sense or measure disturbances, but a secondary variable must exist that directly affects the primary (master) loop and is faster responding than the primary loop. The secondary variable is usually, but not necessarily always, a flow or a pressure that is directly controlled via a control valve. Normally, this secondary (slave) controller is a flow controller, a pressure controller, or a fast-responding temperature controller. The time constant of the slave loop should be less than 25 per cent of the time constant of the master loop for cascade control to be effective ($\tau_M \geq 4\tau_S$). Also, the secondary loop should contain little or no dead time. This allows the secondary variable to be controlled tightly, which provides attenuation for the primary loop.

Build a cascade controller for the heater–tank system using an inner loop that manipulates the steam rate based on the heater outlet (‘hot feed’) temperature. The master controller (i.e. tank temperature) should provide the set point for the slave loop, and

Table W6.2 Feedforward control performance

Disturbance period (min)	Frequency (min^{-1})	Product ΔT (°C)	Attenuation
5			
10			
20			
40			
60			

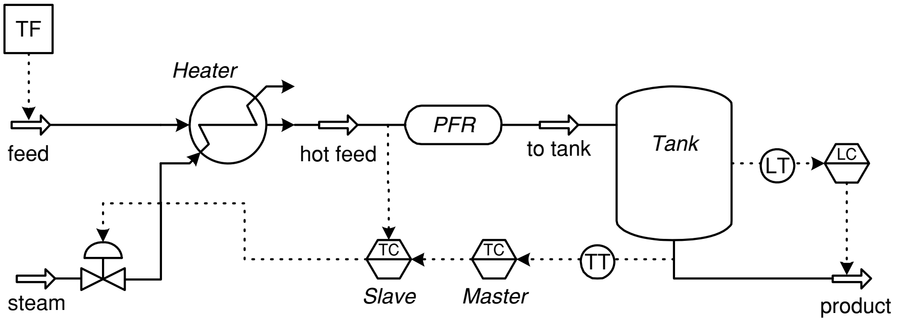

Figure W6.3 Cascade control system

the slave controller should manipulate the steam rate directly. The process is shown in Figure W6.3.

To implement cascade control into your existing simulation, delete the feedforward controller but retain the tank level controller and the original tank temperature controller. The tank temperature controller will now be the master controller for the cascade loop.

Add another controller unit operation to the PFD. Connect the PV point to the 'hot feed' temperature. Connect the OP point to the heater energy stream ('steam') and specify 'Direct Q' between 0 and 5×10^5 kJ h^{-1}. Connect the SP point (cascaded set-point source) to the PV point of the master controller. The slave loop should be tuned tightly; specify a gain of 10 and an integral time of 10 min. The master loop can be tuned more loosely; specify a gain of 1.0 and an integral time of 10 min. Note that the gain of the master loop is not numerically comparable to the gain of the temperature loop from the previous simulation (without cascade control) because a different variable is being manipulated in the two cases. Your process should now be similar to the one shown in Figure W6.3.

- Test the cascade controller for the same range of feed temperature disturbances that you analysed for the previous two systems, which contained FBC only and feedforward control. Record the results for the sinusoidal disturbances in Table W6.3.

- Try varying the tuning constants for both the slave loop and the master loop. Which combination(s) of tuning constants work best?

Table W6.3 Cascade control performance

Disturbance period (min)	Frequency (min^{-1})	Product ΔT (°C)	Attenuation
5			
10			
20			
40			
60			

- What comments can you make about the slave and master controller settings? How sensitive is the overall control performance to the slave loop tuning?

- Overall, how effective is the cascade controller?

  [*Hint: how does the controller respond to changes in the feed rate? How does the controller respond to changes in the master controller set point? Does the duty control valve open and shut excessively, i.e. is there too much control action?*]

- What are the major advantages and shortcomings of cascade control?

- How does cascade control compare with feedforward control?

5 Ratio control

Ratio control is a simple form of feedforward control that is commonly employed in controlling reactor feed compositions and in blending operations. It is also used to control the fuel-to-air ratio in heaters and boilers and to control the reflux ratio in distillation columns. The flow rate of one stream is used to provide the set point for another stream so that that ratio of the two streams is kept constant even if the flow of the first stream varies. Alternatively, the actual ratio between two flows can be used as the input to a controller.

Build a new system consisting of two streams, two tanks, and a mixer using the Wilson thermodynamic package. Pick any two components that are liquid phase at ambient temperatures. The first stream should be pure component 'A' at 25°C and 100 kPa. The second stream should be pure component 'B' at the same temperature and pressure. Set the flow of the first stream to 400 kg h^{-1} and the second stream to 100 kg h^{-1}. These flows are consistent with the desired ratio of 4:1 between components A and B. The tanks are used to simulate dead time in the system, so choose relatively small volumes for the tanks and locate them in series with the first stream. Both tanks should be on level control rather than liquid flow control. Simulate process noise with a sine-wave input to the first stream using an amplitude of 50 kg h^{-1} and a period of 10 min. The system should resemble the one shown in Figure W6.4.

Incorporate ratio control via a spreadsheet. Import the flow on the first stream into cell A1. Put the ratio of 0.25 in cell A2 and add a formula to give the flow of the second stream in cell A3. Export the result of cell A3 to the flow of the second stream.

Run the simulation in dynamic mode with several disturbances. Watch how the combined flow rate and concentration changes with different conditions. You may need to reduce the integrator step size to see the effects of very high-frequency disturbances (period <5 min).

- How effective is the ratio controller at filtering out low-frequency noise?

- How effective is the ratio controller at filtering out high-frequency noise?

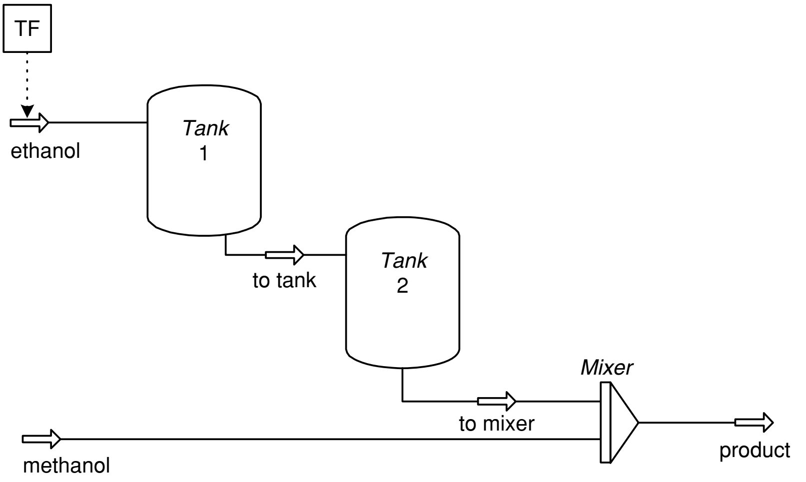

Figure W6.4 Ratio control system

- What are some of the advantages and limitations of ratio control?
- What is the significance of the dead time in the system?

Hint: How would your answers change to the above questions if the dead time was the same for both streams?

Present your findings on diskette in a short report using MS-Word. Also include on the disk a copy of the HYSYS files which you used to generate your findings.

Workshop 7 Distillation control

Have confidence that if you have done a little thing well, you can do a bigger thing well, too.

Joseph Storey

Introduction

Prior to attempting this workshop, you should review Chapter 8 in the book.

Distillation is one of the most important unit operations in chemical engineering. It forms the basis of many processes and is an essential part of many others. It presents a more difficult control problem then with many other unit operations, as at least five variables need to be controlled simultaneously and there are at least five variables available for manipulation. Thus, a distillation column provides an example of a multiple-input–multiple-output control problem. It is critical that variable pairing is done appropriately between controlled and manipulated variables. The overall control problem can usually be reduced to a 2×2 composition control problem since the inventory and pressure loops frequently do not interact with the composition loops. This workshop will highlight some fundamental rules of distillation control and show how a basic distillation control scheme can be selected.

Key learning objectives

1 There are five degrees of freedom present in a simple distillation column with no side draws and a total condenser. The degrees of freedom increase by one with each side draw and for a partial condenser, which has two overhead products.

2 The degrees of freedom equal the number of controlled variables, the number of manipulated variables, and the number of control valves: $\mathrm{DF} = N_{\mathrm{cv}} = N_{\mathrm{mv}} = N_{\mathrm{valves}}$.

3 Steady-state analysis can be used to develop a basic control strategy that provides good sensitivity. Dynamic analysis is required to develop a control scheme that provides good responsiveness.

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4 Vapour and liquid flows have different dead times and response times.

5 Two-point composition control involves the control of both the distillate and bottoms compositions simultaneously. One-point composition control involves operating one of the composition control loops in manual or using a non-composition-related variable as the controlled variable.

6 Distillation control schemes are usually described by the two variables which are manipulated for composition control (i.e. the LV, DV or LB configuration) and may work by manipulating either the mass balance (i.e. the DV and LB configurations) or the energy balance (i.e. the LV configuration).

Tasks

1 Stabiliser configuration

The distillation column shown in Figure W7.1 is typical of a stabilizer, which is found in most refineries. The column is designed to remove volatile components from potential gasoline blendstocks. The feed is usually a mixture of C_3, C_4, and C_5. In this case, the feed contains 5 per cent propane, 40 per cent isobutane, 40 per cent n-butane, and

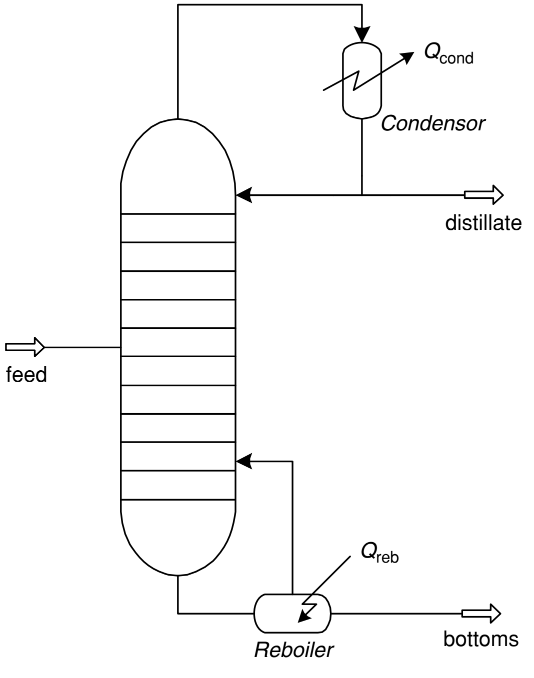

Figure W7.1 Stabilizer

15 per cent isopentane by weight. The total flow rate is 40 000 bbl day^{-1} at 720 kPa and 30°C.

The stabilizer contains 20 trays and a total condenser. Feed enters at tray 10. The normal column overhead pressure is 700 kPa and there is a 20 kPa pressure difference that is evenly distributed between the condenser and the reboiler. Each tray is 2.0 m in diameter with a 0.10 m weir, which is 1.6 m long. The bottoms product specification is 0.01 wt% C_3, and the column is normally operated with a reflux ratio of 1.0.

Use the Peng–Robinson property package and the above information to build a simulation model of the stabilizer. View the results to get a feel for the column in steady state mode.

Hint: The column has two degrees of freedom after the in-built inventory and capacity controls have been considered. These should initially be taken up by a bottoms composition specification and a reflux ratio specification.

2 Control system objectives and design considerations

Ultimately, the importance of process control is seen through increased overall process efficiency allowing the plant engineer to get the most from the process design. This is especially true of distillation control. Most distillation columns are inherently flexible, and a wide range of product yields and compositions can be obtained at varying levels of energy input. A key requirement of any control system is that it relates directly to the process objectives. A control system that does not meet the process objectives or produces results that conflict with the process objectives does not add value to the process.

The first step of control system design is to identify the process objectives, which may not always be obvious. One process objective of the system shown in Figure W7.1 is to control the propane content of the bottoms at 0.01 wt%. However, other process objectives are not clear from the information given above. Controlling the reflux ratio at 1.0 may be desirable for the given conditions, but this is not a key process objective as it does not directly fix any property of the products or the energy consumption.

Some possible process objectives for this system include:

- minimum energy input
- maximum product yield
- minimum isopentane in the overhead product
- maximum recovery of C_4+ components
- overhead temperature to meet utility requirement.

In the stabilizer described above, we will initially use the bottoms composition to set the feed split and the energy input to set the fractionation. As yet, we have not paired these controlled variables to control valves (manipulated variables).

- Define the control objectives for the stabilizer built in the previous task.

Table W7.1 Expected feed variance

	Base case	Minimum flow	Maximum flow
Feed Rate (bbl day^{-1})	40 000	20 000	50 000
C_3 (wt%)	5	1	12
iC_4 (wt%)	40	44	33
nC_4 (wt%)	40	45	37
iC_5 (wt%)	15	10	18

3 Distillation control design via steady-state analysis

The feed to the stabilizer described above is expected to vary in both rate and composition. The preferred control system design should be able to handle the design feed conditions and any extremes that might be expected. The data in Table W7.1 should be used to test any control system design.

Simulations allow essentially all types of variable to be used as controlled variables. However, a control scheme must be implementable in an operating plant. This requires that the variable being controlled can be accurately measured to provide feedback in a control loop. Examples of variables that can be easily measured include flows (especially liquid flows), temperature, and pressures. Some compositions, mainly mass and volumetric fractions, can be measured using analysers, but these instruments are generally expensive and often introduce considerable dead time to processes. Consequently, they are excluded from many control systems.

Preliminary control system design has traditionally been conducted using steady-state data only. Steady-state simulations are performed to gain an understanding of the process and the way it responds to certain changes (disturbances). This information is used to select a candidate control system, which is then either tested with dynamic simulations or immediately implemented in a plant environment in the hope that it provides adequate control with correct tuning. A possible control system design strategy, using only steady-state simulations, is as follows:

1 Select two potential controlled variables (i.e. temperature on tray 3 and the reflux rate). Note the values of these variables from the base case solution (see Task 1) and list them under the ‘Set point’ column in Table W7.2.

2 Use these two potential variables as the new specifications for the column in the steady-state, base case simulation. The value for these two new specifications will be the values noted as the set points in the previous step.

3 Solve the column with the new specifications and record the value of the two manipulated variables under the ‘Prim. MV’ and ‘Sec. MV’ columns. These manipulated

Table W7.2 Control strategy selection using steady-state analysis

				Control objectives					
				Base case		Min. flow		Max. flow	
Primary controlled variable	Set point	Secondary controlled variable	Set point	Prim. MV	Sec. MV	Prim. MV	Sec. MV	Prim. MV	Sec. MV

variables are based on the control objectives, i.e. propane content in the bottoms product and energy input.

4 Repeat step 3 for the expected extremes in flow rate and composition. Again, record the values of the manipulated variables for each case.

5 Repeat starting from step 1 for two new potential controlled variables and/or different set points. Determine which combination of controlled variables produces the best overall control performance for all the feed variances by finding the pair with corresponding set points that meet the control objectives for all variances tested.

6 After the control variables have been established to control the feed-split and column fractionation, set the other variables to control inventory and capacity of the column.

The variable pairings required for inventory and capacity control will often be immediately evident after the composition control variables have been selected. If not, the following guidelines can help:

- Control the pressure with the condenser cooling duty.
- Control levels with an outlet stream from the vessel (i.e. reflux or distillate for the reflux accumulator) or an energy stream that affects the inlet flow to the vessel (i.e. the reboiler duty for the reboiler sump).
- Where more than one stream is available, choose the largest stream for level control.

Use the method listed above to build a control system for the stabilizer given in Figure W7.1 with the following control objectives:

- 0.01 wt% propane in the bottoms

- fixed energy input.

- You should consider the range of feed variance given in Table W7.1 and use only variables that can be easily measured. Record your results in Table W7.2 and provide a diagram of the candidate control system in your report write up.

- Are there any particular advantages or disadvantages associated with your preferred combination of controlled variables?

Once the controlled variables have been chosen, determine what set points will allow the control objectives to be met at all operating conditions (i.e. all three cases from Table W7.1) using the following steps.

1 Modify the column specifications to control the process objectives (i.e. 0.01 wt% propane in the bottoms and fixed heat input) directly.

2 Note the values of the selected controlled variables for all three cases in Table W7.1.

3 Identify which value of each controlled variable is the worst case (most conservative).

4 Re-solve the column with the 'conservative' values of the controlled variables to confirm that the control objectives are met or exceeded for each case, meaning that the concentration of propane in the bottoms is equal to or lower than the specification and the energy input is equal to or lower than the target you have specified.

4 Dynamic responsiveness

Once you have completed the steady-state design, use the results to size all the valves in the system. Add the control loops (two) determined from the steady-state design, with the set points which allow the control objectives to be met at all times. Inventory and pressure control will be available automatically, but you can add your own control loops if desired. Solve the column in dynamic mode. You are free to change set points or create disturbances to the system to examine how the control system performs. A good starting point would be to see how the column responds to step disturbances in the feed between the expected feed cases. You may need to tune the control loops in order to produce an adequate response.

- Does your candidate control scheme provide satisfactory control?

Hint: test the control schemes for a wide range of disturbances, including some outside the bounds of the steady-state design. Are the system dead times and lags dominating the control behaviour? Are both set point changes and disturbances handled adequately?

5 Distillation column control configurations

We noted previously that a simple distillation column with a total condenser normally has five degrees of freedom. Each degree of freedom corresponds to a control valve and a controlled variable. Three of these degrees of freedom must be used to control the inventory and capacity variables, i.e. levels and pressures. The remaining two degrees of freedom are used for composition control. The condenser duty (or a related variable) is usually reserved for pressure control. However, any of the remaining four variables can be used for composition control. The following notation is often used for the four degrees of freedom: L is the liquid flow down the column, reflux rate; V = vapour flow up the column, boil-up or reboiler duty; D = distillate rate; B = bottoms rate.

The relationship between boil-up and reboiler duty is not exact, but it is usually sufficiently close so that the two variables can be used interchangeably.

Distillation control configurations are frequently described by the two variables that are used for composition control or not used for inventory/capacity variable control. For example, the LV configuration uses the reflux rate and reboiler duty to control the product compositions. By inference, the condenser duty is used for pressure control, the distillate rate is used to control the reflux accumulator level, and the bottoms rate is used to control the reboiler sump level.

- Complete Table W7.3 by listing the manipulated variables MV for each of the controlled variables.
- Why is the LD or BV configuration not likely to produce satisfactory control?
- Why is the DB configuration not likely to produce satisfactory control?
- For the DV and LB configurations, which variable should be used as the primary composition control variable and which should be used as the secondary composition control variable?

The LV control configuration is often described as an energy balance configuration and the DV and LB configurations are material balance configurations. This is because the DV and LB configurations manipulate the feed-split or material balance directly

Table W7.3 Distillation column control configurations

Control configuration	MV for reflux accumulator level control	MV for reboiler sump level level control	MV for primary composition control	MV for secondary composition control
LV				
DV				
LB				

by changing one of the product rates. However, the LV configuration only affects the feed-split indirectly through the level controllers.

The basic distillation control configurations have been listed above. However, there are many other configurations which use linear or even nonlinear combinations of the basic manipulated variables. One common example, which is sometimes called Ryskamp's scheme [1], manipulates the reflux ratio L/D, via ratio control, and the reboiler duty V. Another relatively common scheme is the double ratio configuration, which manipulates the reflux ratio and the boil-up ratio. This scheme has been widely recommended as it results in relatively small interactions between the two control loops. This concept will be discussed in further detail at a later stage.

- The principal control objectives for the stabilizer were previously listed as 0.01 wt% propane in the bottoms and fixed energy consumption. Describe how each of the three control configurations listed in Table W7.3 could be set up to satisfy these control objectives.

- Test these configurations via simulation. Are there any significant differences between the three schemes?

Hint: you may have to vary the tuning constants to get equivalent performance.

Present your findings on diskette in a short report using MS-Word. Also include on the disk a copy of the HYSYS files which you used to generate your findings.

Reference

1. Ryskamp, C. J. New strategy improves dual composition column control (also effective on thermally coupled columns). *Hydrocarbon Process*, 1980, June: pp. 51–9.

Workshop 8 Plant operability and controllability

There's a better way to do it. Find it!
Thomas Edison

Introduction

Prior to attempting this workshop, you should review Chapters 8, 9, and 10 in the book.

Traditionally, process design has been performed using steady-state analysis only. Simple rules of thumb have been used to size vessel hold-ups and to set other variables that affect the dynamic performance of a plant. This can sometimes lead to operability and controllability problems, as a design might look good in the steady state but be very difficult to operate or control due to the presence of dead times or insufficient capacitance.

A key consideration for plant operability and controllability is variable interaction. We have learned that dead time is one of our enemies, as it always makes tight control more difficult to achieve. Variable interaction places similar restrictions on the way we can control a process and can significantly reduce the overall control system performance. Three common sources of variable interaction are the nature of the process (i.e. distillation), the combination of multiple unit operations and heat integration. Each of these points can be highly advantageous in the steady state, but they can also create operability and controllability problems that may not be evident without considering the process dynamics at the design phase.

This workshop will investigate several examples where variable interaction is significant and will introduce an analytical technique for finding the best variable pairings in multiple-input–multiple-output MIMO systems. The potential trade-off between capital savings and plant operability will also be demonstrated. The problems in this workshop are more open ended than other workshops. You are encouraged to work more freely and continue your analysis until you are satisfied that you have pursued all paths.

A Real-Time Approach to Process Control, Second Edition W. Y. Svrcek, D. P. Mahoney and B. R. Young
 ISBN: 978-0-470-02533-8

Key learning objectives

1 Distillation control can usually be reduced to a 2×2 control problem. Interaction between variables plays a key role in control strategy selection and performance.

2 Inventory control (i.e. reflux accumulator and reboiler sump) should always be via the largest outlet stream.

3 The RGA is a control system design technique that can be used to minimize control loop interaction.

4 The NI will prove whether the control system design is stable for a 2×2 system.

5 Process understanding and a clear understanding of the key process objectives is essential to the development of a good control scheme.

6 Tight process control requires that the equivalent dead time in a loop should be small compared with the smallest time constant of a disturbance with significant amplitude.

7 There is often a trade-off between steady-state cost savings and dynamic operability.

8 Steady-state minimum-cost designs utilize very small hold-ups and high levels of heat integration. Both of these factors reduce the dynamic operability and controllability of a process.

9 Too much hold-up provides good attenuation but makes the process too slow to respond to disturbances.

Tasks

1 Two-point composition control for distillation

Most industrial distillation columns are operated similarly to the stabilizer that we studied in the previous workshop. One degree of freedom is used to control a product composition and the second available degree of freedom is used to control fractionation or energy consumption. This mode of operation is often called one-point or single composition control.

Sometimes both distillation products are equally important or equally valuable, and so both the bottoms composition and the distillate composition need to be controlled. This is called two-point or double composition control, and results in a much more difficult control problem than one-point composition control [1]. The primary cause of the extra difficulty is the interaction that exists between composition loops in a distillation column. This property of distillation columns (inherent interactions between two or more control loops) is called ill-conditioning.

Among the problems created by ill-conditioning is that there usually exists only a very narrow operating range that satisfies both composition control loops. Essentially, both the manipulated variables being used for composition control need to be adjusted together to produce the required results. Many newer control schemes are based on this principle, including model-based control and dynamic matrix control [2,3].

A second problem with two-point composition control is that there are no degrees of freedom left to operate around equipment constraints such as a reboiler duty limitation or flooding limitation. One-point composition control schemes have one degree of freedom which is not used for composition control and is available for this purpose (cf. our stabilizer, which is operated at a fixed heat input). This can create operational difficulties for many industrial columns.

Reconfigure the controllers on the stabilizer from Workshop 7 to provide two-point composition control. Assume that each product is equally important and that the control objectives are 0.01 per cent propane in the bottoms and 0.1 per cent isopentane in the distillate. Also assume that you have two perfect analysers (i.e. no dead time, no error) available so that the two compositions can be controlled directly.

Select one of the basic distillation control configurations (i.e. the LV, DV, or LB configurations) and tune composition controllers for both the distillate and bottoms products. Test the responsiveness of your candidate control structure using dynamic simulations of the stabilizer and step disturbances.

- Can you configure the system to give tight control of both the bottoms and distillate compositions and give good responsiveness to set-point changes?

- Do the controllers interact?

  Hint: if one loop is stable, does a set-point change in the other loop disturb the first loop?]

- Using steady-state analysis techniques, find an improved control configuration as follows:

  1 For each of the three feed conditions given in Table W7.1, record the distillate rate, bottoms rate, reflux rate and reboiler duty when the two composition specifications are satisfied simultaneously. This information should be already available from your previous steady-state simulation.

  2 Complete Table W8.1 by calculating the ratios of the manipulated variables and/or ratios to the feed rate (i.e. B/F, L/D, Q_R/B and others, if necessary).

  3 Find the combination of two manipulated variables (ratios) that shows the smallest variability for the whole range of feed conditions. This is done

Table W8.1 Manipulated variables and calculated ratios for steady-state analysis technique

Manipulated variables and ratios	Base case	Minimum flow	Maximum flow
D (bbl day^{-1})			
B (bbl day^{-1})			
L (bbl day^{-1})			
Q_R (MW)			
B/F			
L/D			
Q_R/B			

since it will maximize the natural disturbance attenuation of the control system.

4 Outline a candidate composition control structure using the two ratios chosen, making sure that you apply the normal rules of steady-state sensitivity and dynamic responsiveness for distillation control from the previous workshop.

Hint: you will need to use the spreadsheet operation and cascade controllers. Using the ratios from Table W8.2 for direct control is not feasible. The correct approach for the control structure is outlined in Ryskamp [1].

Test the responsiveness of your candidate control structure using dynamic simulations of the stabilizer. Do not forget that you will have to retune the two composition control loops.

- Has loop interaction been reduced?
- Is the overall control better than the basic control configuration that you tested above?
- If you did not have perfect analysers available or did not want to introduce dead time or error that would be present with real analysers, is there a combination of easily measured temperatures that you could successfully use to infer the distillate and bottoms compositions for the whole range of feed variance given in Table W7.1?

2 Relative gain array

The relative gain array is a tool that can be used to select an appropriate control structure from several candidate structures in a multiple-input–multiple-output system. The relative gain is the ratio between the open-loop gain and closed-loop gain in a system. In a distillation column, only the composition control variables are normally considered. The open-loop gain is the process gain between the controlled and manipulated variables with the secondary manipulated variable held constant. The closed-loop gain is the process gain between the controlled and manipulated variables with the secondary controlled variable held constant.

The relative gain array Λ has a property that makes the calculation of all the elements of the array unnecessary. This is shown in Equation W8.1 and applied to a 2×2 system in Equation W8.2. λ_{ij} refers to the relative gain between the ith controlled variable and jth manipulated variable.

$$\sum \lambda_{ij} = \sum \lambda_{ji} = 1 \tag{W8.1}$$

$$\Lambda = \begin{bmatrix} \lambda_{11} & \lambda_{12} \\ \lambda_{21} & \lambda_{22} \end{bmatrix} = \begin{bmatrix} \lambda_{11} & 1 - \lambda_{11} \\ 1 - \lambda_{11} & \lambda_{11} \end{bmatrix} \tag{W8.2}$$

- Determine the λ_{11} element of the relative gain array for each of the basic distillation control configurations (i.e. the LV, DV and LB configurations). Consider only the composition control loops so that the problem reduces to a 2×2 system.

- Which configuration appears to be most suitable for the stabilizer? Does this agree with your dynamic simulation results?

Hint: the gains you need can all be calculated via simulation.

3 Reactor temperature control

A key design consideration with exothermic reactions is the utilization of the heat of reaction. This energy optimization is often critical for plant profitability. The obvious heat integration technique is to use the hot reactor product to heat the cold reactor feed. The alternative would be to employ a hot utility to heat the reactor feed to near the reactor temperature and then a cold utility to cool the reactor product to the desired level.

Build the simple reactor system shown in Figure W8.1. The reactor should be modelled as a separator with a volume of 2 m^3. The reaction, given below, is equilibrium limited. The relationship between K_{eq} (in terms of activities) and the reaction temperature (in kelvin) is also given below. The system is very non-ideal, so an activity model should be used, i.e. the UNIQUAC model, and the reaction equilibrium should be measured in terms of activities rather than molar concentrations.

$$CH_3OH + CH_2{=}C(CH_3) \Leftrightarrow CH_3OC(CH_3)$$

$$\text{methanol} + \text{isobutene} \Leftrightarrow \text{MTBE}$$

$$K_{eq} = -10.54 + 4870/T(\text{K})$$

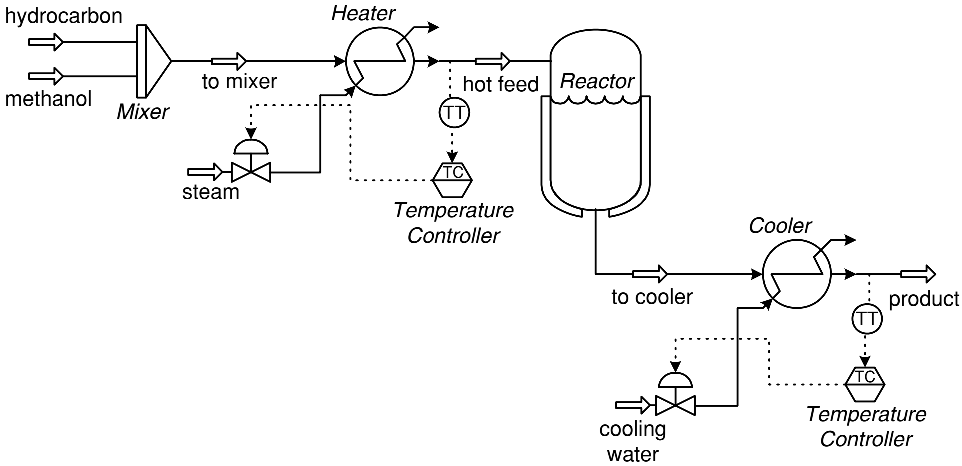

Figure W8.1 MTBE reaction system without heat integration

The feed is composed of two streams. The first stream is a hydrocarbon stream that contains 30 mol % isobutene and 70 mol % 1-butene. The second stream, consisting of pure methanol, is in 5 per cent molar excess of the reaction stoichiometry. The hydrocarbon feed rate is 1000 kg h^{-1}. Both streams are at 30°C and 1500 kPa. The reactor inlet temperature should be controlled at 70°C. The reactor outlet temperature will be higher than the inlet, since the reaction is exothermic and a considerable amount of heat is released. This has the effect of limiting the conversion of isobutene in the reactor. The reactor product should be cooled to around 40°C so that a second reaction stage can increase the isobutene conversion to around 99 per cent. The reactor pressure drop is 140 kPa, and the pressure drops through the exchangers are 70 kPa. The exchanger volumes can be estimated at 0.1 m^3 each.

Set up temperature control loops on both heat exchangers. Then, tune PI controllers for both of these temperature control loops.

- Test your control system for disturbances in the feed rate, feed temperature and feed composition. Use the suggested disturbances, listed in Table W8.2 as starting points for your analysis.

- Which disturbances are most difficult to control?

- How does the isobutene conversion vary during disturbances?

Hint: you may need to use a spreadsheet to calculate the isobutene conversion continuously.

Modify the system to incorporate heat integration between the reactor outlet (hot) and the reactor feed (cold), as shown in Figure W8.2. Again, assume pressure drops of

Table W8.2 MTBE reaction system disturbances

	Base case	Test case 1	Test case 2	Test case 3
Hydrocarbon feed rate (kg h^{-1})	1000	750	1000	1000
Feed temperature (°C)	30	30	45	30
Hydrocarbon feed composition (mol%)	30% i-But 70% 1-But	30% i-But 70% 1-But	30% i-But 70% 1-But	40% i-But 60% 1-But

70 kPa in both sides of the exchangers and a volume of 0.1 m^3. This should also reduce the load on the existing cooler.

Retune both temperature controllers. Test the new control structure against a similar range of disturbances.

- How much energy is saved by using heat integration in the process? Consider both heating and cooling duties.
- Does the system with heat integration still provide adequate control?
- Are the controllers stable? (Calculate the NI)

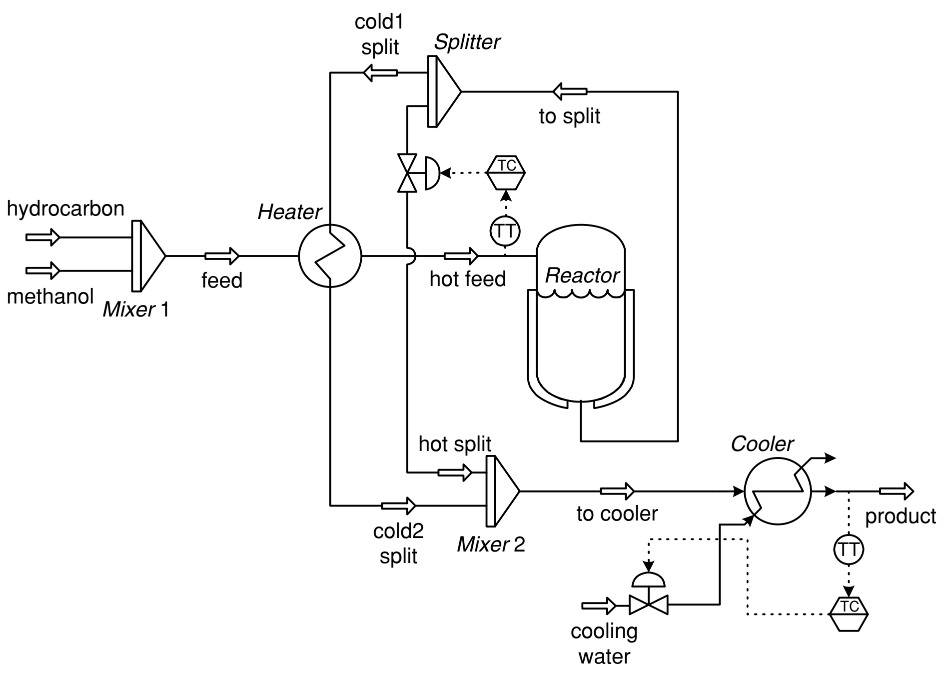

Figure W8.2 MTBE reaction system with heat integration

- What implications does the controllability or lack of controllability have for safety?
- Overall, which process configuration would you prefer?
- How could you modify the system to incorporate elements of both process designs to minimize utility consumption without compromising operability, controllability and safety?

Present your findings on diskette in a short report using MS-Word. Also include on the diskette a copy of the HYSYS files which you used to generate your findings.

Reference

1. Ryskamp, C. J. New stategy improves dual composition control (also effective on thermally coupled columns). *Hydrocarbon Process.*, 1980, June: pp. 51–9.
2. Cutler, C. R. and Ramaker, B. L. Dynamic matrix control – a computer control algorithm. In *AIChE National Meeting Joint Auto Control Conference*, Houston/San Francisco, 1979.
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