A Real-Time Approach to Process Control

Third Edition

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WILEY

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Contents

Author Biographies

William Svrcek is a retired (2009) professor of Chemical and Petroleum Engineering of the University of Calgary, Alberta, Canada, and is currently a professor emeritus of the University of Calgary. He received his B.Sc. (1962) and Ph.D. (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 200 technical articles/reports and has supervised over 50 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 worldwide. This course was 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 Alberta and a member of professional societies that include the Canadian Society for Chemical Engineering, American Institute for Chemical Engineers and the Gas Processors Association of Canada.

Donald Mahoney is vice president and global head of chemicals for SAP, the world's largest enterprise software company. 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's Lerner College of Business and Economics. Mr. Mahoney has held research and teaching positions at the US Navy's Applied Research Lab at University Park, and at Purdue University, where he was awarded the department's highest honour for outstanding teaching. 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 SAP, Mr. Mahoney was a business software entrepreneur and vice president with Hyprotech Ltd, where he led the introduction and launch of more than half a dozen process design, modelling and optimization software products. He also 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™.

Brent Young is a professor in the Department of Chemical and Materials Engineering at the University of Auckland and is currently the head of department. He also holds the position of chair in Food and Process Engineering and is the director of the Industrial Information and Control Centre. He received his B.E. (1986) and Ph.D. (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. In 1991, he joined the University of Technology in Sydney, Australia, as a lecturer, continuing his research in the areas of modelling and control of processes, particularly industrial processes. He was then 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 fellow of the Institution of Chemical Engineers. His research, teaching and practice are centred on two major areas, process simulation and control, and process design and development.

Foreword and Endorsements

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, Principal Consultant, CDI Process & Industrial, Emerson

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 Process Modeling and
Advanced Process Control Specialist, 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

The control engineering profession has produced shelves of books. For the most part they have been written to support academic courses and are authored by lecturers who teach the subject using theory not relevant to the process industry and mathematics that most students find daunting. This book belongs on the shelf labelled 'Process Control for Process Engineers'. It is one of a hopefully growing collection written by authors who recognize that the practical application of control techniques in the process industry is a quite different subject.

The money invested in process control by the process industry has grown substantially over the last few decades. Now around a quarter of the construction cost of a modern plant is associated with its control and optimization. The industry needs professionals that properly understand the technology and what it can achieve. But highly theoretical courses dissuade most process engineering graduates from entering the control engineering profession. Those that do find rewarding that they can have an almost immediate impact on process performance.

This book provides a valuable introduction. It will help students appreciate the true nature of the subject and enable them to make an informed decision about whether to follow it in depth.

—Myke King, Director WhiteHouse Consulting, England

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 non-linearity 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 (SISO) 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. New in the third edition, Chapter 11 provides an introduction to Model Predictive Control (MPC). Also in this third edition, we have included a brief overview of the Fieldbus industrial network system, included in the Appendix. Finally, up-to-date information on computer simulation for the workshops and powerpoint slides can be found on the book web site http://www.wiley.com/go/svrcek-real-time-3e.

While 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 practicing 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.

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 practicing 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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1

A Brief History of Process Control and Process Simulation

In order to gain an appreciation for process control and process simulation it is important to have some understanding of the history and motivation behind the development of both process control and process simulation. Rudimentary control systems have been used for centuries to help humans use tools and machinery more efficiently, effectively and safely. However, only in the last century has significant time and effort been devoted to developing a greater understanding of controls and sophisticated control systems, a requirement of the increased complexity of the processes to be controlled. The expansion of the controls field has driven the growth of steady-state and dynamic process simulation from relative obscurity to the indispensable and commonplace tool that it is today, in particular in the development of operator training systems and the validation of complex control strategies.

1.1 Process Control

Feedback control can be traced back as far as the early third century BC [1,2]. 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 [3]. 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, Cornelis Drebbel [3], a Dutch engineer, used a bimetallic temperature

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regulator to control the temperature of a furnace. Denis Papin [3], 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.

One-port automata (open loop) evolved as part of the Industrial Revolution and focused on a flow of commands that mechanized the functions of a human operator. In 1801, Joseph Farcot [4] fed punched cards past a row of needles to program patterns on a loom, and in 1796, David Wilkinson [5] 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 mechanized 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] utilized piston displacement on a Newcomen engine to perform a deterministic control function. However, the fly ball governor designed by James Watt [7] 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 set 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 [8], 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 [9] 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 [10] 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 the theories of 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 emphasizes the irony seen by Elting Morison [11]. 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 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, Boundy and Cermak [12] that discussed the problem of pH control and showed the advantages of using derivative action to improve controller response. Callender, Hartree, and Porter [13] 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 [14], and Ceaglske, in 1956 [15]. 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 [16], Campbell [17], Coughanowr and Koppel [18], Luyben [19], Harriott [20], Murrill [21] and Shinskey [22]. Process control became an integral part of every chemical engineering curriculum.

Present-day process control texts that include Marlin [23], Seborg *et al.* [24], Smith [25], Smith and Corripio [26], Riggs [27] and Luyben and Luyben [28] have to some extent used a real-time approach via modelling of the process and its control structure using MATLAB Simulink [29] and Maple [30] to provide a solution to the set of differential equations, thus viewing the real-time transient behaviour of the process and its control system.

A book by King [31] titled *Process Control: A Practical Approach* is aimed at the practising controls engineer. It, like this text, focuses on the practical aspects of process control. This book is an excellent addition to the practising controls engineer's library.

The availability of minicomputers in the late 1950s and early 1960s provided the impetus for the use of these computers for centralized process control (DC). For instance the IBM 1800 of that time was equipped with a hardware interface that could convert measured temperatures, flows and so on (analog signals) to the required digital signals (PID). A number of early installations were only digital computer-based data loggers (Figure 1.1).

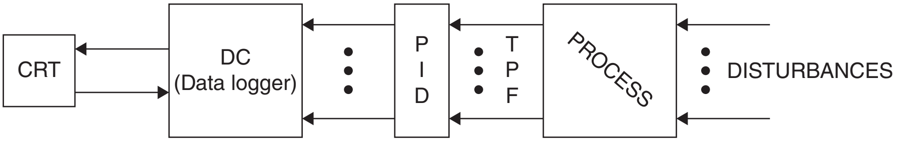

Figure 1.1 *Digital computer-based data logger of the late 1950s and 1960s.*

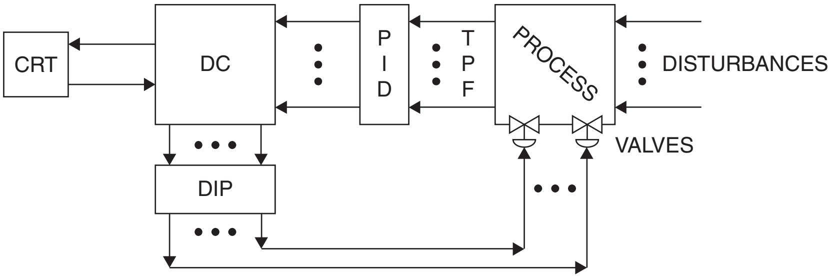

Figure 1.2 *Schematic of centralized digital computer control structures.*

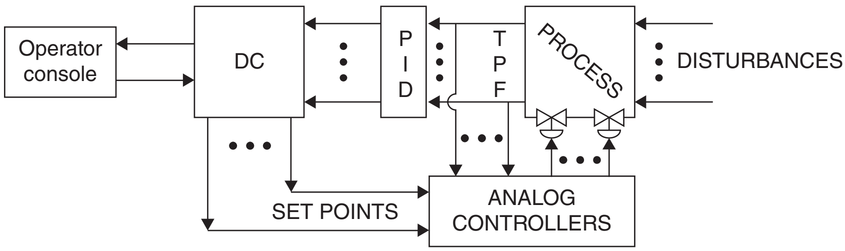

Figure 1.3 *Schematic of supervisory digital control systems of the 1970s.*

The first computer-based central control system [2] was installed in 1959 at the Texaco Port Arthur, Texas refinery, and was based on an RW-300 from Ramo-Woolridge (Figure 1.2). During the following decade a number of centralized digital control systems were installed in chemical plants and refineries [32]. These installations for the most part were supervisory (Figure 1.3) because these facilities could not risk a digital computer failure without a conventional reliable single-loop control structure as a back-up.

In 1975, both Honeywell and Yokogawa provided the first distributed control system (DCS) [2] (Figure 1.4). Over the next decade virtually all the control hardware providers developed and offered a DCS.

The fluid-processing industries quickly adopted this combination of hardware and software. This approach to process control offered a natural extension to the typical

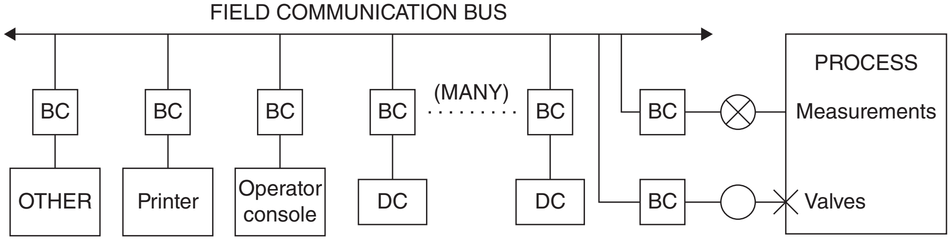

Figure 1.4 *Schematic of a modern distributed control system (DCS).*

process/plant SISO control loops. The plant controllers and measurements were not centrally located but 'distributed' throughout the plant. Hence, basic control is achieved at the local loop level. These local controllers and local measurements are then connected via a communication network for monitoring and display to a central control room. A central aspect of the current DCSs is the quality and detail of the plant equipment hardware displays, the process measured and controlled variable displays.

Figure 1.2 is a schematic of a centralized digital control structure while Figure 1.4 shows the structure of a typical DCS. The obvious difference is the distribution of controllers as groups of digital controllers, that is, only the key loops are backed up with a SISO loop. The DCS has the major advantage that even if the central processor should fail the underlying control system continues to function.

DCS hardware will be discussed further in Chapter 2. DCS software has developed to the point of providing advanced control strategies such as MPC and DMC, detailed graphic displays and user programming capability – in other words, very operator friendly.

1.2 Process Simulation

Prior to the 1950s, calculations had been done manually[1] on mechanical or electronic calculators. In 1950, Rose and Williams [33] 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 [34].

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, most significant of which has been the proliferation of powerful desktop computers [personal computers (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 [35]. Figure 1.5 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 [36]. 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

[1] Using a slide rule.

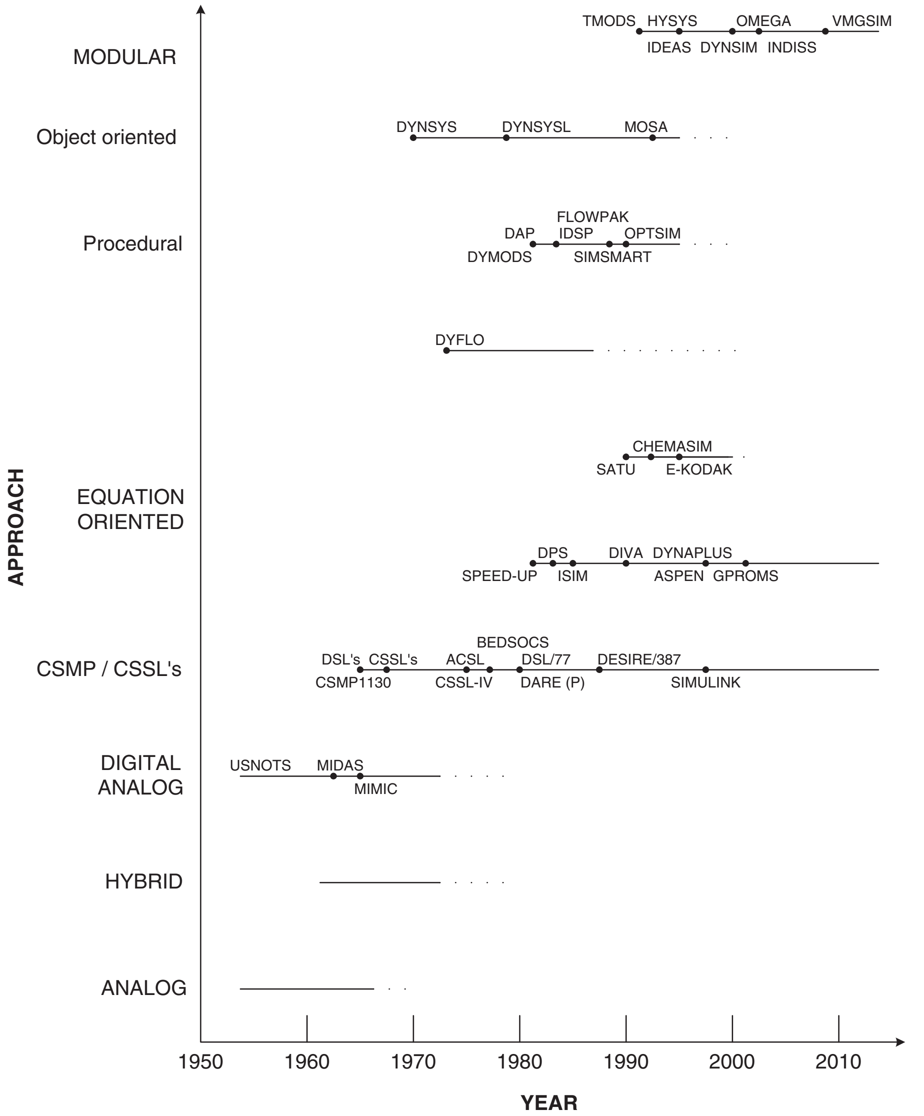

Figure 1.5 *Development of dynamic process simulators.*

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 [36]:

1. Hybrid computers required detailed knowledge of operation for both analog and digital computers. This translated into long training periods[2] before an engineer was able to work with the hybrid computer.

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[2] One week or more.

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, 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.[3]
4. Simulations using hybrid computers were extremely time consuming. An engineer had to reserve time in the hybrid simulation lab and work in this lab 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 when compared to 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. Latter 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 [37] is the most widely used.

Parallel to the previous approach has been the development of equation-based numerical solvers like SPEEDUP [38]. 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 made by Franks and Johnson [39]. The two early modular simulators, DYFLO and DYNSYS, 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 [39], 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 due 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 [40] in 1991 presented a paper summarizing key developments and future challenges in dynamic process simulation.

[3] Human nature.

Since this review three additional dynamic process simulators have appeared – Odysseo [41], Ideas [42] and VMGSim [43].

The key benefits of dynamic simulation [44] 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 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 minimize capital expenditures and operating utility costs, many plant designers have adopted the use of steady-state optimization 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 hold-up 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 optimization 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, minimize waste, maximize yield, reduce utility costs and often 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 minimize 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 to quickly and easily 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 (OTS). In order to be effective, the training simulator should be interactive, be realistic and run real time. By running a relatively high-fidelity model, operators can test 'what if' scenarios, practice 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, 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 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 minicomputers. 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, 50–80% of the time dedicated to a dynamic study was consumed in the model-building phase. Roughly 20% was dedicated to running the various case studies and 10% 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 [35]

- the growth of computer hardware and software technology and
- 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 8 GB, a 512 GB hard drive, and a large flat-screen graphics monitor on his desktop costing less than $1000. Furthermore, a number of powerful and interactive window environments have been developed for the PC and other inexpensive hardware platforms. Windows (2000, NT, XP, VISTA, 7, 8, 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 and user-friendly 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 – must have an intuitive and interactive, graphical environment that involves no writing of code or any compile–link–run cycle.
- Configurable – must provide reusable modules which can easily be linked together to build the desired model.
- Accurate – must provide meaningful results.
- Fast – must strike a balance between rigour and performance so as not to lose the interactive benefits of simulation.
- Broadly applicable – must provide a broad range of functionality to span different industrial applications, as well as varying levels of detail and rigour.
- Desktop computer based – 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 as 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 [45] 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%. 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 [46] limited the acceptance and use of process control theory. However, this situation was changed during the early 1990s with the availability of commercial robust high-fidelity process dynamic simulators and has led to the frequent use of this software tool. The vendors and a book by Luyben [47] do provide a number of examples and guidelines for the development of useful process dynamic simulations.

In summary, the traditional approach to control loop analysis has been through the use of frequency domain techniques such as Bode diagrams, transfer functions and Nyquist plots. 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 different equations. Although these frequency domain techniques are useful for single control loops they are not easily applicable to real multi-loop and nonlinear systems which comprise the actual plants that must be controlled in the fluid-processing industries.

In the real-time[4] 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. It is therefore 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 percent opening has been adjusted to correct for any deviations from the set point.

A Real-Time Approach to Process Control, Third Edition. William Y. Svrcek, Donald P. Mahoney and Brent R. Young.

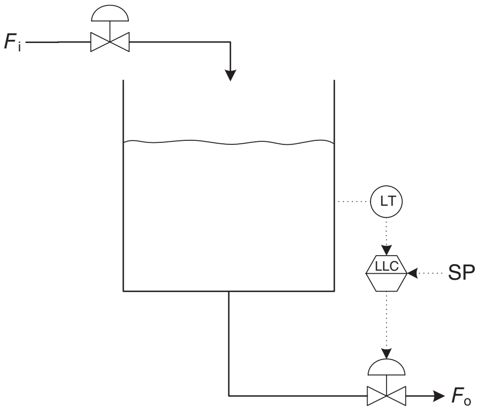

Figure 2.1 *Surge tank level controller.*

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 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 and pressure. A full listing of the types of primary

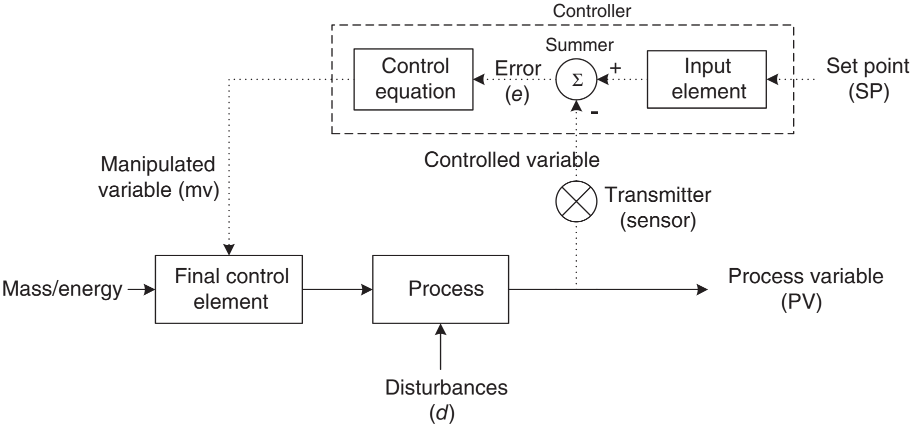

Figure 2.2 *Single-input/single-output block diagram.*

elements available on the market would be very long, but these sensor types 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, GLC, 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. This later point applies particularly to quality or analysis instruments which are only briefly introduced here. Further information and detail can be found in the references.

2.2.1 Pressure Measurement

There are numerous types of primary elements 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 and 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 (ca. 1852) and shown in Figure 2.4, is probably the most common gauge used in industry. Figure 2.4 illustrates the Bourdon tube pressure gauge. 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 $\pm 0.5\%$ of full range for commercial models. Generally, the normal working pressure will be specified as 60% of the full scale.

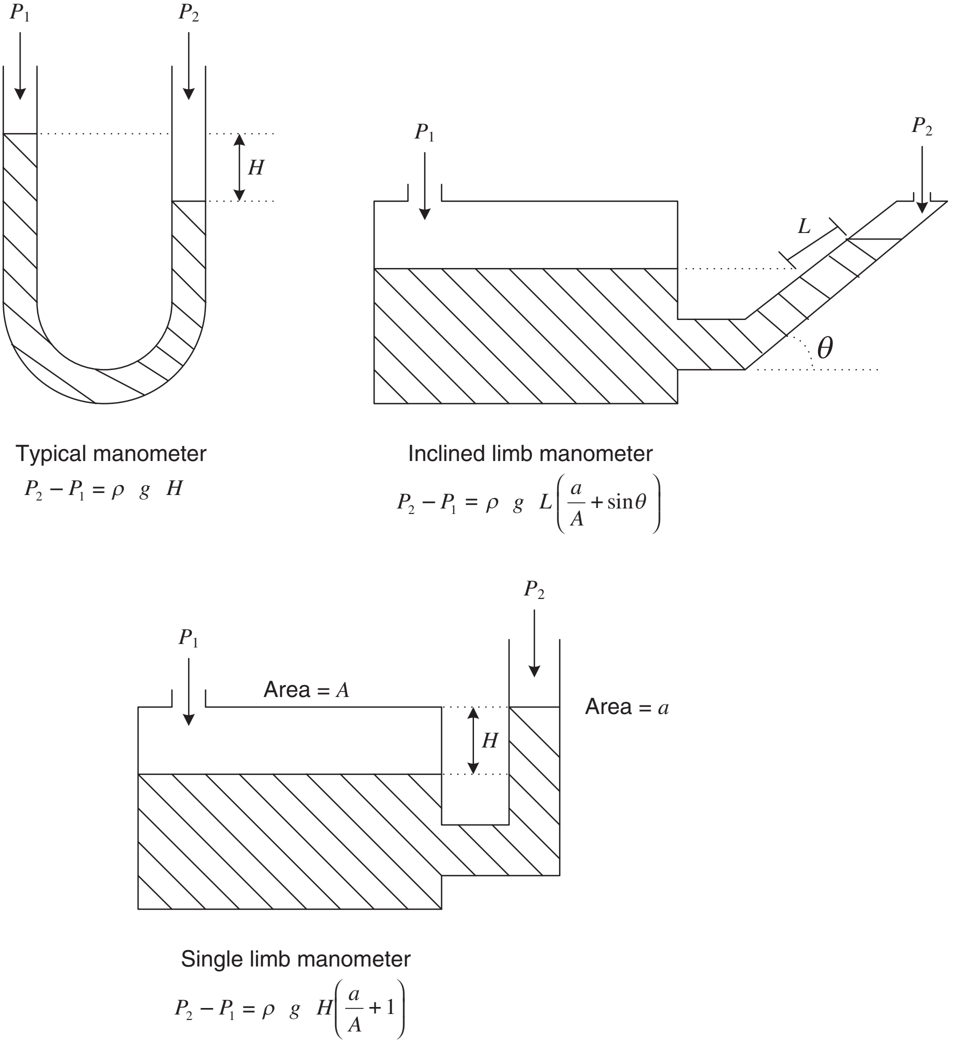

Figure 2.3 *Various manometer types.*

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

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

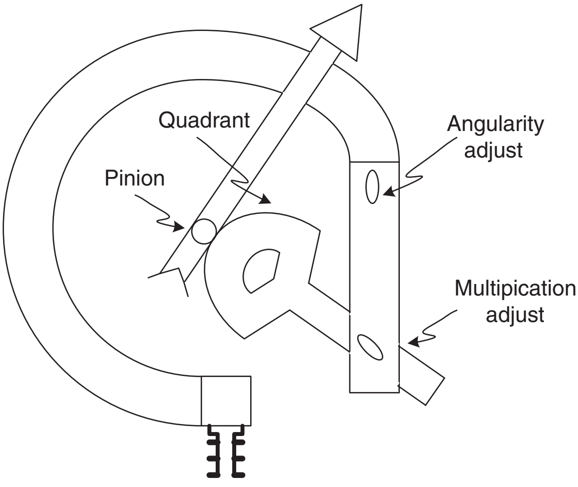

Figure 2.4 *Bourdon tube pressure gauge.*

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) 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 (−73–427°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 DP 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 DP. The range of the cell is 1.20–210 kPa DP with an accuracy of $\pm 0.5\%$ of the range.

Modern DP 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 2.7 is a photo of a modern DP cell.

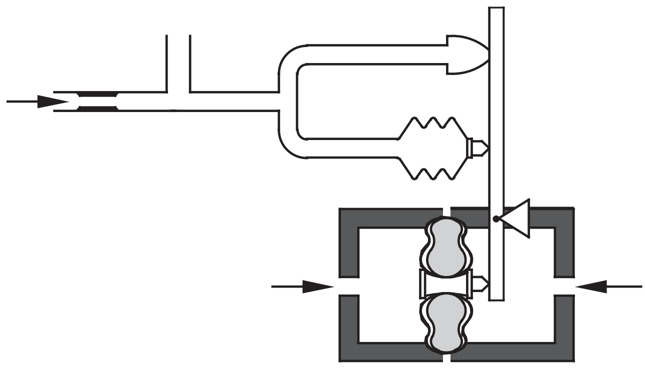

Figure 2.5 Pneumatic DP cell (Reproduced by permission of Emerson Process Management).

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.

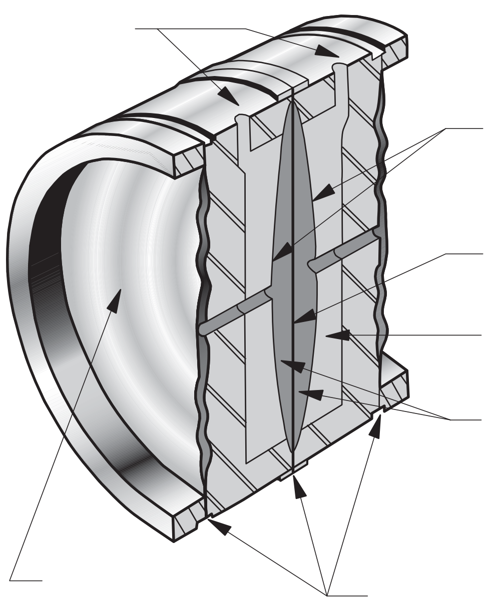

Figure 2.6 Electronic DP cell (Reproduced by permission of Emerson Process Management).

Figure 2.7 *Model 3051 electronic DP cell (Reproduced by permission of Emerson Process Management).*

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, that is, 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 measurements 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 3051SMV 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].

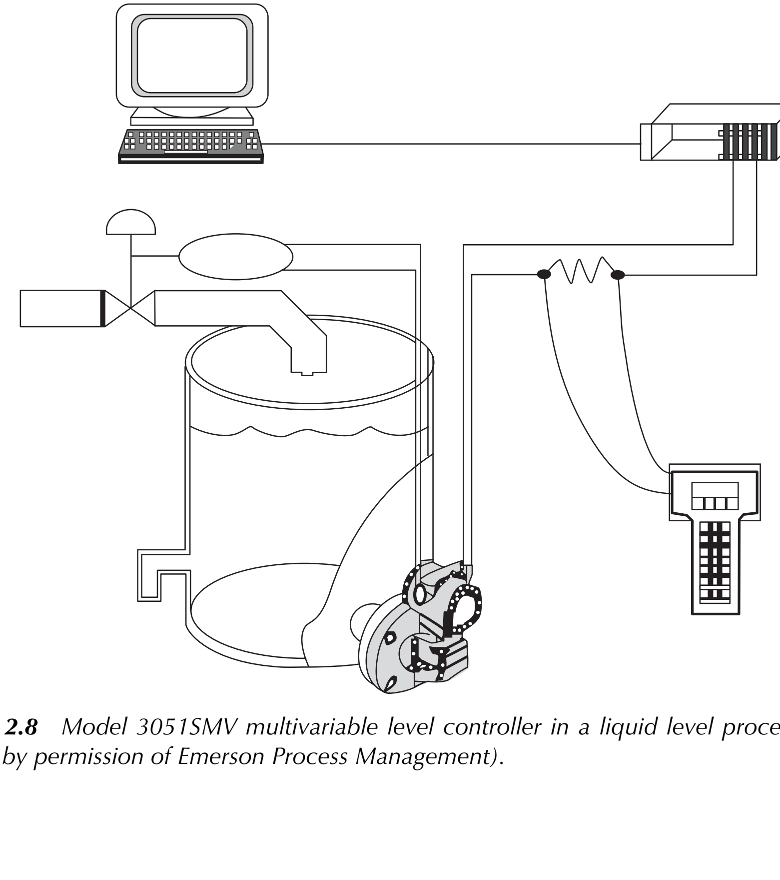

Figure 2.8 *Model 3051SMV multivariable level controller in a liquid level process (Reproduced by permission of Emerson Process Management).*

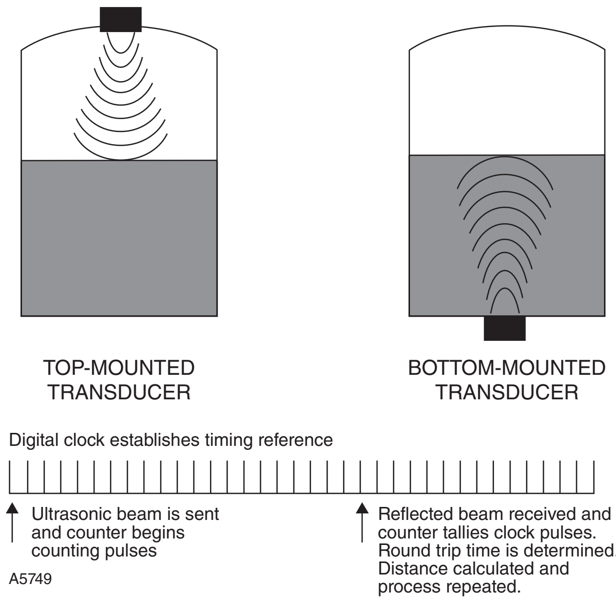

Figure 2.9 *Ultrasonic level transmitter (Reproduced from [9]).*

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 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).

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 measured fluid; 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.

The temperature range for bimetal thermostats is 0–400°C with an accuracy of ±5%, although the accuracy can be increased to ±1% [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 and 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 (emf). This is known as the Seebeck

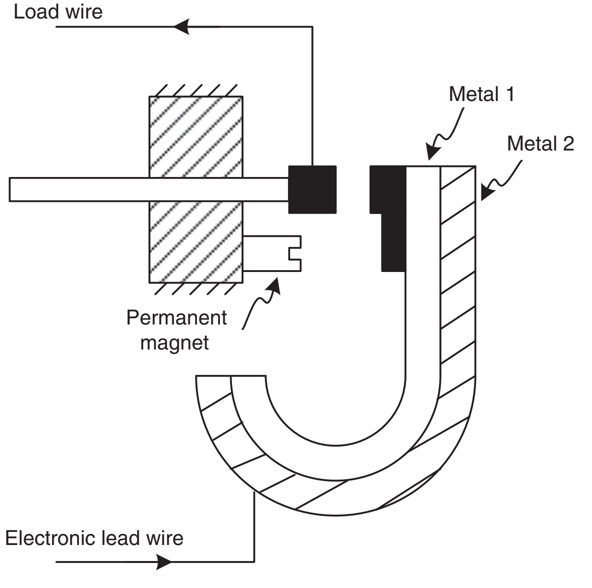

Figure 2.10 *Bimetal thermostat.*

effect, after Seebeck's 1821 discovery of this phenomenon. The magnitude of the emf will depend on the types of materials 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 emf generated will be a measure of the temperature of the other junction. This other junction is called the hot junction.

The emf 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 them together. The relationship between temperature and generated emf is non-linear except over limited ranges. On the steep part of the curve, the relationship is

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

In Equation 2.2, e is the generated emf, T_1 and T_2 are the hot and cold junction temperatures in 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)

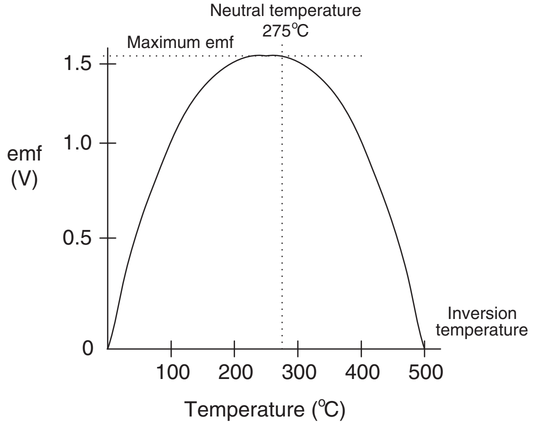

Figure 2.11 *Relationship between temperature and emf for a Cu/Fe system.*

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

- Iron (Fe) deteriorates quickly due to scaling in oxidizing atmospheres at high temperatures.
- Chromel and alumel thermocouples are poisoned by gases that are carbon based, are 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 (RTDs)

RTDs are made of either metal or semiconductor materials as resistive elements that 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 to thermocouples, RTDs have higher accuracy, better linearity and 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).

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.

2.2.4 Flow Measurement

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

1. Obstruction-type meters, such as
   (a) Orifice plates
   (b) Flow nozzles

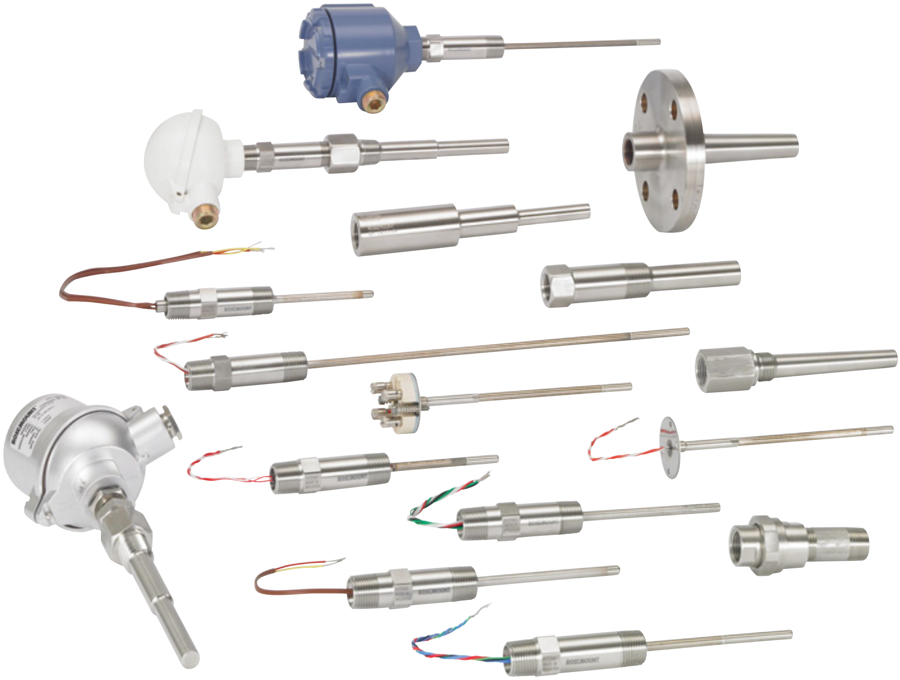

Figure 2.12 *Selection of thermocouples, RTDs and accessories (Reproduced by permission of Emerson Process Management).*

   (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

Figure 2.13 *Temperature sensor and transmitter assembly (Reproduced by permission of Emerson Process Management).*

4. Ultrasonic and thermal-type meters
5. Square root extractors for obstruction-type meters
6. Quantity or total flowmeters, 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 flowmeters, 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 DP 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 homogenous fluids, which include liquids, vapours or gases, whose viscosity does not exceed 65 cP at 15°C. In general the Reynolds number (Re) should not exceed 10 000. The plate thickness should be 1.5–3.0 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 utilized include eccentric and segmental orifice configurations.

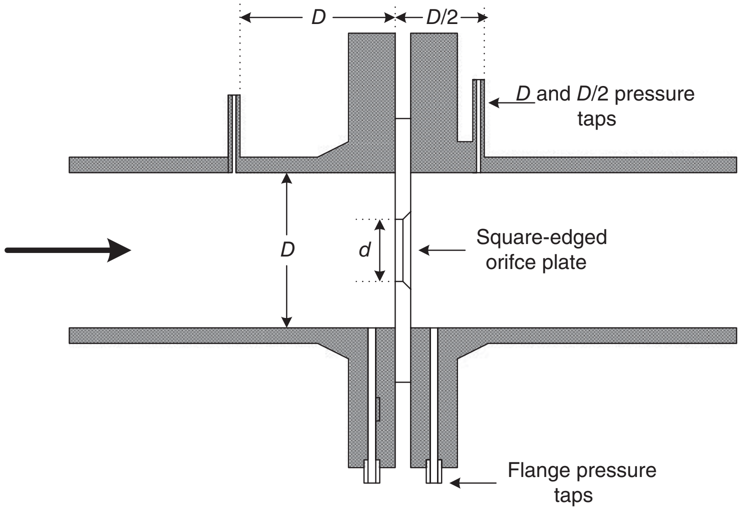

Figure 2.14 *Thin-plate orifice meter.*

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 on Faraday's law (see Equation 2.3), which states that a voltage will be induced in a conductor moving through a magnetic field:

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

In Equation 2.3, E is the generated emf, 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.

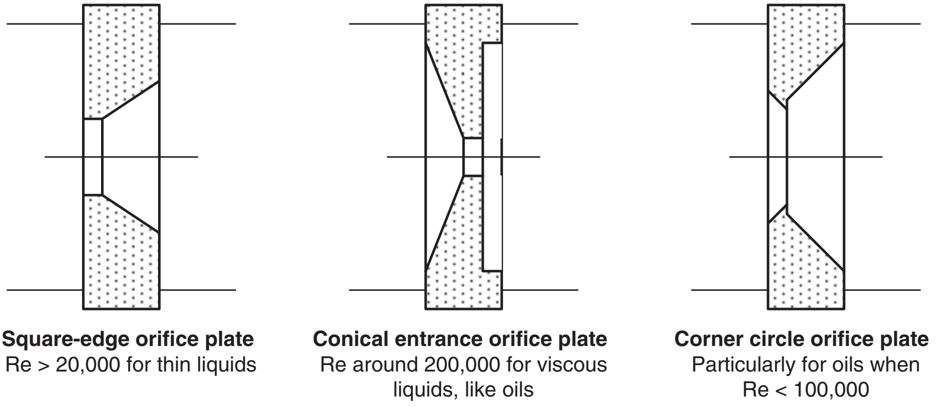

Figure 2.15 *Various orifice plate designs.*

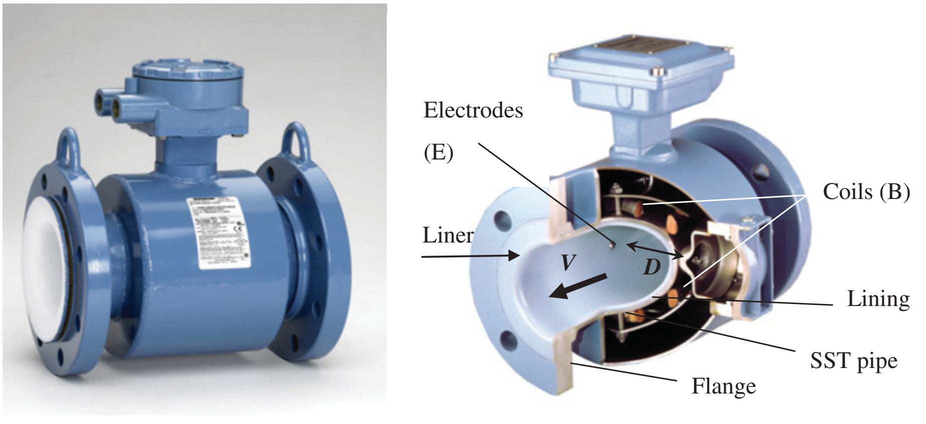

Figure 2.16 *Views of the Model 8705 magnetic flowmeter with junction box (Reproduced by permission of Emerson Process Management).*

Figure 2.16 illustrates the principle of operation of a magnetic flowmeter. In the past, magnetic flowmeters have been very expensive compared to orifice plates and DP cells. However, now the cost is 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 almost any corrosive liquid. The electrodes can be made from very corrosive-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 can give out a digital signal that can be directly fed 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. Note too that proper grounding is critical to magmeter performance to ensure the magnetic field remains isolated from magnetic noise of nearby electrical equipment.

Vortex meters (Figure 2.17) utilize the von Kármán effect. This effect is readily observable when the wind blows and a flag flaps in the wind. In a vortex meter, fluid alternately separates from each side of the shedder bar face. Vortices form behind the face and cause alternating DPs around the back of the shedder bar. The frequency of the alternating vortex development is linearly proportional to the flow rate. The flexing motion is sensed by a piezoelectric sensor element which converts the motion to an alternating electric signal. Vortex flowmeters are suitable for liquids, gas and steam. They are the most cost effective in smaller lines (6 in. and smaller). It is important to recognize that this meter type has a low flow cut-off and does not measure to zero.

Coriolis flowmeters (Figure 2.18) make use of the Coriolis effect. The inertia force must be taken into consideration if Newton’s law of motion of bodies is to be used in a rotating

Figure 2.17 *Model 8800 vortex flowmeter (Reproduced by permission of Emerson Process Management).*

Figure 2.18 *Coriolis flowmeter (Reproduced by permission of Emerson Process Management).*

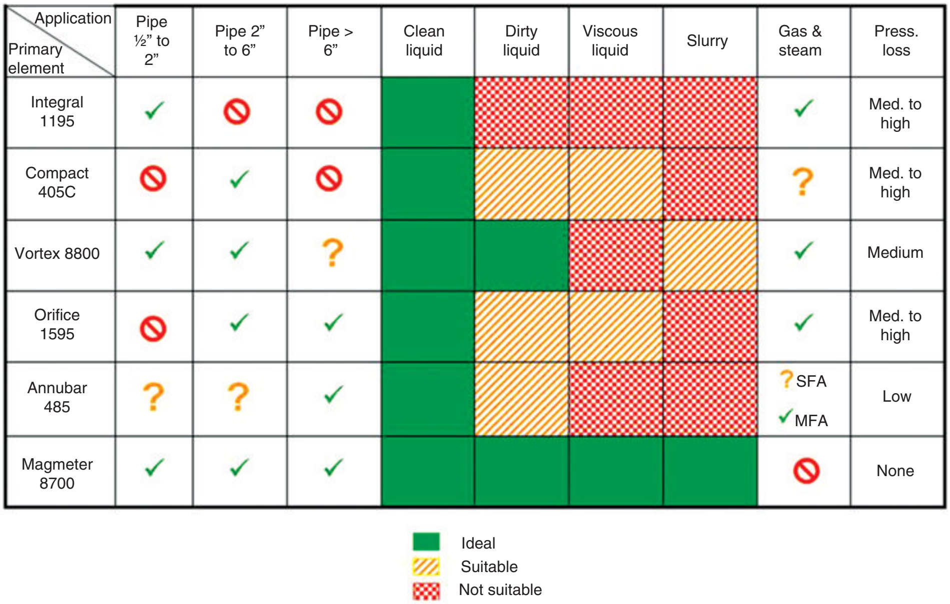

Figure 2.19 *Coriolis flowmeter (Reproduced by permission of Spartan Controls Ltd).*

frame of reference. As the flow passes through, the Coriolis meter tubes twist a bit. The amount of deflection is measured and used to determine the flow rate, but can also be used to determine temperature and density. This flow device is mechanically tolerant, can measure two-phase flow, does not require flow conditioning or straight pipe run lengths and is low maintenance.

A summary of the strengths and weaknesses of different flowmeters in various applications is highlighted in Figure 2.19.

2.2.5 Quality Measurement and Analytical Instrumentation

Primary elements include the rapidly growing range of analytical instrumentation. Chemical engineering graduates need at least an awareness of what is now measurable. For example flue gas oxygen (and CO) analysers are commonplace. Gas and liquid densitometers are used extensively in industry. Other common analysers include viscometers, calorimeters and chromatographs. Near infrared (NIR) spectroscopy has become very widely used as a general-purpose inferential composition measurement technique and nuclear magnetic resonance (NMR) spectroscopy looks like it could go the same way in a decade or so. The oil industry is also a big user of analysers for vapour pressure, flash point, distillation and cloud point. However, because of the specialized and detailed nature of these instruments, which are more the domain of physical chemistry, the reader is referred to the *Process Industrial Instruments and Controls Handbooks* by Considine [3] and McMillan and Considine [13] and the specialist book *Process Analyzer Technology* by Clevett [14] for further details.

2.2.6 Application Range and Accuracy of Different Sensors

Table 2.1 shows the application range and accuracy of most of the different sensors mentioned in the foregoing subsections. More details can be obtained from the books by Considine [3], McMillan and Considine [13] and Clevett [14], and of course instrumentation vendors.

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 are often costly as motor control is expensive, 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% 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)' [16].

The components that control a valve are [16]

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, trim and seats:
   (a) the body is a pressure vessel with a passageway through which the process fluid flows;
   (b) the trim is a closure member such as a plug, ball or gate that modulates the flow of process fluid through the valve;
   (c) 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 butterfly valve in Figure 2.20.

The sliding stem control valves are the most common control valve configuration and have at least half the market in control valves. Figures 2.21a and b show 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,

Table 2.1 *Application range and accuracy of sensors (Adapted from Marlin [15]).*

Variable	Pressure	Differential Pressure	Differential Pressure	Differential Pressure	Differential Pressure
Sensor type	Bourdon tubes	Diaphragm	Capacitance/ inductance	Resistive/strain gauge	Piezoelectric
Range/rangeability	up to 100 MPa	up to 60 kPa	up to 30 kPa	up to 100 MPa	–
Accuracy	0.5–5%	0.5–1.5% of full span	0.20%	0.1–1%	0.50%
Dynamics	–	–	–	Fast	Very fast
Advantages	Low cost	Very small span possible	–	Large range of pressures	Fast dynamics
	Okay accuracy				
	Wide limits of application				
Disadvantages	Hysteresis	Usually limited to low pressures (i.e. below 8 kPa)	–	–	Sensitive to temperature change
	Affected by shock and vibration				
Variable	**Level**	**Level**	**Flow**	**Flow**	**Temperature**
Sensor type	Level	Differential pressure	Orifice	Vortex shedding	Bimetallic
Range/rangeability	Float	Essentially no upper limit	3.5:1	10:1	–
Accuracy	Up to 1 m	–	2–4% of full span	1% of measurement	$\pm 2\%$
Dynamics	–	–	–	–	–
Advantages	–	Good accuracy	Low cost	Wide rangeability	Low cost, physically rugged, local display
		Large range	Extensive industrial use		
		Applicable to slurries with sealed lines	High pressure loss	Insensitive to variations in density, temperature, pressure and viscosity	

Disadvantages	Can be used for switches Cannot be used with sticky fluids which coat the float	Assumes constant density Sealed lines sensitive to temperature	Plugging with slurries	Expensive	Inaccurate Non-digital

Variable	Temperature	Temperature	Temperature	Temperature	Temperature
Sensor type	Type E thermocouple: chromel–constantan	Type J thermocouple: iron–constantan	Type K thermocouple: chromel–nickel	Type T thermocouple: copper–constantan	RTD
Range/rangeability	−100–1000°C	0–750°C	0–1250°C	−160–400°C	−200–650°C
Accuracy	±1.5°C or 0.5% for 0–900°C	±2.2°C or 0.75%	±2.2°C or 0.75%	±1.0°C or 1.5% for −160 to 0°C	$0.15 + 0.2\|T\|$°C
Dynamics	Depends on thermowell, ~2–5 s				Depends on thermowell, ~2–5 s
Advantages	Good Wide range				Good accuracy Small span Linearity
Disadvantages	Minimum span of 40°C Temperature versus emf not linear Low emf corrupted by noise Drift over time				Self-heating error Less physically rugged

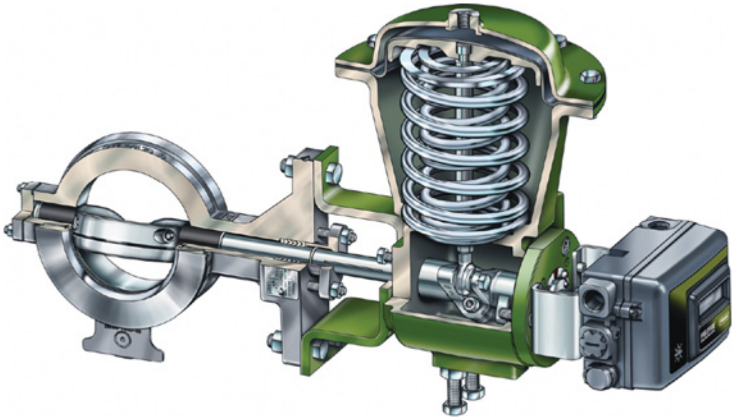

Figure 2.20 *Control valve cutaway (Reproduced by permission of Emerson Process Management).*

especially for high-pressure liquids. Figure 2.22 shows some of the newer styles of valve cages.

Control Valve Sizing

The common equation for the flow of a non-compressible fluid through a control valve is given in Equation 2.4 [17], which can be derived from Bernoulli's equation:

$$Q = C_{\mathrm{v}}\sqrt{\left(\frac{\Delta P}{\mathrm{SG}}\right)}. \tag{2.4}$$

(a) (b)

Figure 2.21 *(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).*

(a) Quick Opening (b) Equal Percentage (c) Linear

Figure 2.22 *Characterized cages for globe-style valve bodies (Reproduced by permission of Emerson Process Management).*

In Equation 2.4, Q is the volumetric flow rate and ΔP is the pressure drop across the valve, SG is the relative density compared to water, and C_v is the valve coefficient. C_v is defined by convention in field or imperial units as the number of US gallons that will pass through a control valve in 1 minute, 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' [17–21].

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.

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 in % 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 2.23.

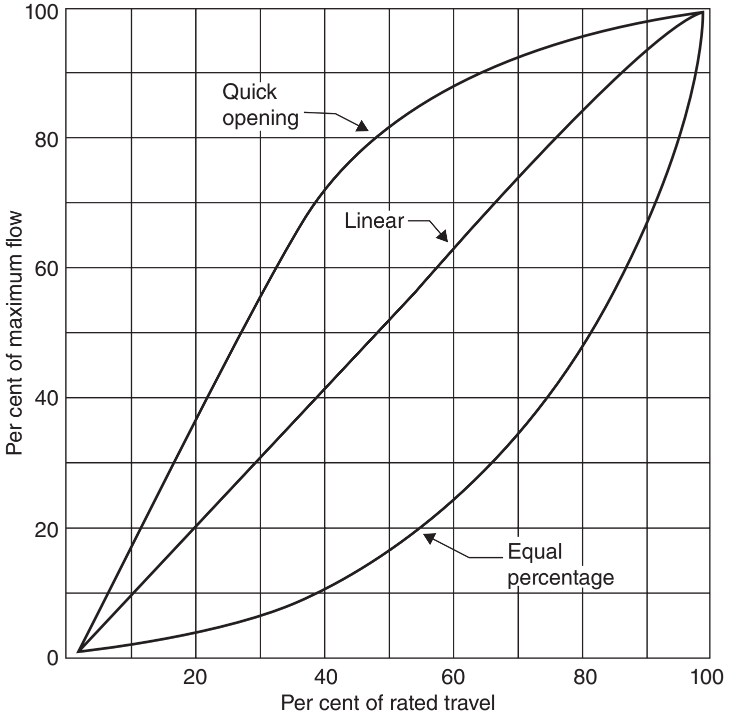

Figure 2.23 *Examples of inherent valve characteristic curves (Reproduced by permission of Emerson Process Management).*

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 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 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% open. The greater this variation, the more the operating characteristic

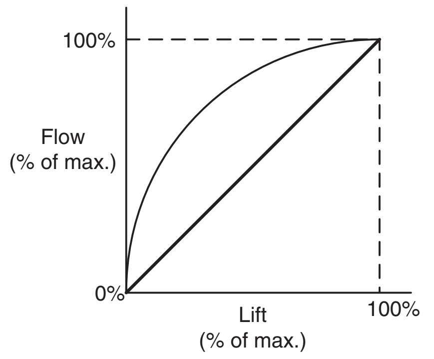

Figure 2.24 *Operating characteristic for a small value of β.*

varies from the inherent characteristic. A measure of this deviation is the parameter, β, defined by

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

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 2.24 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 valve size and valve inherent characteristic.

As previously stated, 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%. 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% 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 oversized 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

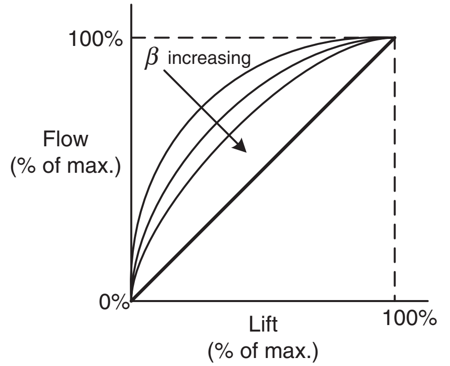

Figure 2.25 *Effect of increasing the value of β.*

should be selected so that the operating characteristic tends towards linear (Figure 2.25). In addition, the equal percentage characteristic is more forgiving when sized incorrectly or if there is insufficient pressure drop across the valve. Equal percentage valve plug trim has become the standard type of trim when valves are put into an existing system.

If the control system is included as part of the original design, 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 non-linear 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 roughly a 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, that is, 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, 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 and so on. 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

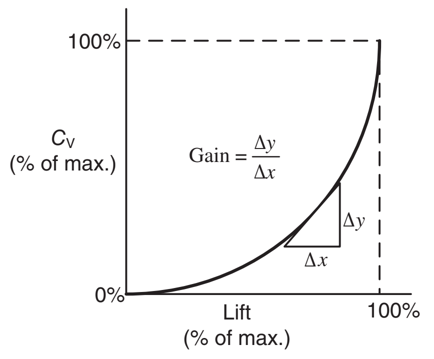

Figure 2.26 *Calculation of the control valve gain.*

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 2.26.

If this is done for the common types of control valves over their whole range then the gain curves shown in Figure 2.27 are obtained.

Quick Opening The gain increases to a maximum at about 20% of the valve opening from where it decreases exponentially resulting in decreasing effectiveness as the valve approaches fully open position. This indicates that the valve would be good for a process whose gain increases with the variable used to control it.

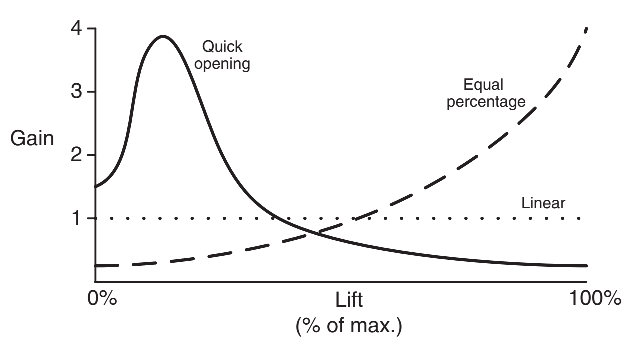

Figure 2.27 *Gain curves for common types of control valves.*

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

Equal Percentage The gain increases exponentially for equal percentage lifts between 10% and 100%. 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. [18, 19]).

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 % open data corresponding to the C_v (or C_g, etc.) values calculated in step 2 (vendor data C_v versus % open).
4. Calculate the control valve gain as

$$K_{cv} = \frac{\%\Delta\text{Flow}}{\%\Delta\text{Opening}}.$$

5. Calculate the average gain.
6. Are the minimum and maximum control valve gains within $\pm 50\%$ 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 valve's characteristics.

Control Valve Rangeability

The required range of C_v should fall between 20% and 80% 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 [22]. At less than 10% open, it can be difficult for the control valve to stabilize and it tends to oscillate, while at greater than 80% 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 which may have a decreased range of control. Often, the pressure drop to be taken across the control valves is specified when detailed plant hydraulics is 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% of the friction pressure drop taken across the equipment plus piping, or 33% of the total system pressure drop (excluding the valve). Minimum pressure drops have been stated as 10% of the system pressure drop for equal percentage valves, or 25% of the system pressure drop for linear valves [23] or 35 kPa for rotary valves and 69 kPa for globe valves [22].

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 at other conditions are governed by system hydraulics. For example, assume that we have a system pressure drop of 50 kPa. Based on taking 33% of the total system drop at design flow, excluding the valve, a pressure balance results in the control valve taking 25% 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% 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 increase of 20% 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 [24] with the results given in Equation 2.8. In this approach, the pressure drop across a control valve can be estimated:

$$\Delta P_{\mathrm{cv}} = 0.05\, P_{\mathrm{s}} + 1.1\left[\left(\frac{Q_{\mathrm{m}}}{Q_{\mathrm{d}}}\right)^{2} - 1\right]\Delta P_{\mathrm{f}} + \Delta P_{\mathrm{b}}, \tag{2.8}$$

where ΔP_{cv} is the pressure drop across a control valve (kPa), P_{s} is the upstream or supply pressure (kPa), Q_{m} is the maximum anticipated flow rate, Q_{d} is the design flow rate, ΔP_{f} is the friction pressure drop at the design flow rate (kPa) and ΔP_{b} is the base (minimum) control valve pressure drop (kPa).

The first term accounts for a fall-off in overall system pressure drop by using 5% 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 [24], given in Table 2.2.

Table 2.2 *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

It should be noted that the values for butterfly and ball valves in Table 2.2 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, for example, 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 maximum flow, not just at design flow or an arbitrary multiple thereof. However, conditions at design flow can also be incorporated and helpful. Note that the procedures use the equations developed by Fisher [18, 19] 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 [17, 20])
- Noise tolerance (ANSI/IEC Standard [17, 20])

Currently manufacturers worldwide are implementing the IEC Valve Sizing Code [20]. This is a procedure that allows tighter noise prediction, particularly in gas service.

Liquid Control Valve Sizing

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

$$C_v = \frac{Q}{\sqrt{\Delta P / \mathrm{SG}}}. \tag{2.9}$$

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

A plot (Figure 2.28) of Equation 2.4 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

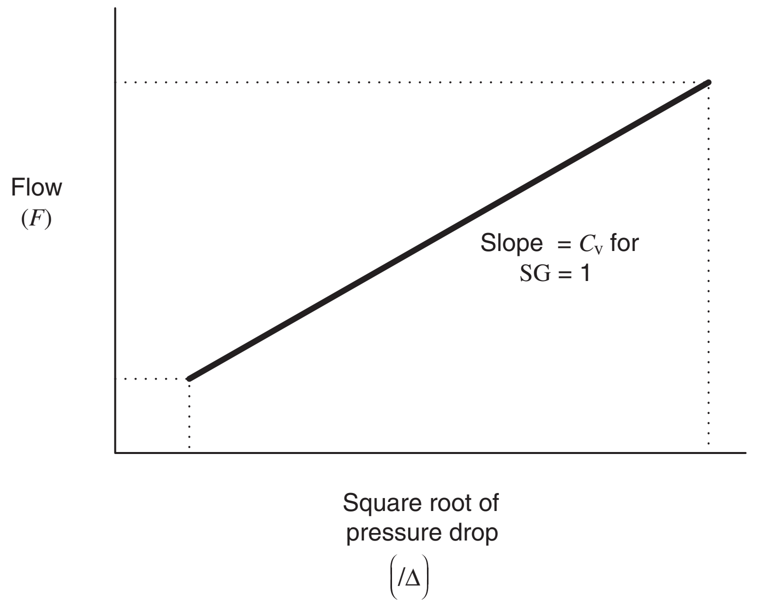

Figure 2.28 *Flow versus square root of the pressure drop across a control valve.*

a pressure drop as shown in Figure 2.29. 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 vaporizes. 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 realized, according to the continuity equation.

The issue becomes how choked flow is integrated into liquid valve sizing. Fisher Controls [18, 19] 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 2.30.

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 an expression for ΔP_{allow} that are functions of vapour and critical pressures. Overall, this results in the following liquid sizing procedure.

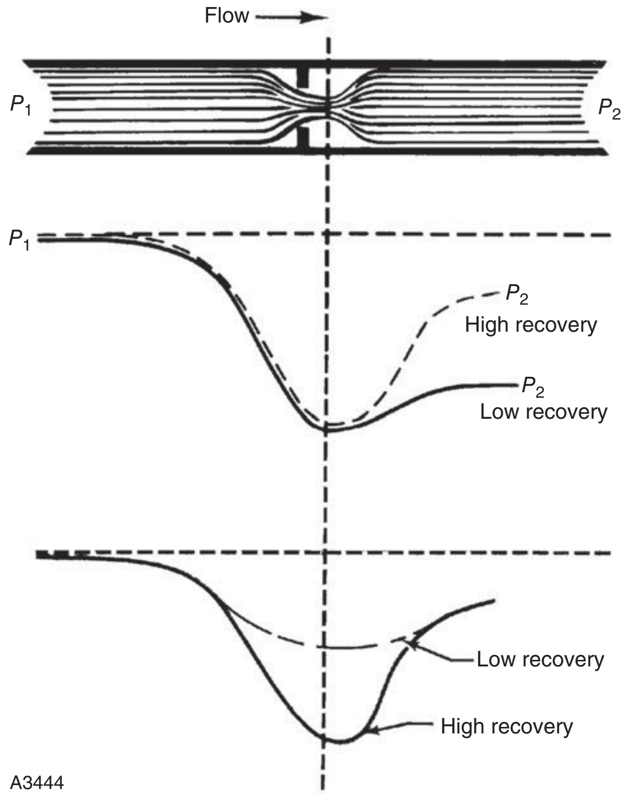

Figure 2.29 Pressure profiles across a control valve (Reproduced by permission of Emerson Process Management).

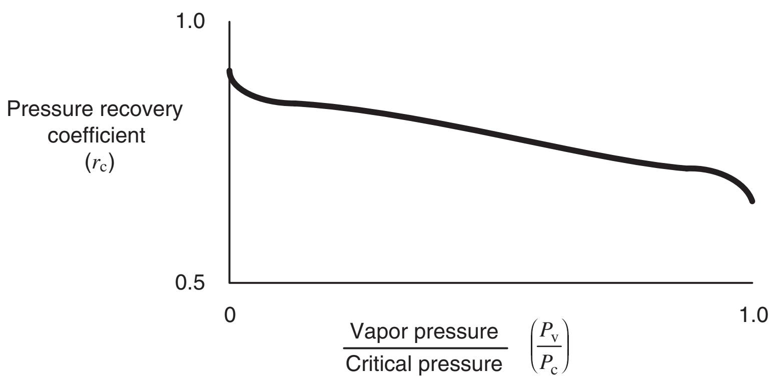

Figure 2.30 r_c versus P_v/P_c.

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% open, the valve is adequate; otherwise a larger valve is required. Note that the design C_v should be at about 50–60% 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 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 and
- 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 while the other parameter represents the type of valve and its effect on critical flow. The Fisher equations are as follows:

$$C_g = \frac{Q_{scfh}}{\sqrt{\dfrac{520}{SGT}}P_1 \sin\left[\left(\dfrac{C_a}{C_1}\right)\sqrt{\dfrac{\Delta P}{P_1}}\right]} \quad \text{(Ideal Gas)}, \tag{2.12}$$

$$C_g = \frac{W_{lb/hr}}{1.06\sqrt{\rho_1 P_1}\sin\left[\left(\dfrac{C_a}{C_1}\right)\sqrt{\dfrac{\Delta P}{P_1}}\right]} \quad \text{(Non-ideal Gas)}, \tag{2.13}$$

$$C_s = \frac{W_{lb/hr}(1 + 0.00065T_{SH})}{P_1 \sin\left[\left(\dfrac{C_a}{C_1}\right)\sqrt{\dfrac{\Delta P}{P_1}}\right]} \quad \text{(Steam)}. \tag{2.14}$$

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:

$$C_v = \frac{Q_{scfh}}{\sqrt{\frac{520}{SGT}} P_1 C_a \sqrt{\frac{\Delta P}{P_1}}}. \tag{2.15}$$

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

$$C_g = \frac{Q_{scfh}}{\sqrt{\frac{520}{SGT}} P_1} \tag{2.16}$$

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

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 falls in the range of 20–80% open, the valve is adequate; otherwise a larger valve is required. Note that the design C_g or C_s should be at about 50–60% 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. 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, it will partially vaporize. If the pressure recovers above the vapour pressure, the gas bubbles collapse on the metal and tend to break it away in small pieces. This is known as cavitation (Figure 2.31). 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 vapour pressure, flashing occurs (Figure 2.32) which can erode the valve plug and seat. Fisher uses a similar equation to Equation 2.18 to describe cavitation pressure drop:

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

Figure 2.31 Typical appearance of flashing damage (Reproduced by permission of Emerson Process Management).

Figure 2.32 Typical appearance of cavitation damage (Reproduced by permission of Emerson Process Management).

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 2.3.

Other control valve vendors use a similar equation.

Control valves can be designed to prevent cavitation or at least minimize 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 (Figure 2.33), 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, and not the control valve, because the outlet pressure is below the vapour pressure of the liquid. However, flashing damage can be minimized by using specially designed valve trims.

Table 2.3 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

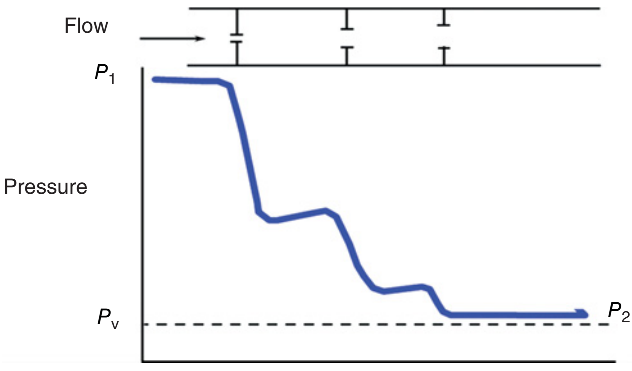

Figure 2.33 *Multistage pressure drop (Reproduced by permission of Emerson Process Management).*

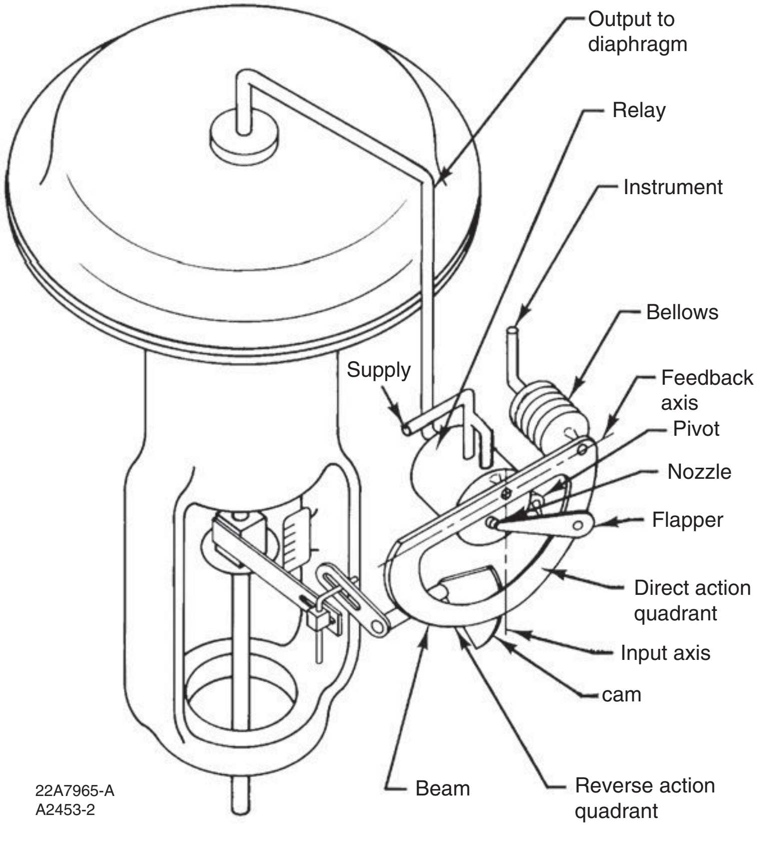

Figure 2.34 *Pneumatic positioner schematic for diaphragm actuator (Reproduced by permission of Emerson Process Management).*

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 2.34 illustrates the function of a pneumatic positioner for a diaphragm actuator, while Figure 2.35 shows a modern control valve utilizing 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; and
- split range operation, which is when two or more valves are operated by one controller.

Figure 2.35 *Modern control valve and digital valve positioner (Reproduced by permission of Emerson Process Management).*

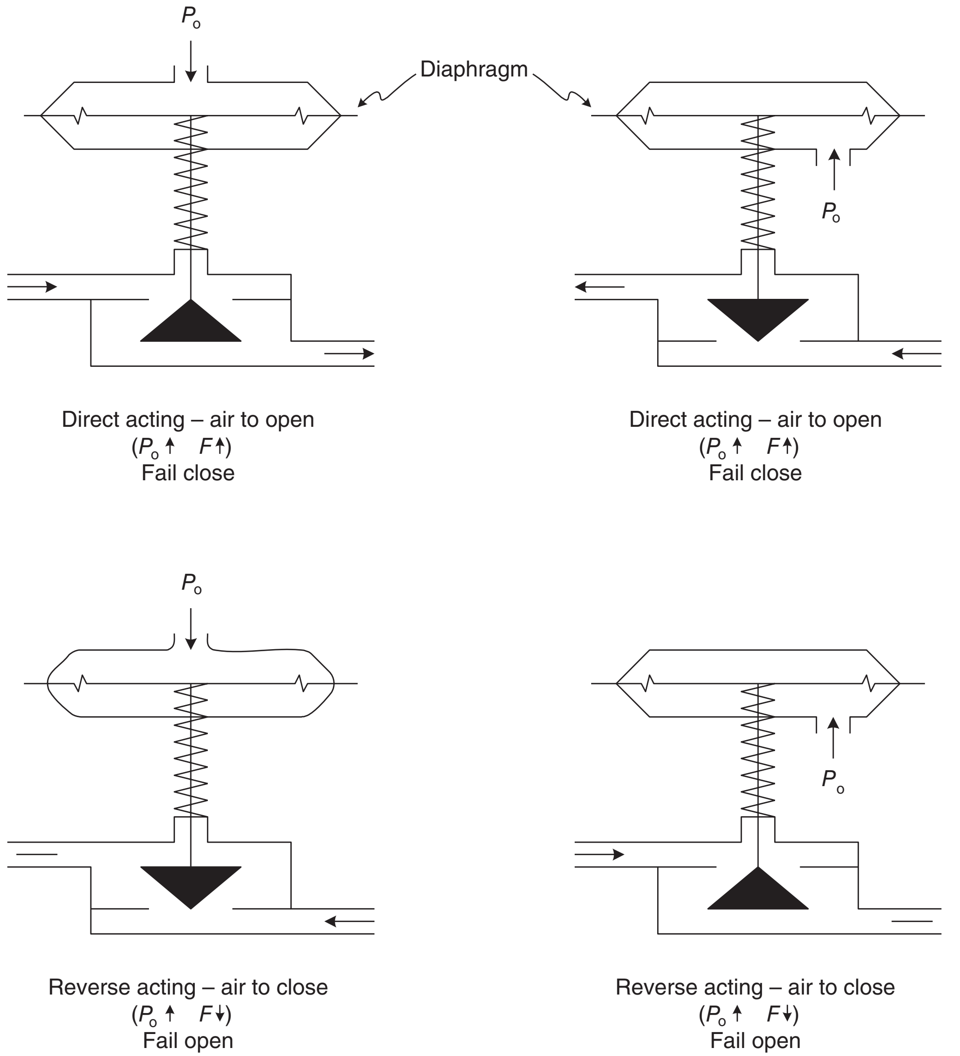

Figure 2.36 *Fail-safe design of valves.*

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 [25].

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 2.36 illustrates these fail-safe designs.

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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 characterization of system responses and provide an introduction to modelling various processes. After studying this chapter, the reader should understand

- the basic components of a 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; and
- 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 the engine itself that converts combustion energy into rotating mechanical energy that turns the wheels at a certain *rpm*. Consider a car on a straight, flat road on a still day. Move the throttle to just the right

A Real-Time Approach to Process Control, Third Edition. William Y. Svrcek, Donald P. Mahoney and Brent R. Young.

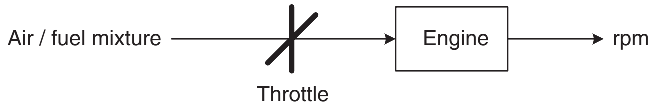

Figure 3.1 *Illustration of car metaphor.*

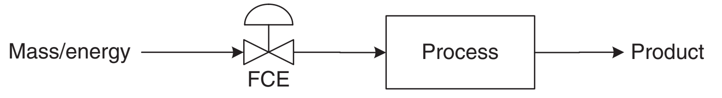

Figure 3.2 *A simplified process view.*

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 rpm 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 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 rpm, we call the measured output of the process, the *controlled variable/process variable.*

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 is suddenly rising 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

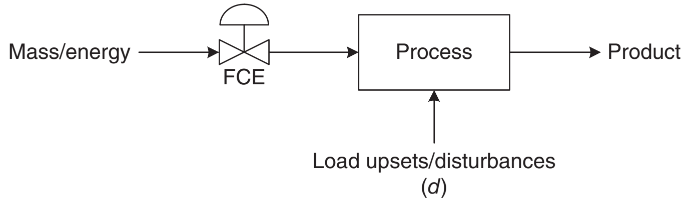

Figure 3.3 *A realistic process view.*

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 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 primarily focus 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 rpm, we need to be able to constantly adjust the throttle position 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. Existing control practitioners should recognize this figure is an alternative form of a classical block flow diagram (Figure 3.5) that is seen in most process

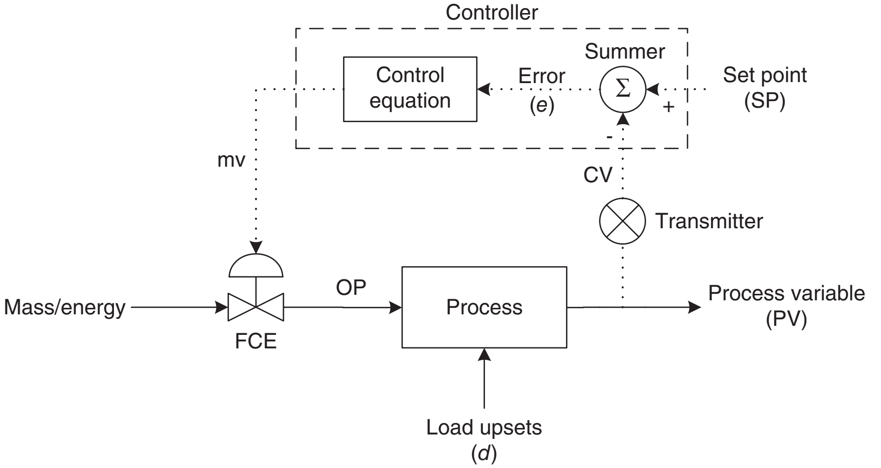

Figure 3.4 *Single-input/single-output feedback control loop.*

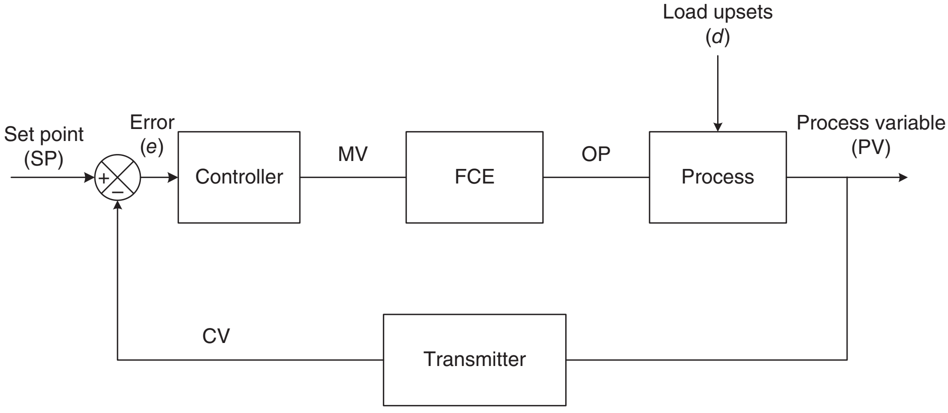

Figure 3.5 *Alternative, classical form of the single input/single output feedback control loop.*

control textbooks. Both Figures 3.4 and 3.5 illustrate the same things – both the information flow and physical connections of the feedback control loop. (In this book we use a format that more closely resembles instrumentation or control systems diagrams that are used in real process design documentation, e.g. P&IDs and control systems narratives).

In the feedback control loop 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 to 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*, or 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 successfully control a process, it is important to select both the right process variable and the 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 3.6 [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

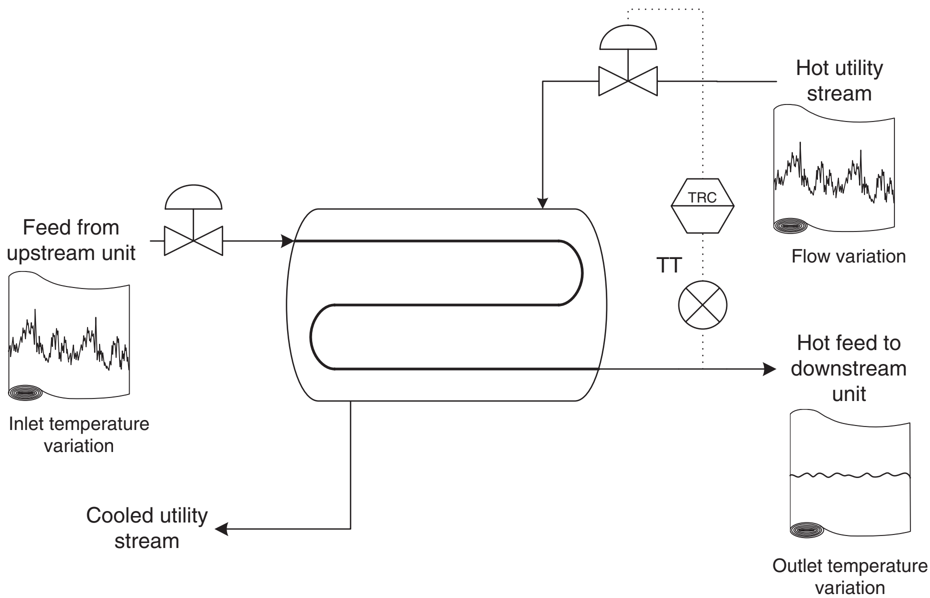

Figure 3.6 *Transformation of variation from temperature to flow (Adapted from [4]).*

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, and, therefore, require some sort of control. Automatic feedback control is the most common form of control.
- 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 represents the magnitude 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; and
  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 to 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 feedbacks.

3.4.1 Positive and Negative Feedbacks

Positive feedback represents a controller contribution that reinforces the error, and therefore 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, and therefore tends to add to stability. The cruise control in our automobile example works with negative feedback. If the speed is too high, 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's 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 rooms decreases. This is known as an increase/decrease (I/D) relationship, or a reverse-acting element.

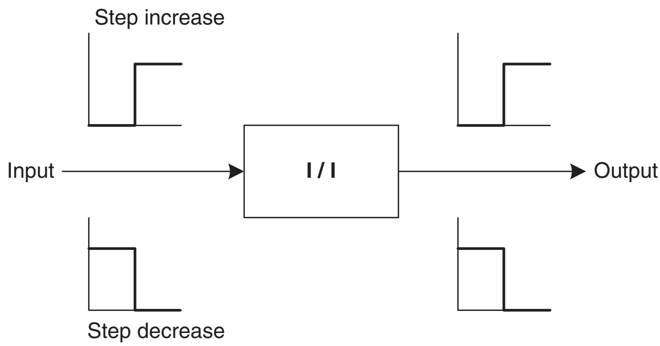

Figure 3.7 *Increase/Increase component action.*

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 3.7. Ignoring the relative amplitudes between input and output, if there is an increasing or decreasing input there will be a corresponding increasing or decreasing output.

Consider a general I/D component as shown in Figure 3.8. For this element, if there is an increasing or decreasing signal, a resulting decreasing or increasing output signal will result.

Connecting several I/I components in series, as shown in Figure 3.9, will result in an overall I/I action.

As seen in Figure 3.10, if a single I/D component is placed anywhere in the sequence the overall action is I/D.

Figure 3.11 shows that when two I/D blocks are in series there is an overall I/I action. It can further be shown that whenever there is 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 3.12) 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

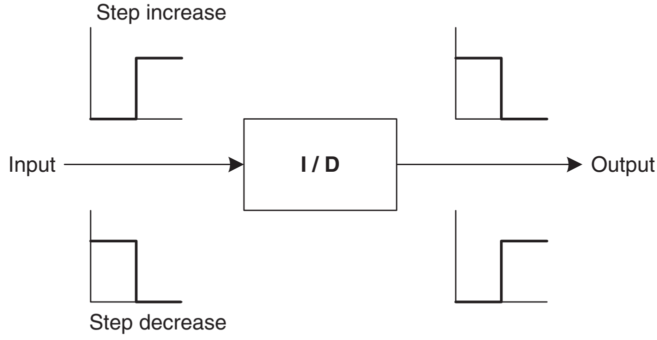

Figure 3.8 *Increase/Decrease component action.*

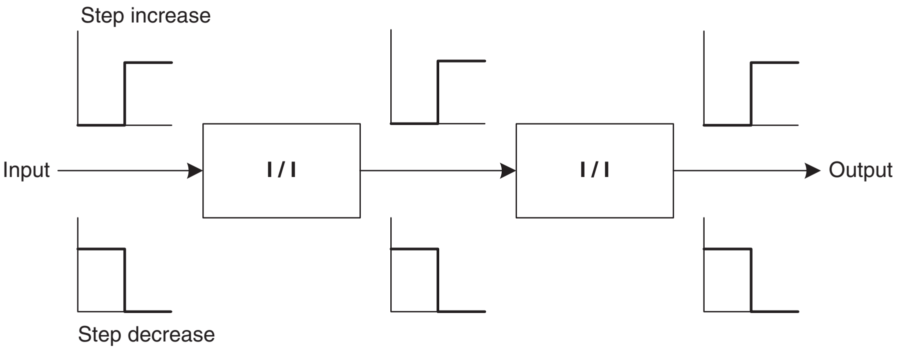

Figure 3.9 *Increase/Increase components in series.*

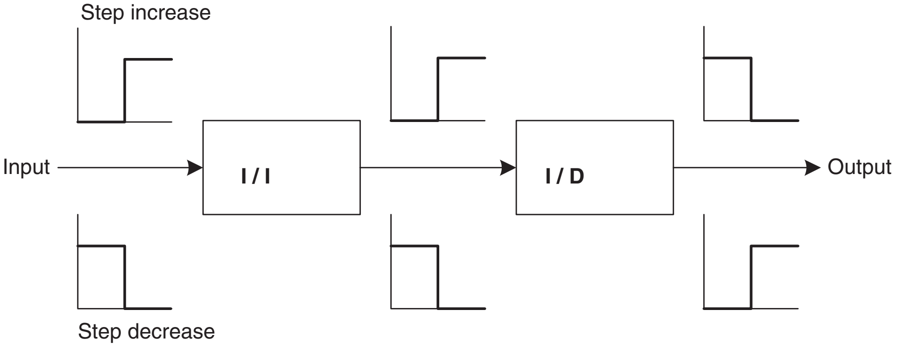

Figure 3.10 *Combined components in series.*

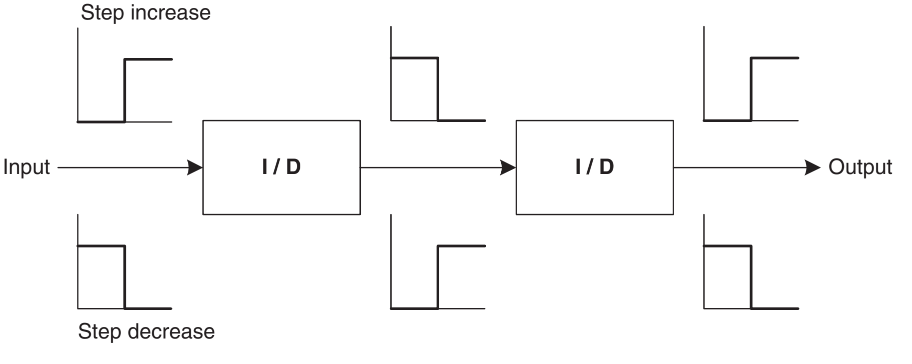

Figure 3.11 *Increase/Decrease components in series.*

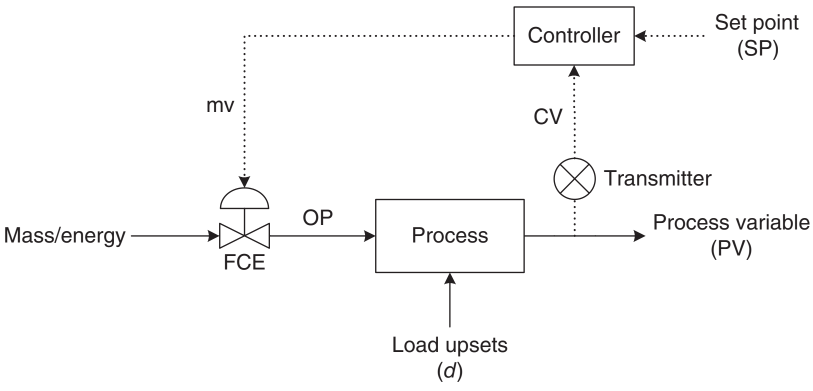

Figure 3.12 *Typical single input single output (SISO) loop.*

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's 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 processes and determine the relationships between the sign of their inputs and outputs.

The first process is a single tank shown in Figure 3.13. 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.

Now consider an energy flow or heating process illustrated in Figure 3.14. 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 3.15. We assume that the feed and the 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.

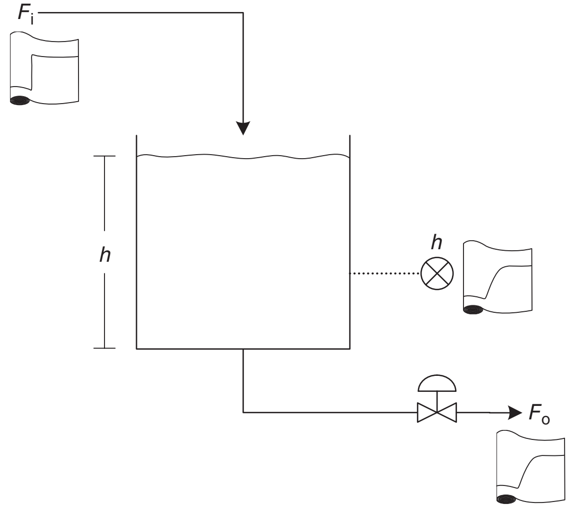

Figure 3.13 *Mass flow process.*

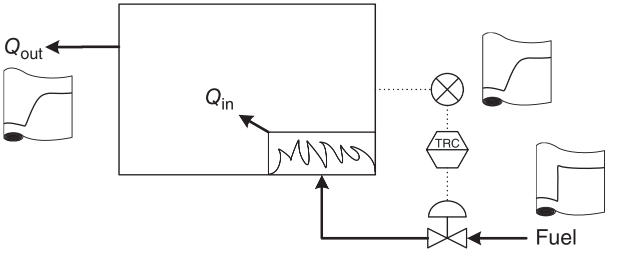

Figure 3.14 *Energy flow process.*

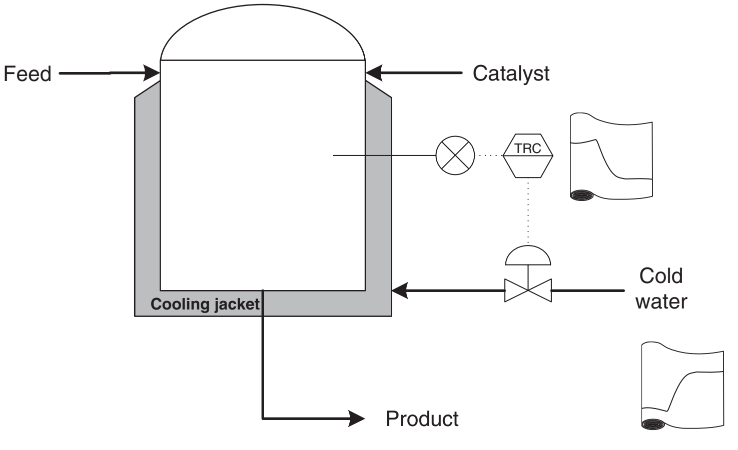

Figure 3.15 *Exothermic reactor.*

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, and so on. However, in the fluid-processing industries about 90% of all FCEs are valves. Hence, it is necessary to understand the action of a control valve. For a manual control valve, as the stem 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 close), 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 close' 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 3.16, assume the action shown in Figure 3.17, 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 3.18. In this situation, the controller must be set to direct action in order to achieve the overall negative feedback required for the loop.

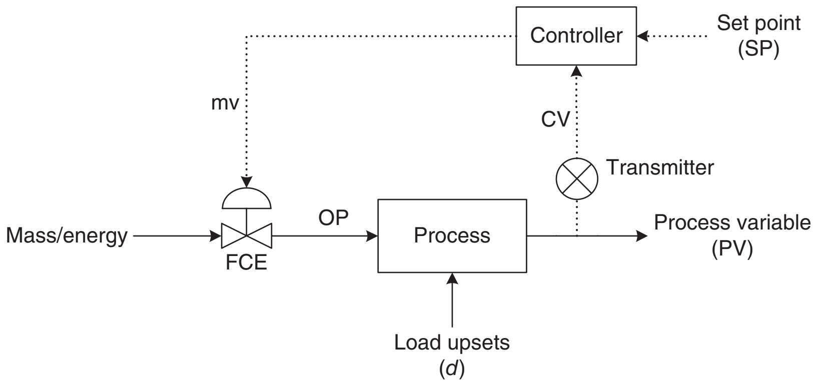

Figure 3.16 *SISO feedback control loop.*

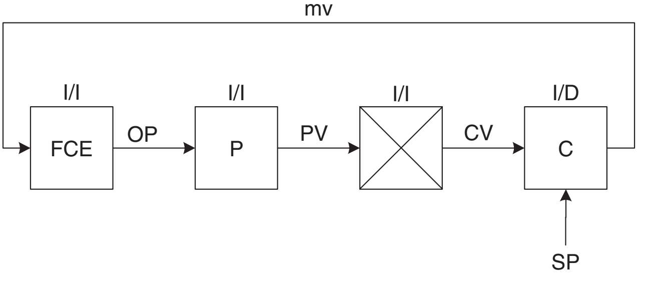

Figure 3.17 *Component input/output for air-to-open actuator.*

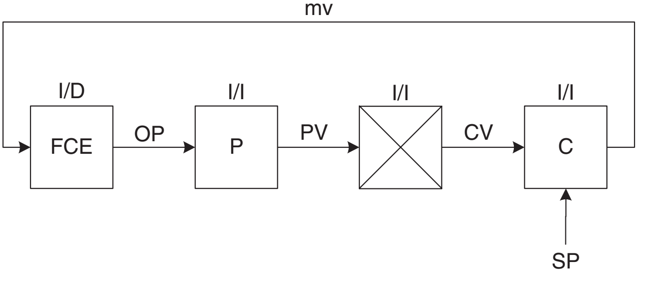

Figure 3.18 *Component input/output for air-to-close actuator.*

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 this complexity, many processes behave as if they were first-order systems (i.e. systems that may be described by first-order differential equations – more on that in Section 3.7), many

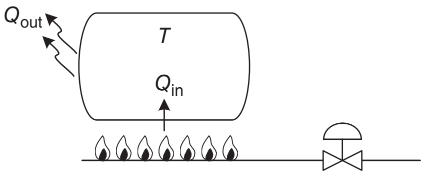

Figure 3.19 *Capacity dominated process – energy storage.*

exhibiting transport delay or dead time. Because of this, it is important to understand two fundamental process dynamic characteristics: capacitance and dead time.

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, that is, inertia. A common example of a capacity-dominant process is one that stores energy (Figure 3.19).

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 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 3.20: 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 in the tank, H, if the inflow, F_i, were increased. One would certainly expect the level to rise. However, if F_i was increased by 10%, the level would not

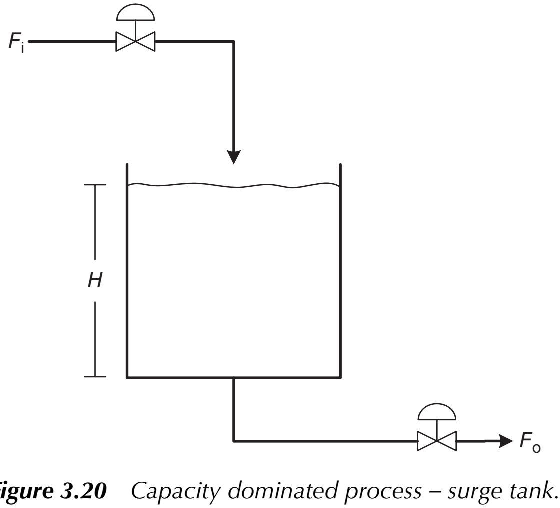

Figure 3.20 *Capacity dominated process – surge tank.*

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):

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

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

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 Equation 3.3:

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

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 Equation 3.4:

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

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 valves. The symbol g is the acceleration of gravity. Substituting for Q_{out} in Equation 3.3, which is a first-order differential equation, would result in a non-linear 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 the first-order differential equation can be solved analytically for a disturbance in flow into the tank. This linearization for the purpose of analytical solution represents a simplification of the process dynamics. However, fortuitously most processes can be well approximated as being linear close to their operating set points. If we have designed an adequate control system, this assumption holds. Therefore we will proceed to linearize Equation 3.4 and solve it analytically. For a single variable the Taylor series can be written as shown in Equation 3.5:

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

In performing a first-order linearization, shown in Figure 3.21, the higher order terms (HOT) are neglected since $h = (H - H_{\text{o}})$ is small. Setting $F(H)$ (Equation 3.6) as a

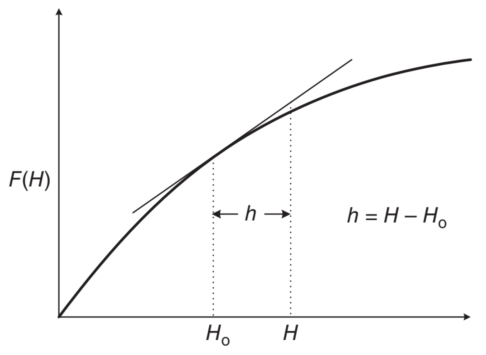

Figure 3.21 *First-order linearization.*

function of head, H, and substituting in Equation 3.5 results in the linear form shown in Equation 3.10:

Let

$$F(H) = V(Xp)\,C_{\mathrm{V}}\sqrt{\rho\,g}\sqrt{H} = K_{\mathrm{V}}\sqrt{H}, \tag{3.6}$$

$$F(H) = F(H_{\mathrm{o}}) + \left(\frac{\partial F}{\partial H}\right)_{H_{\mathrm{o}}}(H - H_{\mathrm{o}}), \tag{3.7}$$

$$\left(\frac{\partial F}{\partial H}\right)_{H_{\mathrm{o}}} = \frac{1}{2}\left(\frac{K_{\mathrm{V}}}{\sqrt{H_{\mathrm{o}}}}\right), \tag{3.8}$$

or

$$F(H) = F(H_{\mathrm{o}}) + \left(\frac{K_{\mathrm{V}}}{2\sqrt{H_{\mathrm{o}}}}\right)(H - H_{\mathrm{o}}), \tag{3.9}$$

or

$$F(H) - F(H_{\mathrm{o}}) = \left(\frac{K_{\mathrm{V}}}{2\sqrt{H_{\mathrm{o}}}}\right)(H - H_{\mathrm{o}}). \tag{3.10}$$

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

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

At the initial or steady-state condition, Equation 3.11 can be written as Equation 3.12:

$$A\frac{\mathrm{d}H_{\mathrm{o}}}{\mathrm{d}t} = Q_{\mathrm{in}_{\mathrm{o}}} - F(H_{\mathrm{o}}). \tag{3.12}$$

Subtracting the above two equations results in Equation 3.13:

$$A\frac{\mathrm{d}(H_{\mathrm{o}} - H)}{\mathrm{d}t} = Q_{\mathrm{in}_{\mathrm{o}}} - Q_{\mathrm{in}} - [F(H_{\mathrm{o}}) - F(H)]. \tag{3.13}$$

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

Equation 3.15:

$$A\frac{\mathrm{d}h}{\mathrm{d}t} = q_{\mathrm{in}} - \frac{K_{\mathrm{V}}}{2\sqrt{H_{\mathrm{o}}}}h, \tag{3.14}$$

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

In Equation 3.15, R is the resistance to flow in units of time/length^2 (min/m^2). Equation 3.15 can be written in the standard form for a first-order LDE (Equation 3.16):

$$RA\frac{\mathrm{d}h}{\mathrm{d}t} + h = R\,q_{\mathrm{in}}, \tag{3.16}$$

where $RA = \tau =$ 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 Equation 3.19:

$$\frac{\mathrm{d}h}{\mathrm{d}t} + \frac{h}{\tau} = K\,q_{\mathrm{in}}, \tag{3.19}$$

where $K = 1/A$.

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

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

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'll find out, capacitance can be a control engineer's best friend. A capacity-dominated system is described by Equation 3.21:

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

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:

$$\Delta\text{mv} = K_c(\text{SP} - \text{PV}) = K_c e, \tag{3.22}$$

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 the K_c, 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, that is, there is no 100% 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's 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. Recognize 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, 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's 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, when you turn the tap, it takes several seconds for you to feel the effect. Assuming you've been lucky enough to get the water to a comfortable temperature, there's always the inconsiderate or unaware houseguest who flushes the toilet mid shower. Quickly, you race to cut back on the scalding hot water source. There is no immediate effect and

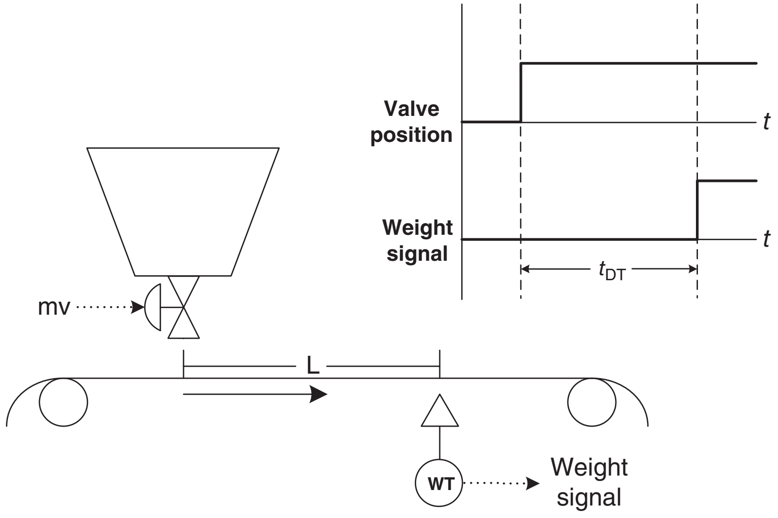

Figure 3.22 *Continuous weighing system.*

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 realize 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's 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 3.22.

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 in Equation 3.23:

$$t_{\mathrm{DT}} = \frac{L}{v}[=]\left[\frac{\mathrm{m}}{\mathrm{m/min}}\right][=][\mathrm{min}]. \tag{3.23}$$

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, 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 of 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.

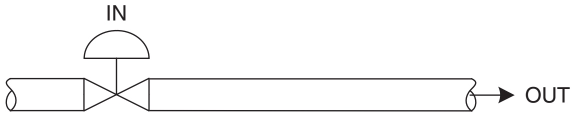

Figure 3.23 *Valve/pipe flow system.*

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 following scenario shown in Figure 3.23.

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 per cent of valve span, $A\%$, an increase in flow of B m^3/s 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:

$$K_{ss} = \frac{\Delta\text{out}}{\Delta\text{in}} = \frac{B(\text{m}^3/\text{s})}{A(\%)}. \tag{3.25}$$

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, that is, the pipe alone, $K_{ss} = 1$. 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's look at why.

A system with capacitance and dead time (actually, quite a common combination) is described by Equation 3.26:

$$h(t) = Kq_{\text{in}}\left(1 - e^{-(t-\text{DT})/\tau}\right). \tag{3.26}$$

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:

$$\Delta\text{mv} = K_c(\text{SP} - \text{PV}) = K_c e, \tag{3.27}$$

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 over-react. 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 non-linear process response) that keeps control engineers earn 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 minutes of dead time in a chemical process with long response times (large capacitance) will not cause too much trouble. What about 2 minutes of dead time in the control loops of a jet airplane?

Let's summarize what we've just covered:

- Two types of feedback exist: *positive* and *negative* feedbacks. Only negative feedback produces stable control.
- Each element in a control loop has a particular action or sign relationship between its input and output responses. 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, 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

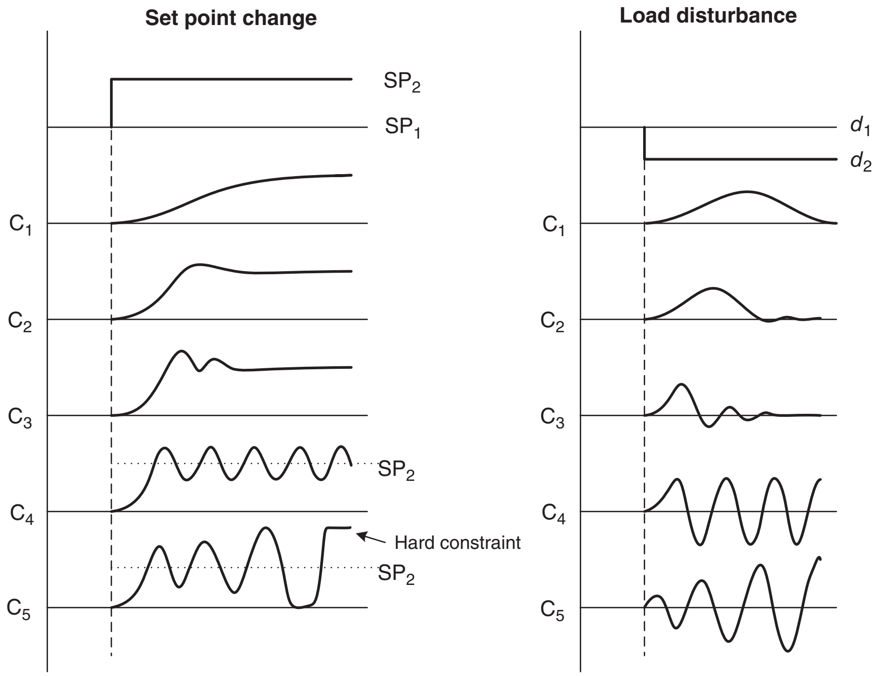

Figure 3.24 *Typical PV response to set point and load disturbance upsets.*

understanding is important. In fact, the more we understand about the process, the easier our overall control system design may become. We'll touch on this more in Chapter 10 on plant-wide control. For now, let's 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'll use the set point change disturbance and the load disturbance as a means of illustrating process dynamic response.

As qualitatively illustrated in Figure 3.24, 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 above would be classed as overdamped, that is, a slow, sluggish return to set point. C_2 presents the case for critical damping, that is, 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, that is, 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 responses are characteristic of specific types of processes. 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 five time constants, 5τ, for the output of the capacity process to reach its final, steady-state value. A time constant, τ, is defined as the amount of time it takes the output of the system to reach 63.2% 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 (see Equation 3.28):

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

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, that is, seconds, minutes and so on.

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 3.25.

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, non-linear differential equations result when the system is

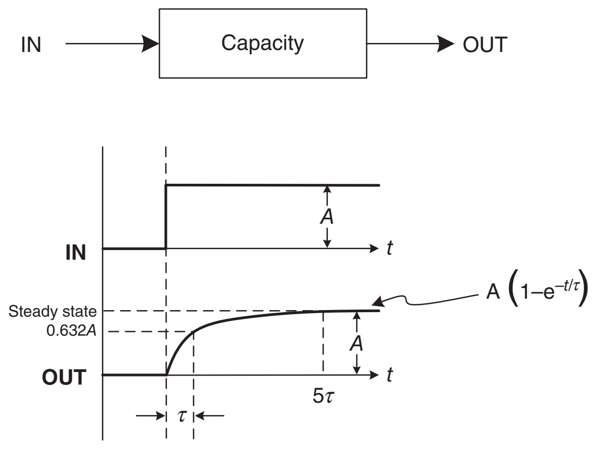

Figure 3.25 *First order system response to a step input.*

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 3.26.

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

$$\sum F_\mathrm{i} = Ma. \tag{3.29}$$

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

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

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

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

The force is a function of time 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:

$$W\frac{\mathrm{d}^2X_\mathrm{v}}{\mathrm{d}t^2} + b\frac{\mathrm{d}X_\mathrm{v}}{\mathrm{d}t} + KX_\mathrm{v} = AP(t), \tag{3.33}$$

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 better understand what type of response these second-order systems will display we will examine another common system and

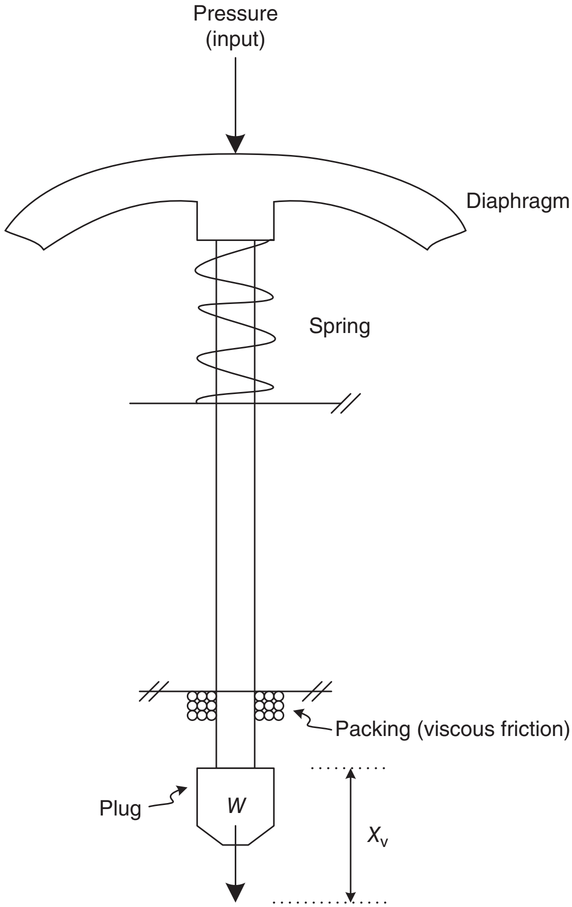

Figure 3.26 *Diaphragm operated control valve.*

generalize 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 3.27. Note that this is really just a simplification of the control valve example just cited.

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

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

Since the rest, or equilibrium, position Y_R is constant, Equation 3.34 can be rewritten in terms of the displacement from the rest position, y. In this manner, we will be looking at variations about the equilibrium position, that is, steady state. This is a common approach

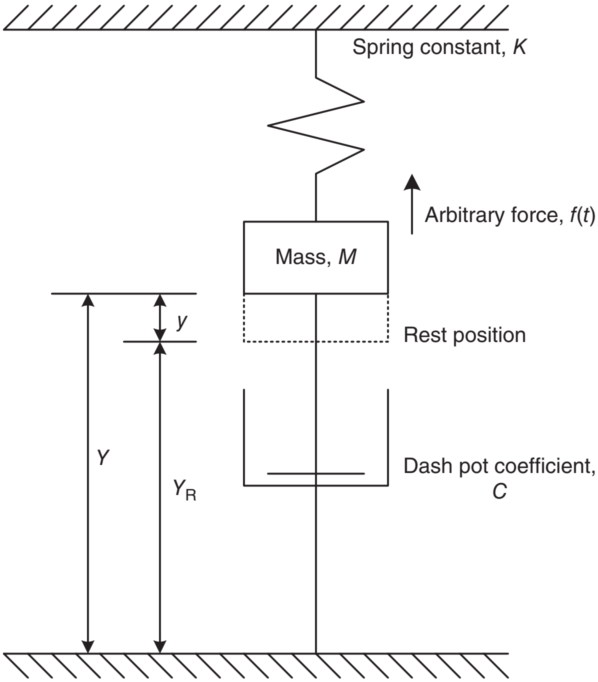

Figure 3.27 *Spring, mass and dash pot (damper) system.*

in system analysis because analysis of even non-linear systems about a steady state results in a linear system, that is, ordinary differential equations (ODEs):

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

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

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

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

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

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 ODE 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:

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

$$\frac{\mathrm{d}[Y(0)]}{\mathrm{d}t} = 0. \tag{3.40}$$

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 in Equation 3.41:

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

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 as follows:

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

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 3.38 by M:

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

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

$$\omega_\mathrm{n} = \sqrt{\frac{K}{M}}, \tag{3.45}$$

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

These generalized terms then characterize the response of the system. The first term, ω_n, is called the undamped natural frequency, while 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 as the control scheme for a real chemical process is undertaken. Remember that

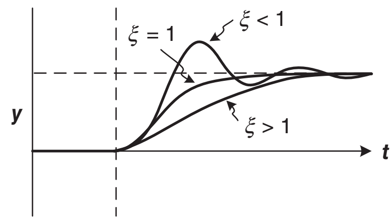

Figure 3.28 *Typical second order response.*

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 characteristic equation, shown in Equation 3.47:

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

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

which simplifies to

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

or

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

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 3.28.

Now let us examine the case of multiple capacities in series. Consider two non-interacting tanks in series, shown in Figure 3.29.

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

$$A_1\frac{\mathrm{d}h_1}{\mathrm{d}t} = F_i - F_1, \tag{3.51}$$

$$A_2\frac{\mathrm{d}h_2}{\mathrm{d}t} = F_1 - F_2. \tag{3.52}$$

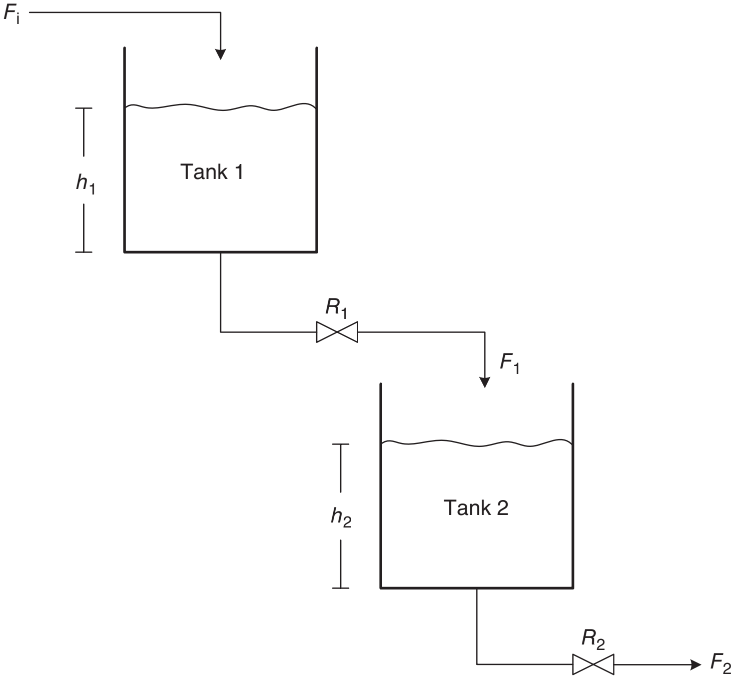

Figure 3.29 *Two non-interacting tanks in series.*

If linear resistance to flow is assumed for the valves in the system 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

$$A_1 R_1 \frac{\mathrm{d}h_1}{\mathrm{d}t} + h_1 = F_i R_1, \tag{3.55}$$

$$A_2 R_2 \frac{\mathrm{d}h_2}{\mathrm{d}t} + h_2 = \frac{R_2}{R_1} h_1, \tag{3.56}$$

Differentiating Equation 3.56 with respect to time and re-writing h_1 in terms of h_2 gives a second-order ODE:

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

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

$$2\xi\omega_n = \frac{1}{A_2R_2} + \frac{1}{A_1R_1}, \tag{3.58}$$

$$\omega_n = \sqrt{\frac{1}{A_1A_2R_1R_2}}. \tag{3.59}$$

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 tanks display?

3.7.3 Simple System Analysis

Often questions 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, 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 non-linearities 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 3.30.

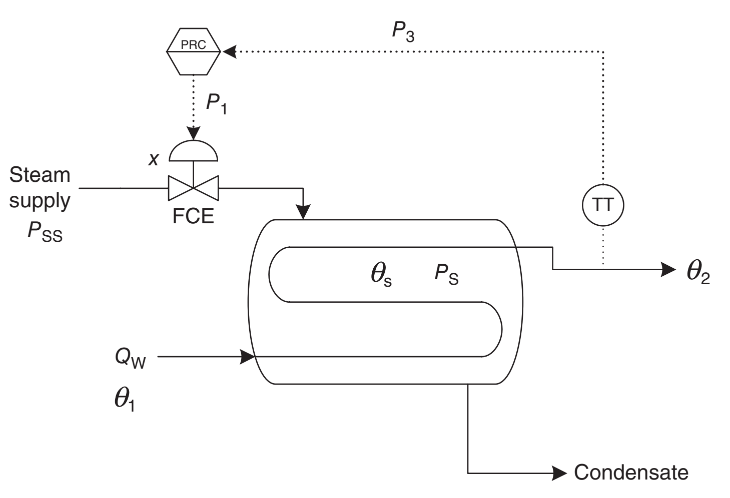

Figure 3.30 *Typical process schematic of shell and tube heat exchanger.*

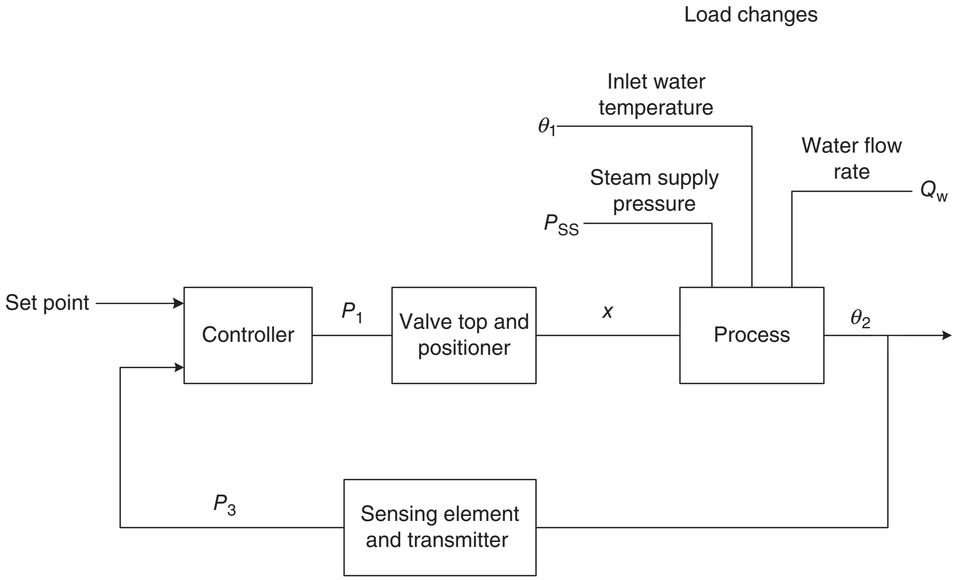

Figure 3.31 *Word block diagram of shell and tube heat exchanger.*

Constructing a Word Block Diagram

Before starting to analyse the process, it is helpful to construct a system block diagram (Figure 3.31). 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)

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 and the time constant is $\tau = 0.033$ minute. Therefore, we can write Equation 3.60 as follows:

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

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 under 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.

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 is the thermal resistance (°C/kW), A is the surface area (m^2) and h is the film coefficient (kW/°C m^2).

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

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

where C_t is the thermal capacitance (kJ/°C), W is the mass of sensing element (kg) and C is the specific heat (kJ/kg °C).

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:

$$\tau = R_\mathrm{t}C_\mathrm{t} = \frac{WC}{A\,h}. \tag{3.65}$$

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, h will primarily be a function of fluid velocity. Assuming that the bulb has a time constant over a limited velocity range as follows:

$$\tau = 1.15\left(\frac{1}{v^{0.58}}\right), \tag{3.66}$$

where τ is the time constant (seconds) and v is the fluid velocity (m/s).

Thus, for a fluid velocity of 0.686 m/s, the time constant of the sensing element is 0.024 minute (1.43 seconds).

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 lumped, linear system analysis approach, we assume that a local linear model about the operating set point holds and the heat exchanger is modelled as one lump (lumped parameter approach) in which a small change in the valve stem position, ΔX, and its effect on the outlet water temperature, θ_2, are 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, 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:

$$Q_s = f(X, P_s) = \left(\frac{3}{2}\right) C_v P_s = 0.00086 X P_s, \tag{3.67}$$

where Q_s is the steam flow (kg/s), X is the valve stem position or travel (mm) and P_s is the steam pressure (kPa).

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

$$\Delta Q_s = \alpha\, \Delta P_s + \beta\, \Delta X, \tag{3.68}$$

where

$$\alpha = \left(\frac{\partial Q_s}{\partial P_s}\right)_X = (0.00086)\, X_{op},$$

$$\beta = \left(\frac{\partial Q_s}{\partial X}\right)_{P_s} = (0.00086)\, P_{s_{op}}$$

and op is the operating point around which X and P_s are defined.

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

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

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

The coefficient γ can be evaluated from steam tables at the shell nominal operating pressure of 432 kPa; refer Table 3.1. As can be seen from Figure 3.32, at an operating pressure of 432 kPa, γ can be linearized to give a value of 11.0 kPa/°C.

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

$$\Delta Q_{\mathrm{s}} m = (W_{\mathrm{t}} C_{\mathrm{t}} + W_{\mathrm{s}} C_{\mathrm{s}})\, \Delta\theta_{\mathrm{s}} + h_{\mathrm{t}} A_{\mathrm{t}} \left(\Delta\theta_{\mathrm{s}} - \frac{\Delta\theta_2}{2} \right), \tag{3.70}$$

where m is the latent heat of condensing steam (kJ/kg; 2129 kJ/kg for this example), W_{t} is the tube weight (kg), C_{t} is the specific heat of the tube material (kJ/kg °C; 0.5 kJ/kg °C for this example), W_{s} is the shell weight (kg; 19.6 kg for this example), C_{s} is the specific heat of the shell material (kJ/kg °C), h_{t} is the tube side film coefficient (kW/°C m^2), A_{t} is the tube area (m^2; $h_{\mathrm{t}}A_{\mathrm{t}} = 0.376$ kJ/s °C), θ_{s} is the steam temperature (°C) and θ_2 is the outlet temperature (°C).

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, that is, the shell is well insulated.

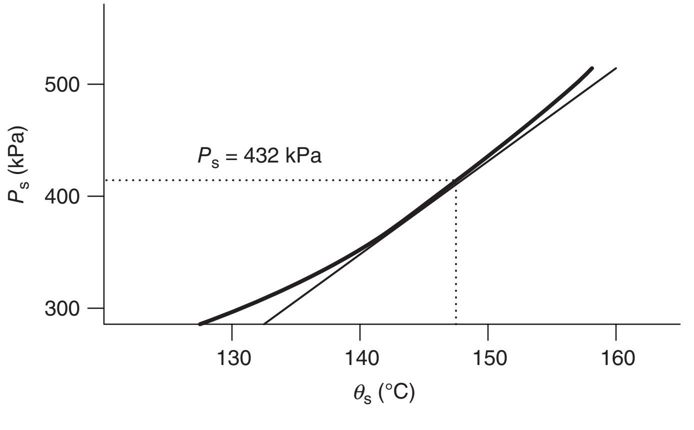

Figure 3.32 *Steam pressure versus steam temperature.*

In the development of a mathematical model the validity of assumptions is always debatable and depends on the use of the model, required accuracy, equipment size and 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:

$$W_{\mathrm{w}} C_{\mathrm{w}} \frac{\mathrm{d}\left(\Delta\theta_2/2\right)}{\mathrm{d}t} + Q_{\mathrm{w}} C_{\mathrm{w}} \Delta\theta_2 = h_{\mathrm{t}} A_{\mathrm{t}} \left(\Delta\theta_{\mathrm{s}} - \frac{\Delta\theta_2}{2}\right), \tag{3.71}$$

where Q_{w} is the water flow into the tube (kg/s), C_{w} is the specific heat of water (kJ/kg °C; 4.2 kJ/kg °C for this example), W_{w} is the water weight in the tube (kg), h_{t} is the tube-side film coefficient (kW/°C m^2), A_{t} is the tube area (m^2), θ_{s} is the steam temperature (°C) and θ_2 is the outlet temperature (°C).

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:

$$\Delta P_1 = K_{\mathrm{c}} \left[\Delta e + \frac{1}{T_{\mathrm{i}}} \int_0^t \Delta e \,\mathrm{d}t \right], \tag{3.72}$$

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 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:

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

$$\Delta Q_{\mathrm{s}} = \left(0.00086 X_{\mathrm{op}}\right) \Delta P_{\mathrm{s}} + \left(0.00086 P_{s_{\mathrm{op}}}\right) \Delta X, \tag{3.68}$$

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

$$\Delta Q_{\mathrm{s}} (2129) = (0.5 W_{\mathrm{t}} + (0.5)(19.6))\, \Delta\theta_{\mathrm{s}} + (0.376) \left(\theta_{\mathrm{s}} - \frac{\Delta\theta_2}{2}\right), \tag{3.70}$$

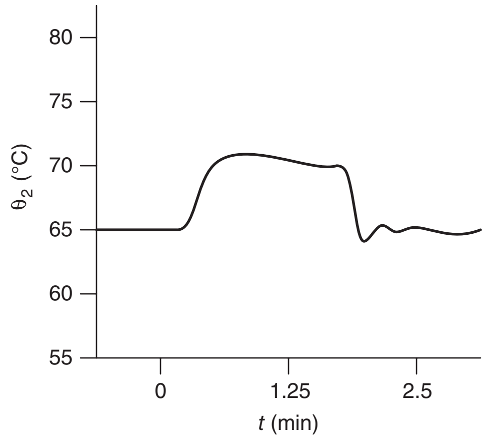

Figure 3.33 Typical response for a PI controller set point change.

$$(3.96)\,\frac{\mathrm{d}\left(\Delta\theta_2/2\right)}{\mathrm{d}t} + Q_{\mathrm{w}}\,(4.2)\,\Delta\theta_2 = (0.376)\left(\Delta\theta_{\mathrm{s}} - \frac{\Delta\theta_2}{2}\right), \tag{3.71}$$

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

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

The resulting response for this system to a PI controller set point change is shown in Figure 3.33.

The response in Figure 3.33 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 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

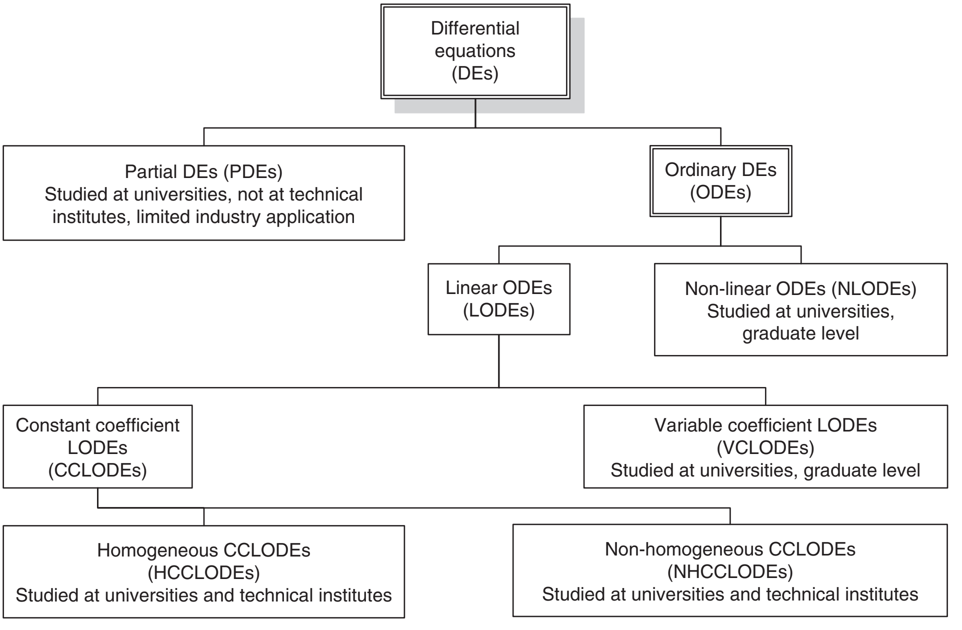

Figure 3.34 *Differential Equations.*

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 may be classified as per the taxonomy in Figure 3.34.

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

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

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 Equation 3.75:

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

An example of a physical system represented by Equation 3.75 is the linearized equation for change of level h in a cylindrical tank of cross-sectional area A, with no 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$y = y(0)e^{\left(\frac{-a_0}{a_1}t\right)}$, 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, for example, 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: $L\left\{a_1\frac{\mathrm{d}y}{\mathrm{d}t} + a_0y = 0\right\} = a_1(sY - y_0) + a_0Y = 0$

or$Y = \frac{y_0}{s+\frac{a_0}{a_1}}$, from which $\frac{\mathrm{d}y}{\mathrm{d}t} = \frac{-a_0}{a_1}y$ and$y = y(0)e^{\left(\frac{-a_0}{a_1}t\right)}$.

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 the 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 $j\omega$, where $j = \sqrt{-1}$ 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, $G(s) = 1\big/\tau s + 1$.

Table 3.2 *Transfer functions of common process systems dynamics.*

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

The Laplace variable s is replaced by $j\omega$ 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 and as 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, linearization and simplification, for example, Allen [13].

Mathematical models imbedded in today's simulation software provide a means to handle both variations in operating level and process non-linearities. 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.

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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 final control elements (FCEs) that are non-throttling in nature, that is, 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% 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% 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, that is, for the case of flow smoothing. A good example of this type of process is a surge tank. A large capacitance is important due 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. Due 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.

A Real-Time Approach to Process Control, Third Edition. William Y. Svrcek, Donald P. Mahoney and Brent R. Young.

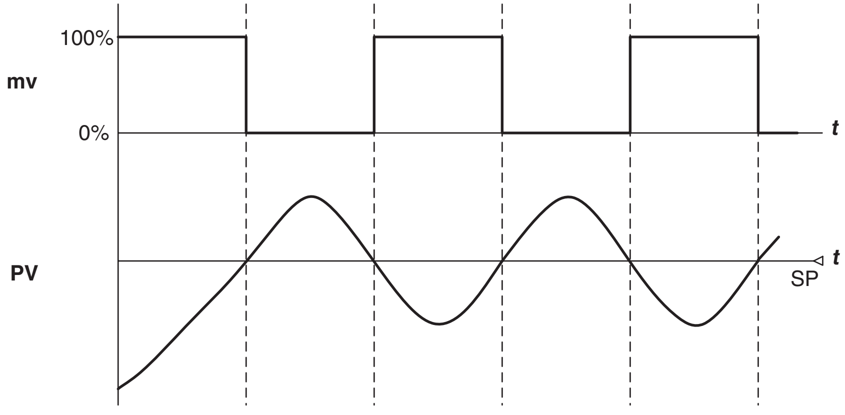

Figure 4.1 *On–off controller response.*

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

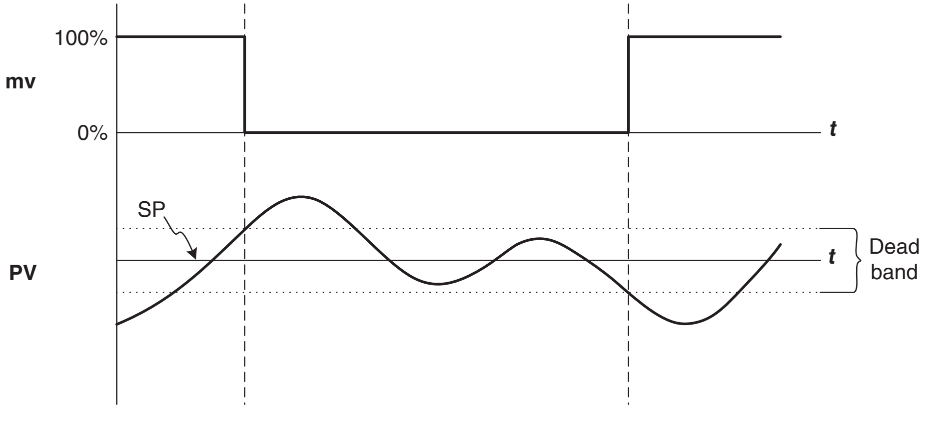

Figure 4.2 *On–off controller with dead band.*

band, inside which no control action takes place. The intent of this differential gap is to minimize the cycling of the controller output and extend the operational life of the FCE.

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 the frequency of cycling is increased but the deviation from the 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°. Setting the dead band equal to 10% of the span, the dead band in degrees would be 10°. If the set point were 75°, then the upper edge of the dead band would be $75^\circ + 5^\circ = 80^\circ$ and the lower edge of the dead band would be $75^\circ - 5^\circ = 70^\circ$, giving the dead band a width of $80^\circ - 70^\circ = 10^\circ$.

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, 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 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_o, then the level, as seen in the sight glass, remains constant. If F_o increases such that it is greater than F_i, the level will begin to drop. In order to stop the level from dropping F_i must be increased

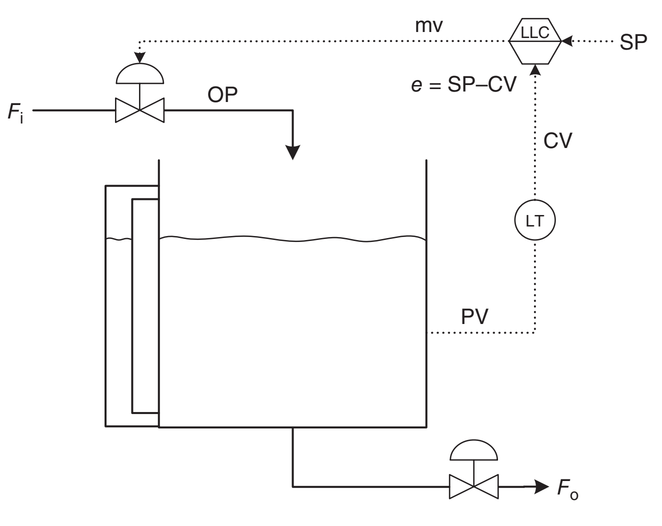

Figure 4.3 *Liquid level control – proportional mode.*

by opening the inflow valve until it is equal to F_o, 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 the level drops depends on how much the inflow valve had to move to make F_i equal to F_o. A similar situation would occur if F_o 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[1] (Equation 4.2):

$$\text{mv} = K_c e. \tag{4.2}$$

In Equation 4.2, K_c is the controller gain, e is the error and mv is the manipulated variable. Remember:

$$e = \text{SP} - \text{CV} \quad \text{(for reverse acting)},$$
$$e = \text{CV} - \text{SP} \quad \text{(for direct acting)}.$$

To allegorize 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_o, the level should stay at the set point. Also, set $F_o = 50\% = F_i$, $\text{CV} = \text{SP} = 50\%$ and $K_c = 2$. Note that here we are using percent of span units. Now, if the controller is placed into auto what will happen to the output? At the instant the controller is placed into auto, the error will be zero since CV is equal to SP, and the controller output will also be zero:

$$\text{mv} = 2(50 - 50) = 0 \quad \text{(from Equation 4.2)}.$$

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_o must equal 50% again. If a linear relationship is assumed between inflow and controller output, then for $F_i = 50\%$ we will have $\text{mv} = 50\%$.

$$\begin{aligned} \text{Since} \quad & \text{mv} = K_c, e = 2(\text{SP} - \text{CV}) = 2(50\% - \text{CV}). \\ \text{For} \quad & \text{mv} = 50\%, e = 25\%. \\ \text{Therefore,} \quad & \text{mv} = 50\% \quad \text{when CV} = 25\%. \end{aligned}$$

Thus, the controller output becomes 50% when the measurement, CV, drops by 25%, creating a 25% 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 $F_i = F_o$.

Suppose K_c was set equal to 4, giving $\text{mv} = 4e$. Now, the error would only need to be 12.5% for $\text{mv} = 4(12.5\%) = 50\%$. Logically, it would appear that the larger the controller gain, the smaller the error. In theory if K_c is set to infinity the error can be reduced to zero. The problem with this extrapolation is that the gain of the controller, 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, the loop gain will be greater than one, 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,

[1] Error is the deviation of the measurement, CV, from the set point, SP.

there is another approach to reducing the error to zero. Suppose another term was added to the proportional controller equation, as shown in Equation 4.3:

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

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 equal to 2. Also, manually adjust CV = SP = 50%, $F_i = F_o = 50\%$ and set $b = 50\%$. Now, when the controller is set to auto, 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% (Equation 4.3). Since F_o is equal to 50% and mv is also equal to 50%, the level will stay the same. In general, if the bias equals the load, mv, the error will always be zero.

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

As previously mentioned, 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, is equal to one, 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_o, the response shown in Figure 4.4 could be expected.

CV damps out with a quarter decay ratio[2] 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

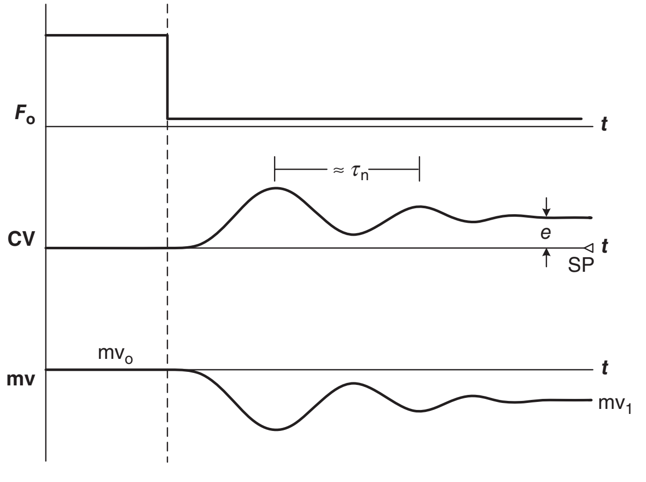

Figure 4.4 *Typical proportional-only controller response.*

[2] Quarter decay ratio is discussed in greater detail in Chapter 5.

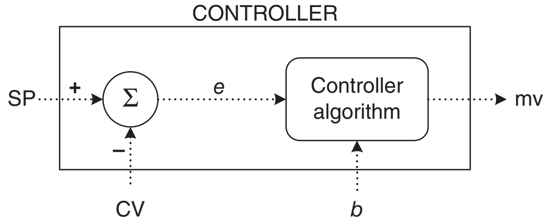

Figure 4.5 *Block diagram of proportional-only controller.*

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 Equation 4.4:

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

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 Equation 4.5:

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

Since there is a one-to-one relationship between CV and e the controller gain can be written as per Equation 4.6:

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

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, that is, % level instead of milliamperes.

Assuming that a linear relationship exists between CV and mv, as shown in Figure 4.6, Equation 4.7 may be written as follows:

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

The controller gain, K_c, in Equation 4.7, is the amount that CV must change to make the controller output change by 100%. The gain of the transmitter is similar and is given by Equation 4.8:

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

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

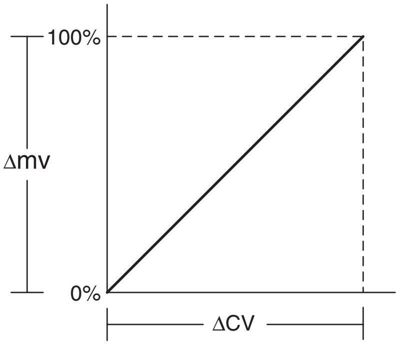

Figure 4.6 *Controller input/output relationship.*

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%. Using this definition of PB we can define the controller gain as shown in Equation 4.9:

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

If the proportional band setting on the controller is set to 40%, the output of the transmitter, which is the input to the controller, changes over 40% of its output span. The output of the controller would change by 100%, 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 $K_c = \frac{100\%}{\text{PB}\%}$, 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 follows:

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

Note that

$$e = \text{SP} - \text{CV} \quad \text{(for reverse acting)},$$
$$e = \text{CV} - \text{SP} \quad \text{(for direct acting)}.$$

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 previously discussed, is not plausible since as PB $\to 0$, $K_c \to \infty$ and the loop becomes unstable. Furthermore, it is not possible to set PB $= 0$, because on many controllers the minimum setting is usually 2–5%. However, if PB was very small, that is, 2% or $K_c = 50$, the error would certainly be minimized, provided the loop remained stable. This case can be illustrated using Figure 4.7.

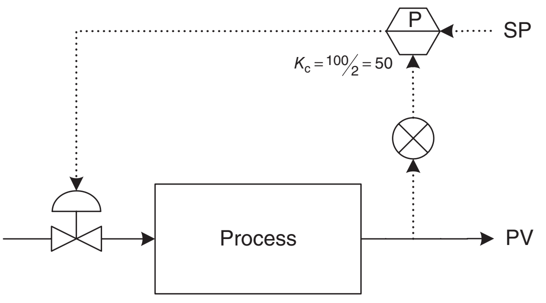

Figure 4.7 SISO feedback control loop.

If in Figure 4.7, $K_v \times K_P \times K_T < 1/50$, 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, that is, a large surge tank. Due 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 and hence make this option viable, as in Equation 4.11:

$$e = \frac{1}{K_c}(\mathrm{mv} - b). \tag{4.11}$$

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 to 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 given as

$$\mathrm{mv} = \frac{1}{T_i}\int e\,\mathrm{d}t + \mathrm{mv}_o. \tag{4.12}$$

mv_o 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.

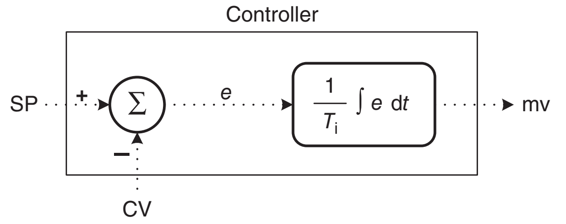

Figure 4.8 *Block diagram of integral-only controller.*

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, was increased in a stepwise 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, that is, $e = 0$.

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 the form of Equation 4.12 some manufacturers measure the reciprocal of T_i or repeats per minute in a controller (Equation 4.13):

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

As a result of this reciprocal relationship, if the controller is adjustable in min/rep, then increasing the adjustment gives less integral action, whereas in rep/min, increasing the number produces greater integral action. Therefore, it is important to be aware of how an

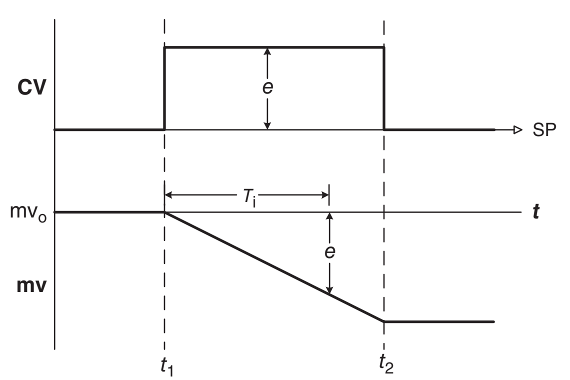

Figure 4.9 *Integral controller response to square wave input.*

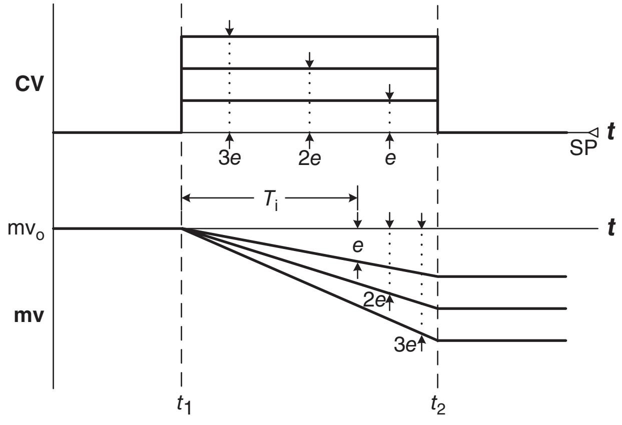

Figure 4.10 Effect of error magnitude on integral control response.

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 to proportional-only controller. As mentioned earlier, the output of the proportional-only controller changes as quickly as the measurement changes; in other words, the controller tracks the error. So, if the measurement changes as a step, the controller output also changes as a step by an amount

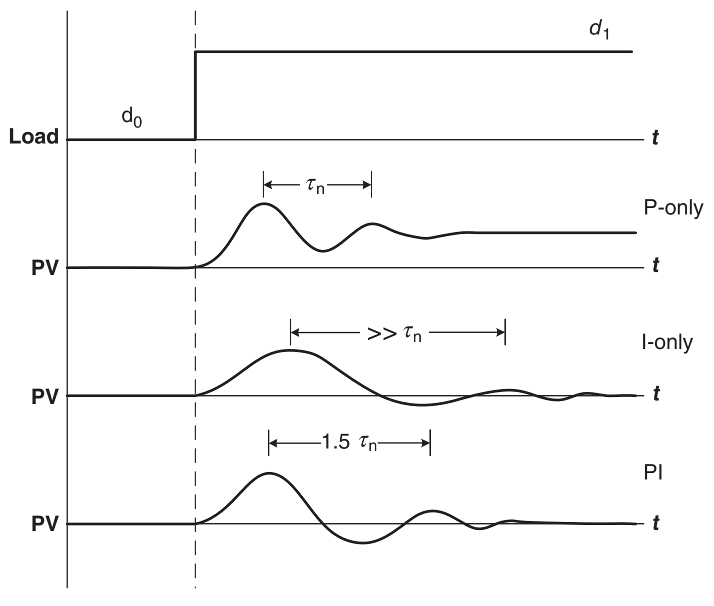

Figure 4.11 Response of P-only, I-only and PI controllers.

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 than 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 sulfur 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 sulfur plant furnace air demand. In energy or 'BTU' control of a coal-fired power station (L. Neumeister, Personal Communication) the integral-only control compares the energy leaving the boiler (steam) with the energy entering the boiler (coal). If there is a sustained difference, 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 supplied coal. Typically it takes some 12 hours for the BTU content of the coal to change a significant amount.

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% 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%) of controllers found in plants are PI controllers. The equation for a PI controller is given in Equation 4.14:

$$\mathrm{mv} = K_c\left(e + \frac{1}{T_i}\int e\,\mathrm{d}t\right) = K_c\,e + K_c\frac{1}{T_i}\int e\,\mathrm{d}t. \tag{4.14}$$

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) to 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 Equation 4.15:

$$b = K_c\frac{1}{T_i}\int e\,\mathrm{d}t. \tag{4.15}$$

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

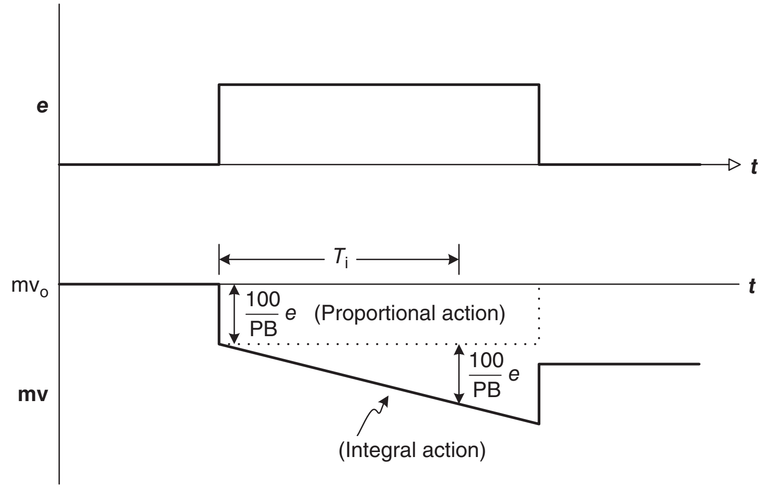

Figure 4.12 *Proportional plus integral controller response to square wave input.*

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, the gain of the PI controller, K_{PI}, 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. While $T_i = \infty$ is not realizable, it can be set to a very large number in min/rep 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 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

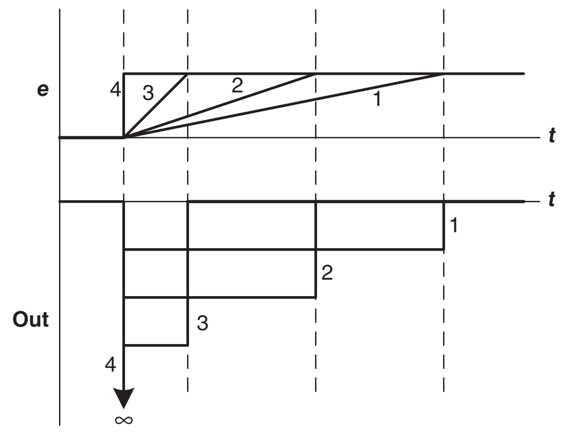

Figure 4.13 *Effect of error on derivative mode output.*

K_{PI}, which 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% 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 hours. 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, de/dt. 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 (see Equation 4.17):

$$\text{output} = T_d \frac{de}{dt}. \tag{4.17}$$

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 signals 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 to the average temperature signal, 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, 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 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 times 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 and 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 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:

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

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, $\mathrm{mv_P}$ is the proportional portion of the output and $\mathrm{mv_D}$ is the derivative portion. In the example, the measurement changes at a fixed rate of change, and 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.

Now, let us superimpose $\mathrm{mv_P}$ and $\mathrm{mv_D}$ to get the actual output for a PD controller, 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 $\mathrm{mv_D}$, or the contribution of the derivative action to the movement of the final control element.

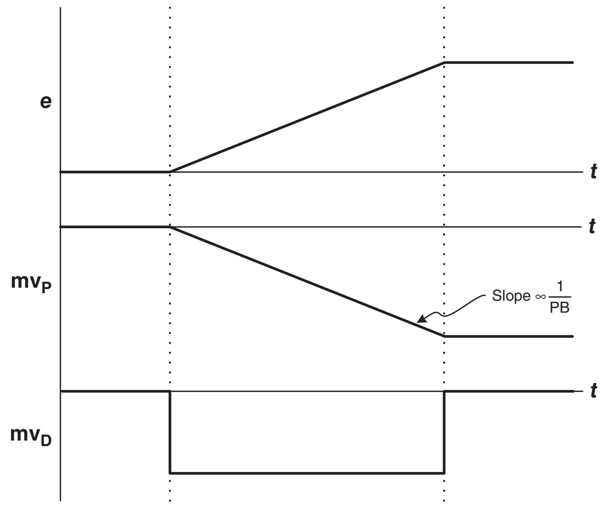

Figure 4.14 *Responses for P-only and D-only portions of a PD controller.*

In Equation 4.18, for the PD controller the derivative action acts on the error. Since $e =$ SP – 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 (Equation 4.19):

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

If there is a load upset to the process, 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 percent

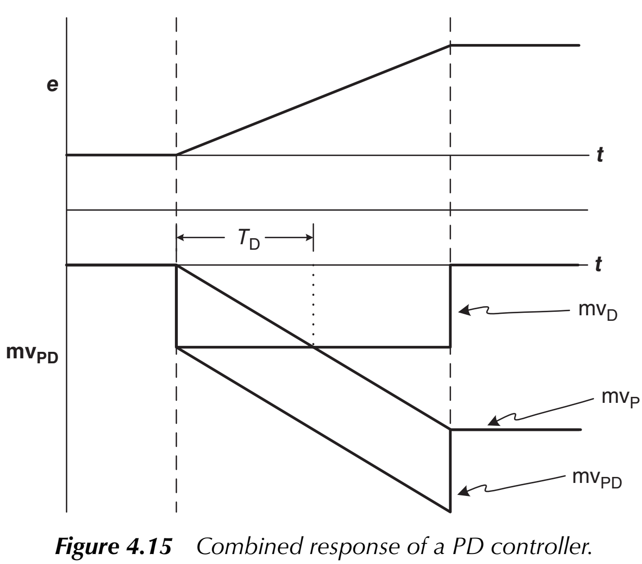

Figure 4.15 *Combined response of a PD controller.*

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, 4.21 and 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

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

Hence,

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

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, that is, if dCV/dt is large.

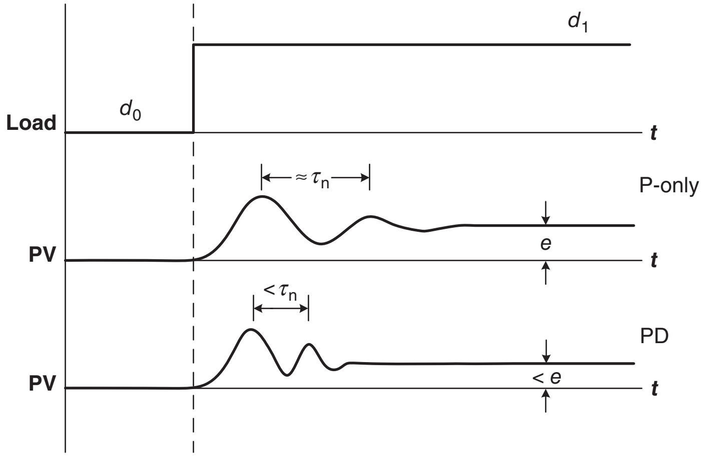

Figure 4.16 *Proportional derivative controller response to a load disturbance.*

In electronic controllers and distributed control systems (DCSs) the derivative action can be eliminated by setting T_d to zero. In a pneumatic controller the derivative action cannot be eliminated but can be reduced to a minimum value of approximately 0.01 minute. If a PD controller is installed on a flow loop there will still be considerable derivative action due to the noisy flow measurement. It is therefore 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 actions to produce a three-mode controller, 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:

$$\mathrm{mv} = K_c\left(e + \frac{1}{T_i}\int e\,\mathrm{d}t - T_D\frac{\mathrm{CV}}{\mathrm{d}t}\right). \tag{4.23}$$

Figure 4.17 presents a comparison of the responses for a 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.

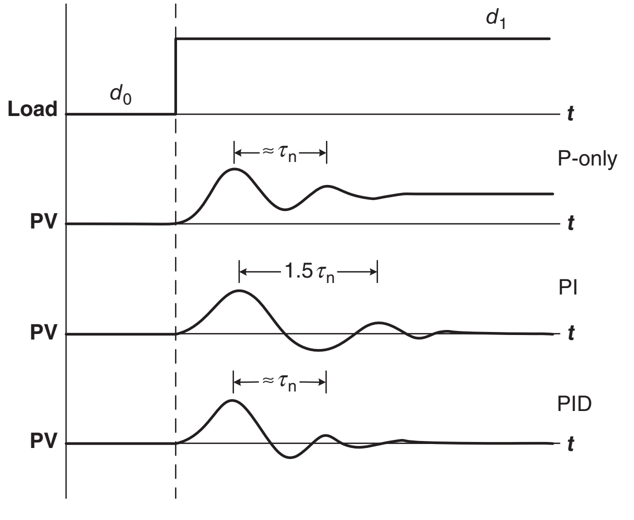

Figure 4.17 *P-only, PI and PID controller response to a load disturbance.*

4.8 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.

The positional form of the PID controller algorithm is

$$\mathrm{mv}(t) = K_{\mathrm{c}}\left(e(t) - e(t-1) + \frac{1}{T_{\mathrm{i}}}\sum_{k=0}^{n} e(kh) - T_{\mathrm{D}}\frac{\mathrm{CV}(t) - \mathrm{CV}(t-1)}{h}\right), \quad (4.24)$$

where $\mathrm{mv}(t)$, $e(t)$ and $\mathrm{CV}(t)$ are the current controller output, error and controlled variables, respectively; t is the enumerated sampling instant in time; $\mathrm{mv}(t-1)$, $e(t-1)$ and $\mathrm{CV}(t-1)$ 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

$$\mathrm{mv}(t) = \mathrm{mv}(t-1) + K_{\mathrm{c}}\left(e(t) - e(t-1) + \frac{1}{T_{\mathrm{i}}}e(t) - T_{\mathrm{D}}\frac{\mathrm{CV}(t) - 2\mathrm{CV}(t-1) + \mathrm{CV}(t-2)}{h}\right). \quad (4.25)$$

For the positional form it is important to note how to handle properly the summation term associated with the integral action. 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.

4.9 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, a proportional-only controller can be used. If the

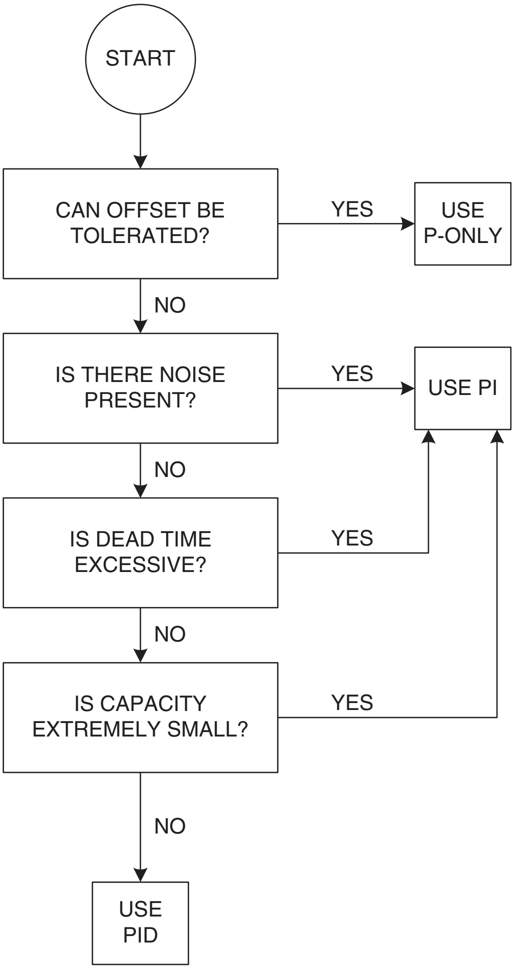

Figure 4.18 *Flow chart for controller selection.*

answer is no, 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, proceed 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, 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 of applications in the fluid-processing industries. Finally, if the process capacitance is large, a PID controller can be effectively 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, while there are four decision blocks that must be passed through to reach a PID controller.

4.10 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 an old version of the electronic digital controller. 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, startup values, on/off modality and gain schedule limits. A more current DeltaV™ digital controller is shown in Figure 4.21.

Figure 4.22 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

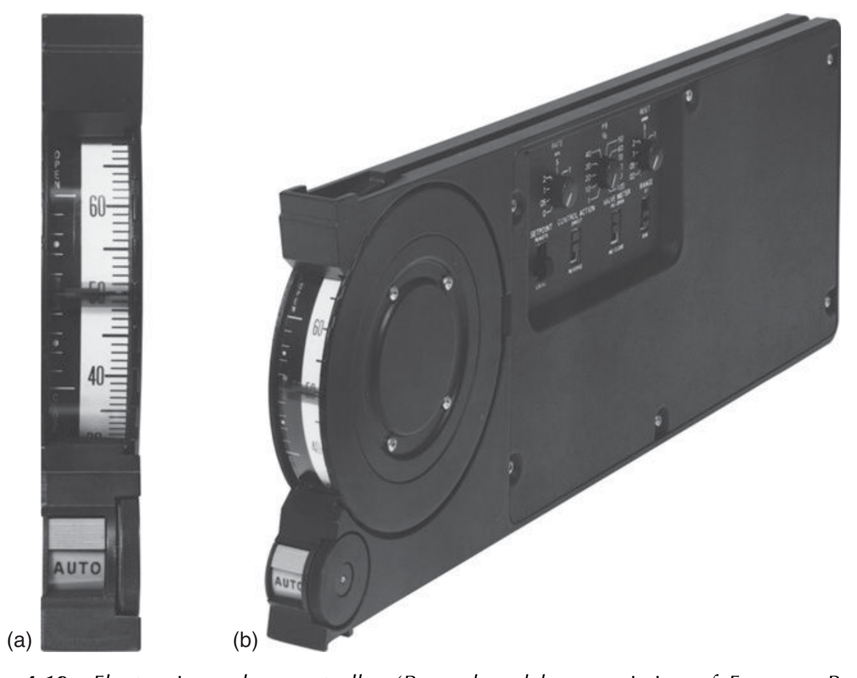

Figure 4.19 *Electronic analog controller (Reproduced by permission of Emerson Process Management).*

Figure 4.20 *Electronic digital controller – Fisher DPR series (Reproduced by permission of Emerson Process Management).*

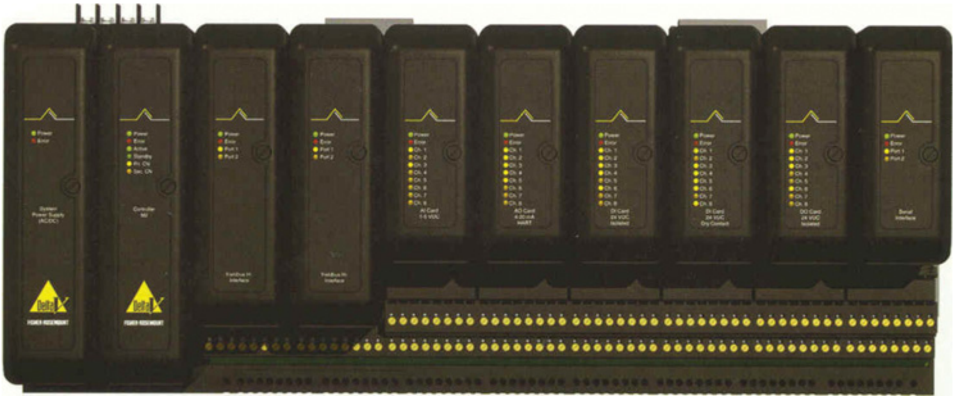

Figure 4.21 *DeltaV Controller (Reproduced by permission of Emerson Process Management).*

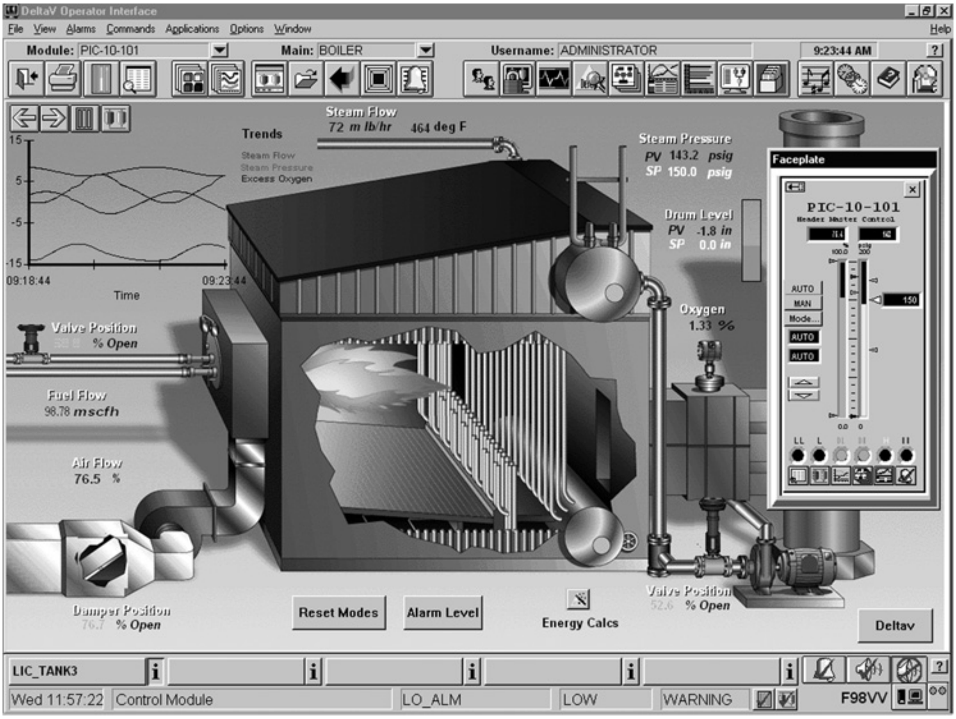

Figure 4.22 *Screen shot from a DeltaV distributed control system (Reproduced by permission of Emerson Process Management).*

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 the digital devices that can communicate directly with each other to fully exploit each other's capabilities with remote diagnostics and supervisory control and data acquisition (SCADA). This control technology is known as Fieldbus (http://www.fieldbus.org/) [2]. Additional detail pertaining to Foundation Fieldbus for instrumentation and control can be found in Appendix 3 (A. Dicaire, Personal communication; B. Van Vliet, Personal communication).

With the ability to send data wirelessly, information to and from field devices can be sent and received at a single I/O point via a gateway that has coverage for up to 100 different devices. Examples are shown in Appendix 3 (B. Van Vliet, Personal communication). Back in the panel, electronic marshalling unique to DeltaV can be done too. It is based on traditional I/O practices, but delivers value in reduced infrastructure and associated engineering design [3].

The PC explosion of the 1980s and 1990s has also impacted modern DCSs with the advent of PC-based control systems [4, 5] which feature object linking and embedding (OLE) software for process control (OPC) (http://www.opcfoundation.org/).

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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 type of disturbances 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 and, hence, most systems are optimized for regulatory control.

This chapter will discuss control quality and optimization, 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 Optimization

Controller tuning can be defined as an optimization 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, Third Edition. William Y. Svrcek, Donald P. Mahoney and Brent R. Young.

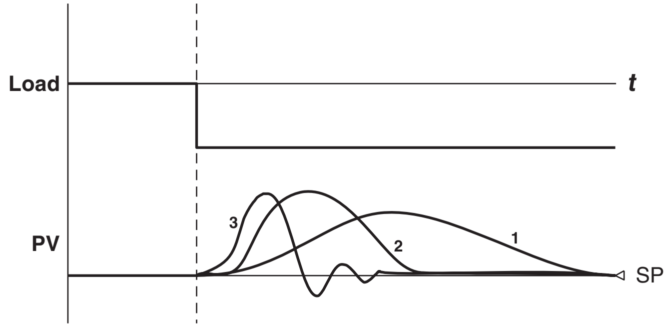

Figure 5.1 Typical responses to a load change.

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 the selection of good control is a tradeoff 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 the 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

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

If Equation 5.2 is substituted into Equation 5.1, we obtain

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

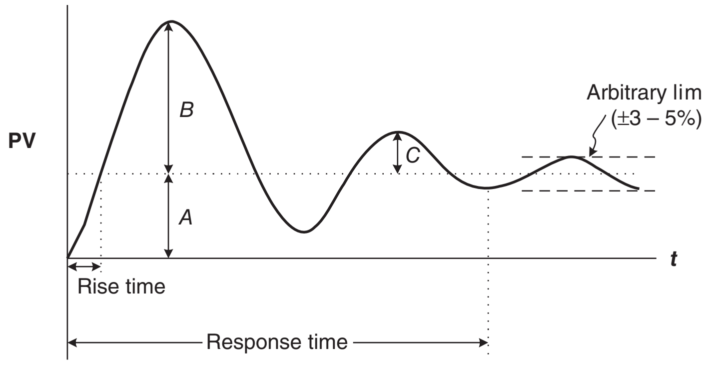

Figure 5.2 *Second- or higher-order typical response to a set point change.*

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

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

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 or similar has been shown through experience to provide a reasonable tradeoff 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, leading to many recommending less than QDR for process control. For a second-order system it can be shown that

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

Rise Time

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

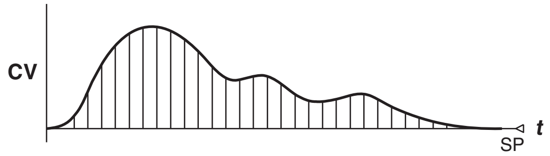

Figure 5.3 *General response curve.*

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 $\pm$ 3–5% of the PV steady-state value.

5.1.2 Error Performance Criteria

The previously discussed simple performance criteria, that is, decay ratio, overshoot, and so on, 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 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 minimize 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 minimize 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 IE may not be 100% reliable if there is no averaging elsewhere in the process. For example, if there is a sinusoidal oscillation about the set point, 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, IE is a perfectly adequate error criterion.

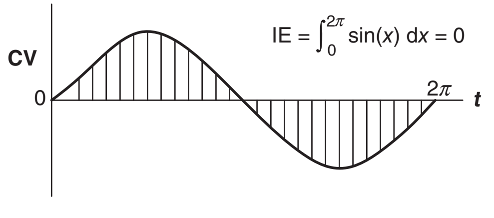

Figure 5.4 *Sinusoidal error response.*

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, that is, faster return to the set point:

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

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 penalized 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 IAE, ISE and ITAE.

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 optimized for a particular error performance criterion, often QDR. The first method described is based entirely on trial and error, while the rest are based upon some understanding of the physical

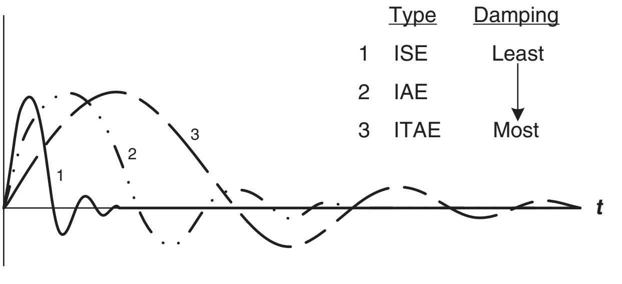

Figure 5.5 *Response to various error criteria.*

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 retuning or 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 2, that is, 0.25, 0.5, 1.0, 2.0, 4.0 and so on.
4. The optimal response is one's chosen criterion, for example, QDR or less.
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 but rather 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:

$$K_c = 0.4\text{–}0.65,$$
$$T_i = 0.1 \text{ minutes (or 6 seconds).}$$

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% open when the level is at 50%, wide open when the level is at 75% and shut when the level is at 25%:

$$K_c = 2,$$
$$\text{Bias term}\,(b) = 50\%,$$
$$\text{Set point}\,(\text{SP}) = 50\%.$$

Typical PI controller settings are

$$K_c = 2\text{–}20,$$
$$T_i = 1\text{–}5 \text{ minutes.}$$

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

$$\begin{aligned}\text{Vapour:}\quad & K_c = 2\text{–}10,\\ & T_i = 2\text{–}10 \text{ minutes.}\\ \text{Liquid:}\quad & K_c = 0.5\text{–}2,\\ & T_i = 0.1\text{–}0.25 \text{ minutes.}\end{aligned}$$

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

$$\begin{aligned}K_c &= 2\text{–}10,\\ T_i &= 2\text{–}10 \text{ minutes,}\\ T_d &= 0\text{–}5 \text{ minutes.}\end{aligned}$$

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 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 are as follows:

L is the lag time (min); T is the time constant estimate (min); P is the initial step disturbance (%); ΔC_p is the change in PV in response to step disturbance, (change in PV)/(PV span) $\times$ 100, (%); $N = \Delta C_p/T$ is the reaction rate (%/min); and $R = L/T = NL/\Delta C_p$ is the lag ratio (dimensionless).

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.

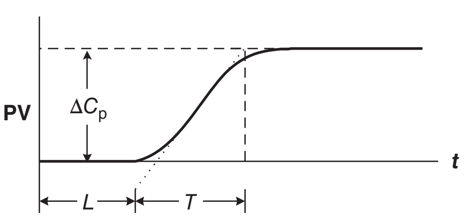

Figure 5.6 *Process reaction curve.*

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, which is the basis of the general process reaction curve 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 QDR are as follows:

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

$$\text{PI:} \quad K_c = 0.9\left(\frac{P}{NL}\right), \tag{5.13}$$

$$T_i = 3.33\,(L), \tag{5.14}$$

$$\text{PID:} \quad K_c = 1.2\left(\frac{P}{NL}\right), \tag{5.15}$$

$$T_i = 2.0\,(L), \tag{5.16}$$

$$T_d = 0.5\,(L). \tag{5.17}$$

These settings should be taken as recommendations only and tested thoroughly in closed loop, fine-tuning the parameters to obtain QDR or whatever is the desired response criterion. It is worthwhile noting that for chemical process control, Ziegler–Nichols recommendations are often considered very aggressive as they were designed for servo control for US Naval applications and not process control. Consequently, tuning may have to be adjusted or fine-tuned for process control application (lower gain, higher integral time). Also, other workers have developed a series of less aggressive recommendations, often with less than QDR as a desirable response criterion.

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 in the relatively rare occasion when process dead time is large relative to the dominant open loop time constant. In other regards, the Cohen–Coon recommendations are still considered relatively aggressive for process control.

The Cohen–Coon-recommended controller settings are as follows:

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

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

$$T_i = L\left(\frac{30 + 3R}{9 + 20R}\right), \tag{5.20}$$

$$\text{PID:} \quad K_c = \frac{P}{NL}\left(1.33 + \frac{R}{4}\right), \tag{5.21}$$

Table 5.1 *Simplified IMC rules [6].*

	$\frac{\tau}{L} > 3$	$\frac{\tau}{L} < 3$	$L < 0.5$
K_c	$\frac{P}{2(NL)}$	$\frac{P}{2(NL)}$	$\frac{P}{N}$
T_i	$5L$	τ	4
T_d	$\leq 0.5L$	$\leq 0.5L$	$\leq 0.5L$

$$T_i = L\left(\frac{32 + 6R}{13 + 8R}\right), \tag{5.22}$$

$$T_d = L\left(\frac{4}{11 + 2R}\right). \tag{5.23}$$

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 QDR and are also usually considered aggressive so that de-tuning may be required.

Internal Model Control Tuning Rules

As mentioned above, 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, for example [4], is a term that is also used to refer to controller tuning methods that are based on a specified closed loop time constant. More recently Skogestad has proposed IMC tuning rules that have also been shown to exhibit superior performance in a number of studies [5]. 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*. [6] 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.

5.2.3 Constant Cycling Methods

Ziegler–Nichols Closed Loop Method

The closed loop technique of Ziegler and Nichols [1] 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 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.

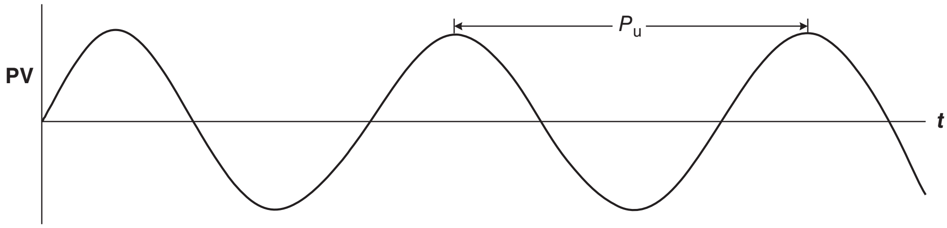

Figure 5.7 *Constant amplitude limit cycle.*

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 following parameters from the constant amplitude limit cycle (Figure 5.7):

   $P_u \equiv$ ultimate period = period taken from limit cycle;

   $K_u \equiv$ ultimate gain = controller gain that produces the limit cycle.

5. Calculate the tuning parameters using the following equations:

$$\text{P-only:} \quad K_c = \frac{K_u}{2}, \tag{5.24}$$

$$\text{PI:} \quad K_c = \frac{K_u}{2.2}, \tag{5.25}$$

$$T_i = \frac{P_u}{1.2}, \tag{5.26}$$

$$\text{PID:} \quad K_c = \frac{K_u}{1.7}, \tag{5.27}$$

$$T_i = \frac{P_u}{2}, \tag{5.28}$$

$$T_d = \frac{P_u}{8}. \tag{5.29}$$

6. Fine tune by adjusting K_c, T_i and T_d (as required) to find the desired response, for example, QDR or less. Again, it is important to note here that the Ziegler–Nichols recommendations are considered relatively aggressive for chemical process control and this step is usually very necessary.

Yuwana–Seborg Closed Loop Tuning

Another closed loop method that retains the advantage of being performed under control in the loop but addresses the disadvantage of the oscillatory response that results from the Ziegler–Nichols technique is that of Yuwana–Seborg [7]. The technique is based on the assumption that the plant dynamics, while unknown, can be approximated by a first-order

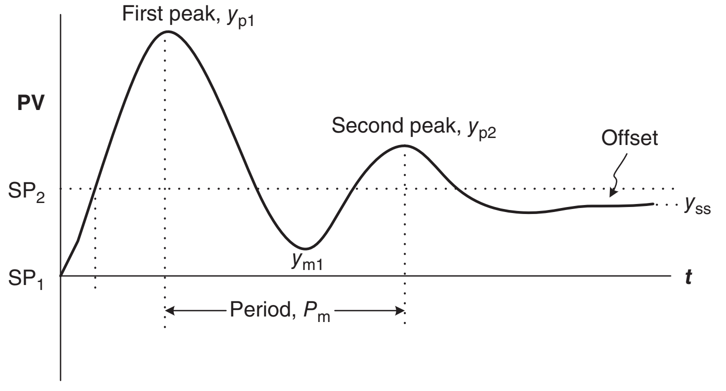

Figure 5.8 *Typical response of a stable first-order plus dead-time system using P control.*

plus dead-time model of process gain, K_p, time constant, τ, and dead time, L. The method is briefly as follows:

1. Put the controller in automatic (close the loop).
2. Turn the integral and derivative action off (i.e. use P-only control).
3. Introduce a step change in the set point and adjust the controller gain to obtain a decaying underdamped response (Figure 5.8).
4. Log the measured period between the first and second peaks, P_m, the first peak, y_{p1}, the second peak, y_{p2}, the midpoint between them, y_{m1}, and the final steady-state value, y_{ss}.
5. Compute the plant model from the following equations:

$$\text{Final steady state:} \quad y_{ss} = \frac{y_{p2}y_{p1} - y_{m1}^2}{y_{p1} + y_{p2} - 2y_{m1}}, \tag{5.30}$$

$$\text{Open process loop gain:} \quad K_p = \frac{|y_{ss} - y_0|}{K_c\left(|SP_{ss} - SP_0| - |y_{ss} - y_0|\right)}, \tag{5.31}$$

$$\text{Time constant:} \quad \tau = \frac{P_m \beta_1 \beta_2}{2\pi}, \tag{5.32}$$

$$\text{Dead time:} \quad L = \frac{P_m \beta_2}{\pi \beta_1}, \tag{5.33}$$

where

$$\beta_1 = \psi\sqrt{K_c K_p + 1} + \sqrt{\psi^2(K_c K_p + 1) - 1 + K_c K_p}, \tag{5.34}$$

$$\beta_2 = \sqrt{(1 - \psi^2)(K_c K_p + 1)}, \tag{5.35}$$

$$\psi = \frac{\ln\alpha}{\sqrt{\pi^2 + (\ln\alpha)^2}} \tag{5.36}$$

and

$$\alpha = \frac{y_{p1} - y_{ss}}{y_{ss} - y_{m1}}. \tag{5.37}$$

Table 5.2 *Yuwana–Seborg tuning recommendation constants.*

Controller	*A*	*B*	*C*	*D*
P	0.49	1.084	–	–
PI	0.859	0.977	1.484	0.680
PID	1.357	0.947	1.176	0.738

6. Compute the P, PI or PID tuning from the following recommendations:

$$\text{Controller gain: } K_c = \frac{A}{K_p}\left(\frac{L}{\tau}\right)^{-B}, \tag{5.38}$$

$$\text{Integral time: } \quad \frac{1}{T_i} = \frac{1}{C\tau}\left(\frac{L}{\tau}\right)^{D}, \tag{5.39}$$

$$\text{Derivative time: } \quad T_d = 0.381\tau\left(\frac{L}{\tau}\right)^{0.99}, \tag{5.40}$$

   where the constants *A*–*D* are obtained from Table 5.2.
7. Test and fine tune with the new recommended tuning values. As with other methods, if the response is not satisfactory, further manual, fine-tuning may be needed.

The algorithm is actually quite easy to implement. The main challenge is the reliable determination of the peaks and troughs.

Auto Tune Variation Technique

The auto tune variation (ATV) technique of Åström and Hagglund [8] is another closed loop technique used to determine the two important closed loop system constants, 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 [9] 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 5.9 shows the instrument setup, and Figure 5.10 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 significant dead time. The ultimate period will just equal the sampling period if the dead time is not significant.

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%). 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.

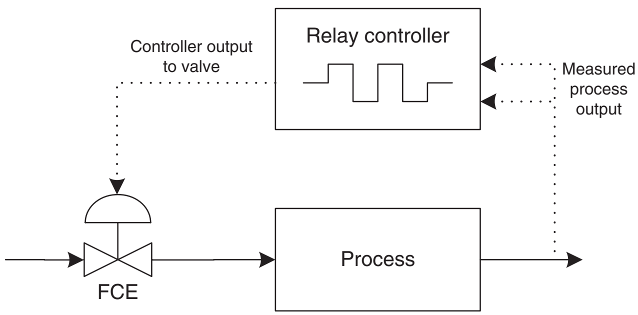

Figure 5.9 *ATV tuning instrument setup.*

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 5.10.
6. Record the value of a by picking it off the response graph.
7. Perform the following calculations to determine the ultimate period, ultimate gain and the controller gain and integral time.
8. Implement the recommendations, test and fine tune. These Tyreus–Luyben [9] recommendations are less aggressive than Ziegler–Nichols so the expected closed loop response is less than QDR with settings and usually considered more suitable for chemical process control.

$$P_u \equiv \text{ultimate period} = \text{period taken from limit cycle},$$

$$K_u \equiv \text{ultimate gain} = \frac{4h}{3.14\,a},$$

$$K_c = \frac{K_u}{3.2}, \tag{5.41}$$

$$T_i = 2.2\,(P_u)\,. \tag{5.42}$$

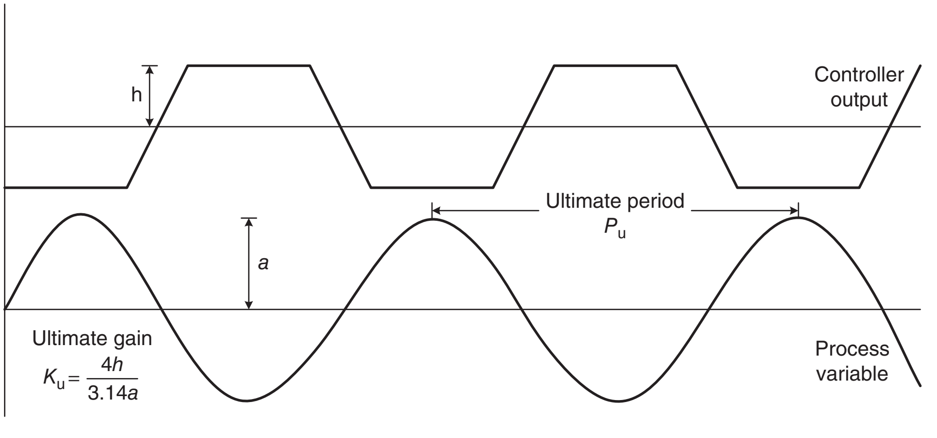

Figure 5.10 *ATV critical parameters.*

Table 5.3 *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 5.3 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 emphasizes minimizing overshoot. ATV is, therefore, often the preferred technique for process control.

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6

Advanced Topics in Classical Automatic Control

Up to this point 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 should also be considered [1], which can provide improved and economic process control. 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

Leading industrial practitioners have indicated that cascade control is a widely used controller configuration. Hence, it needs to be stressed in an undergraduate control course and its inclusion here is as the fore most advanced classical control technique.

Cascade control is a 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 final control element (FCE). A typical cascade control loop is illustrated in Figure 6.1.

To better understand cascade control, 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 was properly tuned and there is a change in F_w, 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

A Real-Time Approach to Process Control, Third Edition. William Y. Svrcek, Donald P. Mahoney and Brent R. Young.

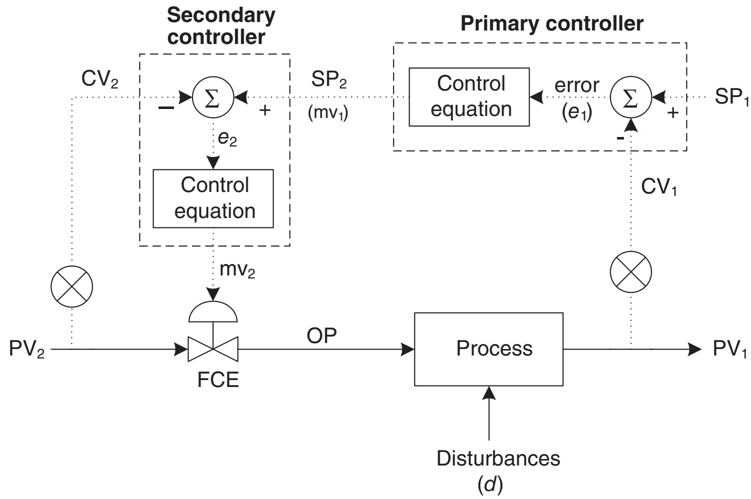

Figure 6.1 *Cascade control system.*

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 which is also servicing other users, there is a possibility that as the other users' needs vary, pressure upsets will occur causing 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 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.

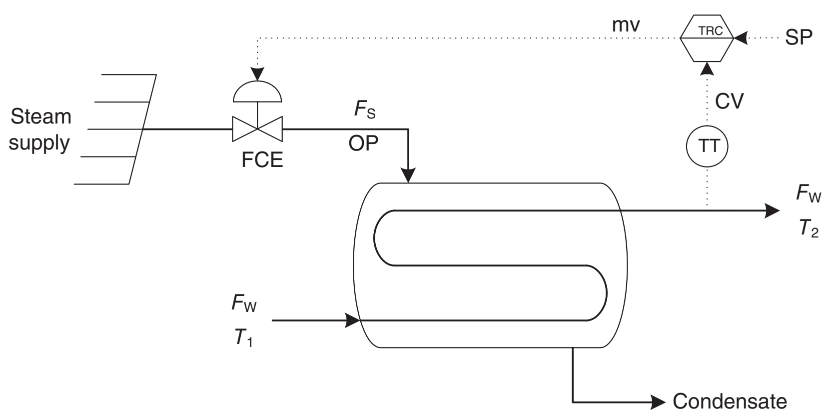

Figure 6.2 *Feedback control loop for a heat exchanger (steam supply loop).*

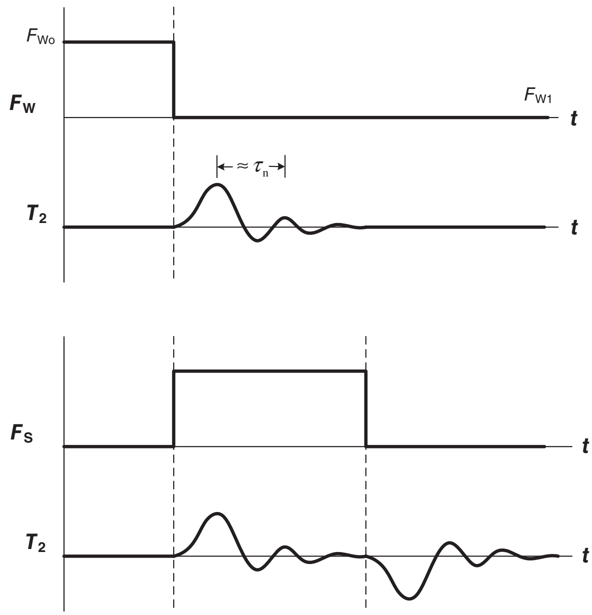

Figure 6.3 *Heat exchanger outlet temperature response.*

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%) supplies the needed amount of steam, the outlet temperature will be at the set point. However, the flow through the valve is a function of the pressure drop across it. Therefore, if there is a decrease in inlet pressure the flow will decrease, even though the valve is 50% 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, 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 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, the output of the temperature controller will change the steam flow controller set point. The flow loop operates so much faster than the

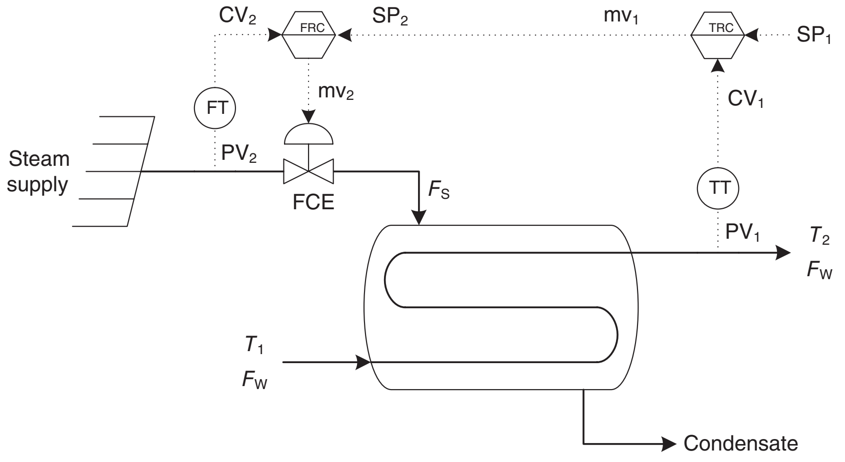

Figure 6.4 Typical cascade control loop for a heat exchanger.

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 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 that cascade control works 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 normally.

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 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 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 to 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. Other common industrial examples of cascade control include level to flow cascade control, heat exchanger temperature control as per the example above, temperature control of furnaces/fired heaters, temperature control of exothermic stirred chemical reactors and distillation composition to temperature to flow cascade control. Most modern advanced process control schemes such as model predictive control are also implemented in a cascade arrangement over a regulatory control layer.

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, that is, 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 to that of the disturbance, 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, the feedforward dynamic compensation would have to be a predictor, which of course is impossible unless an exact and very fast dynamic model

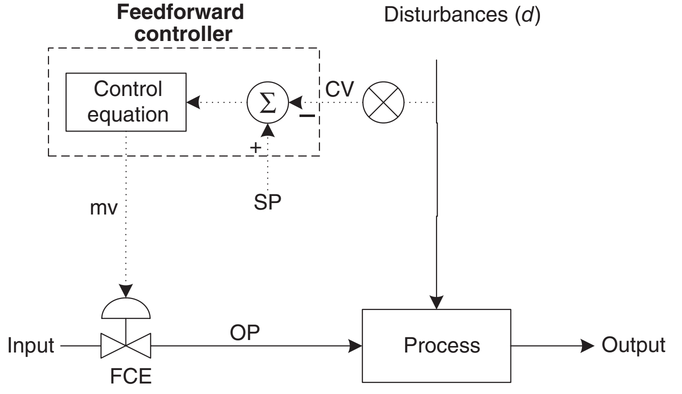

Figure 6.5 *Feedforward control system.*

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 precisely known. If unexpected disturbances enter the process when only feedforward control is used, 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, 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.

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.

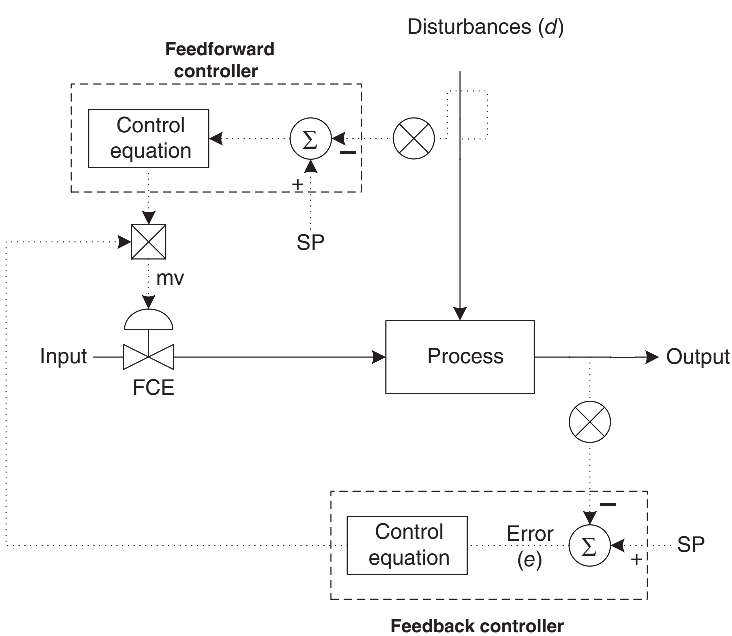

Figure 6.6 *Feedforward/feedback control system.*

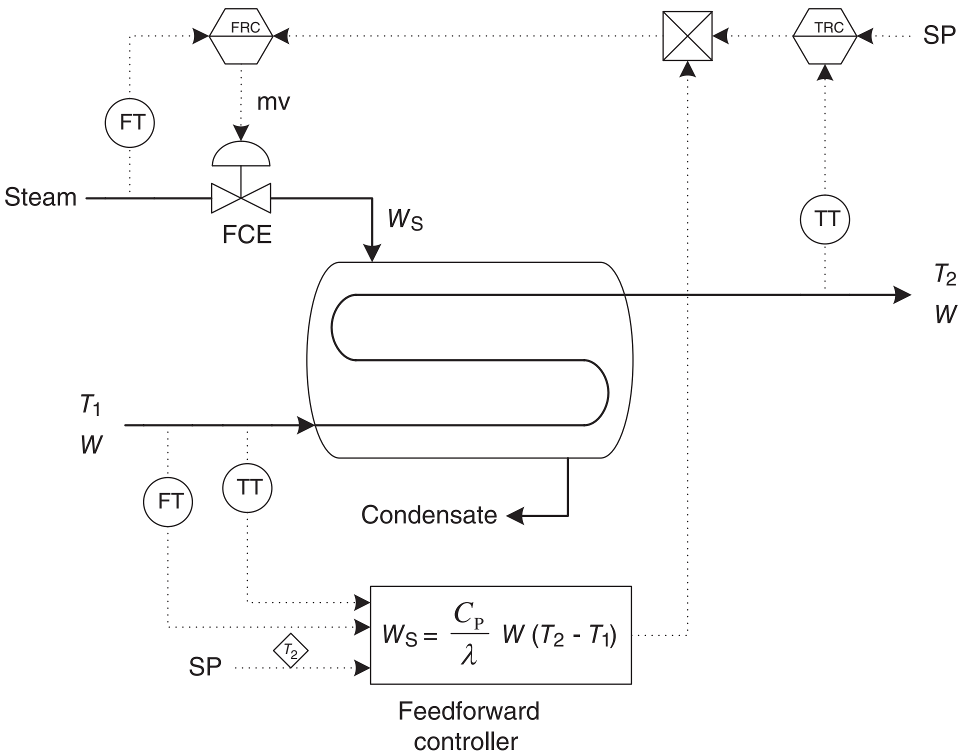

Figure 6.7 *Combined feedforward and cascade control of a heat exchanger.*

At steady state an overall heat balance can be written for the process as shown in Equations 6.1, 6.2 and 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 is the enthalpy transferred by the steam condensing to form condensate (kJ/kg) and C_p is the heat capacity of the process fluid (kJ/kg K).

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 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.

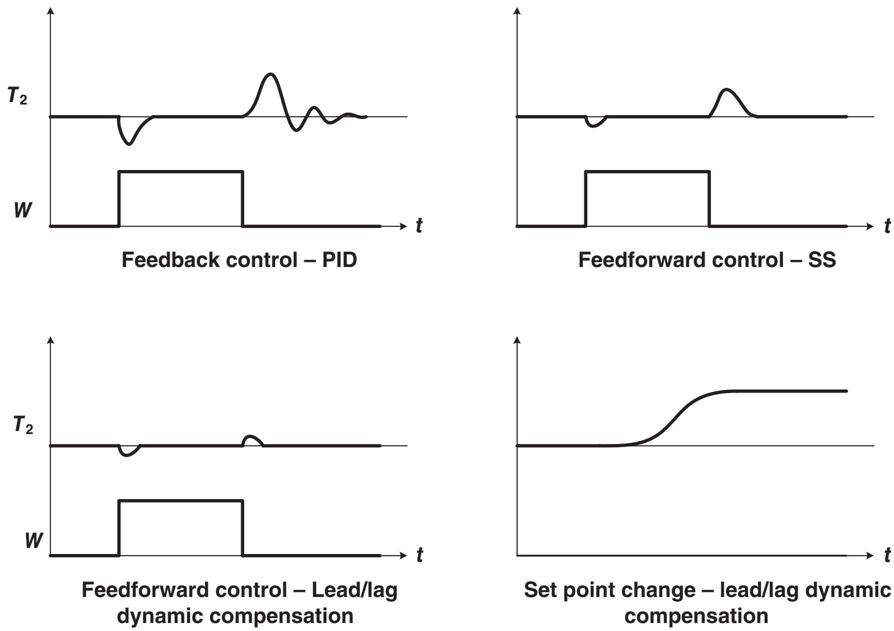

Figure 6.8 *Typical responses of the heat exchanger.*

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.

Common industrial examples of feedforward control in addition to heat exchanger control as shown above include control of fired heaters, chemical reactors and distillation columns. Other examples include control of biological process systems, such as fermentors and activated sludge processes and so on. An industrial illustration of when feedforward control was combined with feedback to control dissolved oxygen addition to a municipal wastewater treatment process resulted in significant savings in air blower energy consumption for the wastewater treatment plant [5]. Figure 6.9 illustrates the successful control scheme.

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

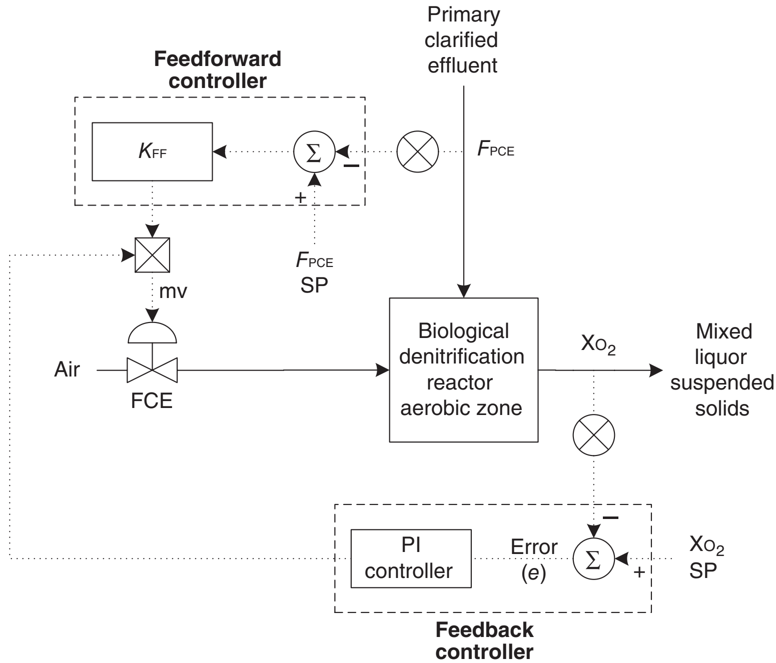

Figure 6.9 Feedforward/feedback control of a wastewater denitrification process.

streams. Ratio control would ensure that these streams were being 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 6.10.

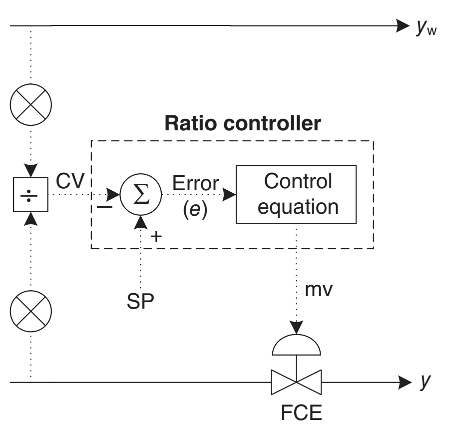

Figure 6.10 Typical ratio control system.

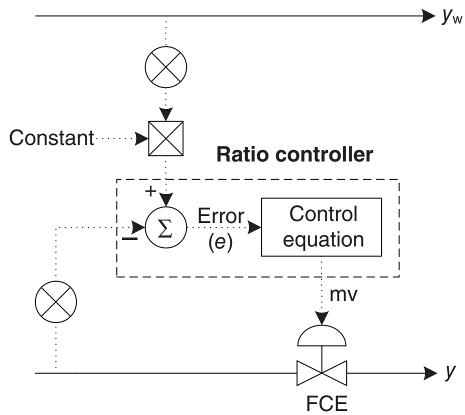

Figure 6.11 *Typical ratio control system.*

In this first scheme, both flows are measured and divided to obtain the actual ratio. This is then compared to 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 6.11, 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 to the measured flow and adjusts the flow of y accordingly. In this scenario, the desired ratio is a constant variable in the multiplier and if a new value for the ratio is needed, it must be set in the multiplier.

The ratio control configuration shown in Figure 6.10 will not have a steady loop gain because the ratio calculation itself is in the loop. The loop in Figure 6.10 may become nonlinear, making the control configuration in Figure 6.11 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 to the column, V, while 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, a selection must be made for control purposes (single input/single output). 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].

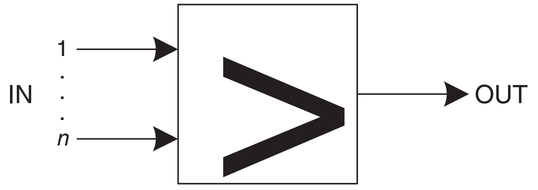

Figure 6.12 High selector.

The two basic building blocks for selector systems are the high and the low selector. The high selector, shown in Figure 6.12, will pass the highest value of the multiple inputs to the output signal, ignoring all other inputs.

The low selector, shown in Figure 6.13, 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 selectors, such as a median value selection, shown in Figure 6.14. 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).
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 6.15. 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_o exceeds a certain maximum pressure, close the valve. (Assume $P_o > P$ shut off).

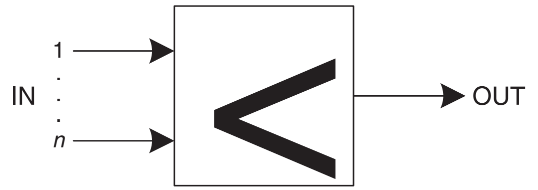

Figure 6.13 Low selector.

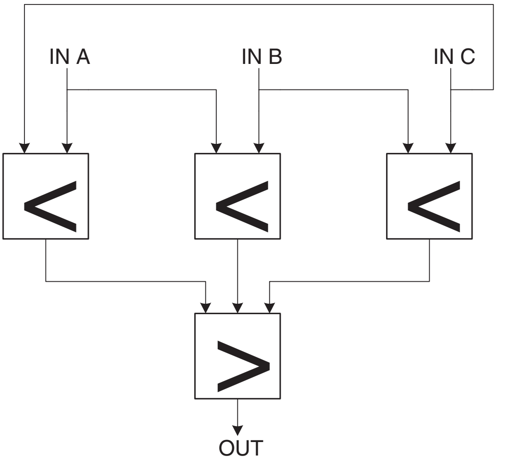

Figure 6.14 *Median value selector.*

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 selector 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.

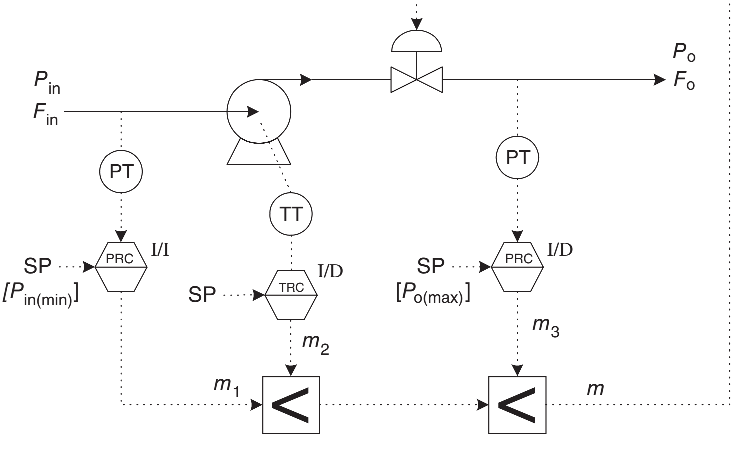

Figure 6.15 *Protection of equipment – pump.*

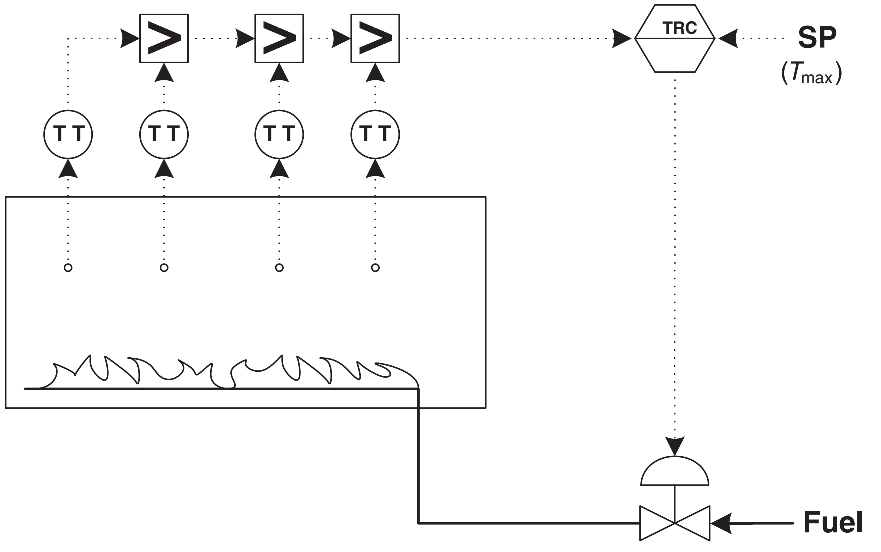

Figure 6.16 Auctioneering – temperature control in an oven.

6.4.2 Auctioneering

The objective of auctioneering is to protect against the highest temperature sensed by one of many temperature transmitters. In the example shown in Figure 6.16, 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 6.17) too much catalyst can prove disastrous. By implementing a fail-safe scheme that consists of two composition transmitters 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, 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 it will get selected and the catalyst flow will be stopped. Production is stopped and a possible hazardous situation is avoided.

An alternate scheme, shown in Figure 6.18, 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, either upscale or downscale, the selector will still choose the median value.

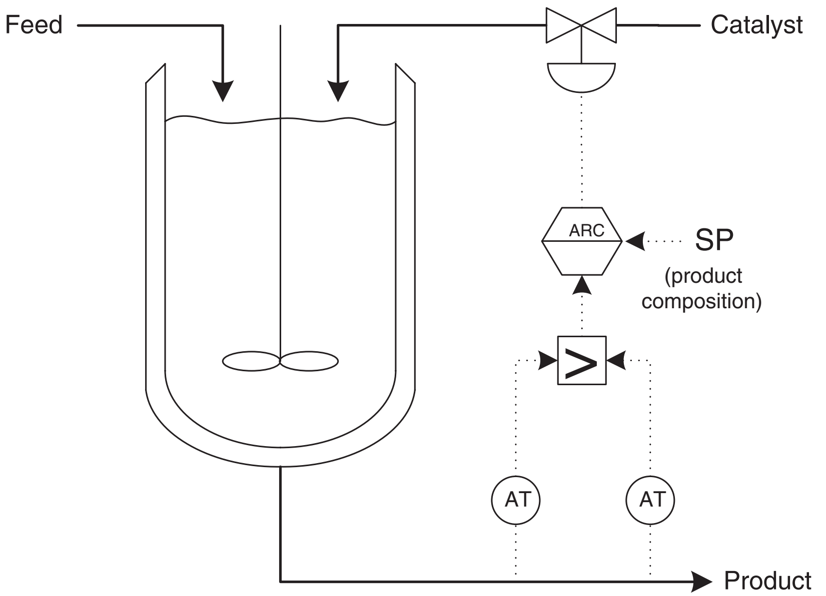

Figure 6.17 *Redundant instrumentation – reactor.*

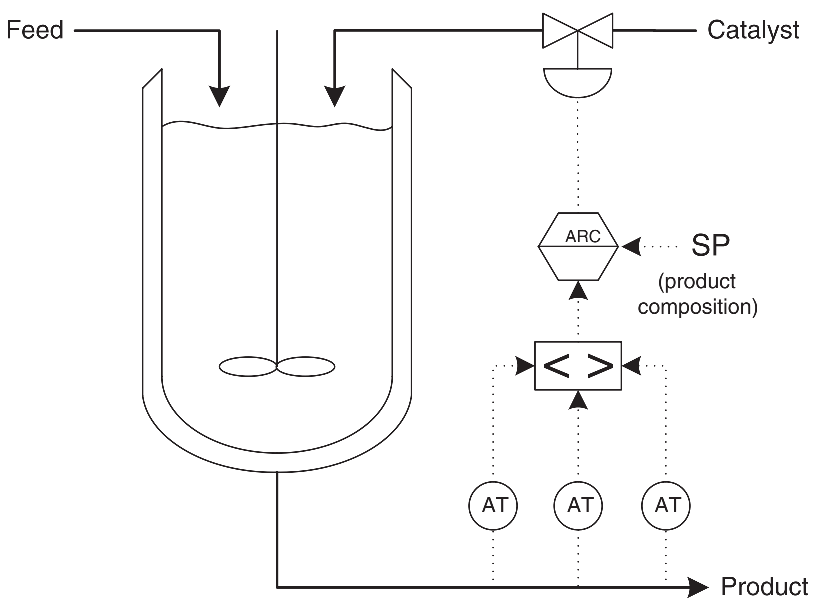

Figure 6.18 *Redundant instrument – reactor/median selector.*

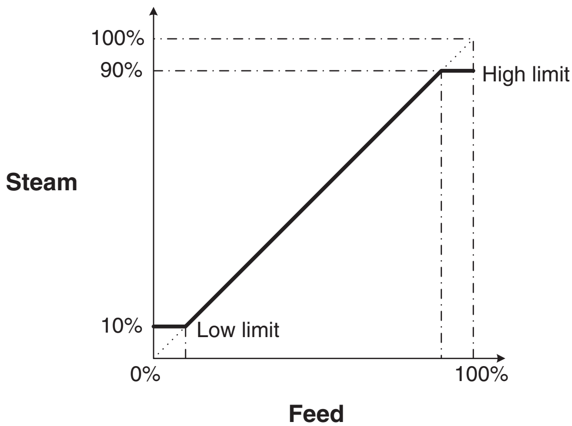

Figure 6.19 *Feed/steam characteristic of a distillation column.*

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 6.19.

Instead of operating the steam versus feed flow on a straight line, operating constraints are set. The operating constraints require a minimum steam rate of 10%, 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%, the high limit. These constraints can be implemented as shown in Figure 6.20.

If the feed flow is within the safe operating region, the signal from the multiplier will pass through the high selector since it is higher than the low limit. It will also pass 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 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.

6.5 Split Range Control

Split range control may be useful in processes where there is one controlled variable and there are extra manipulated variables. Each of the extra manipulated variables must be able to affect the controlled variable. An example of split range control is illustrated in the control of an exothermic reactor, Figure 6.21, which will also be discussed in further detail in Chapter 7 when we deal with reactor control more comprehensively.

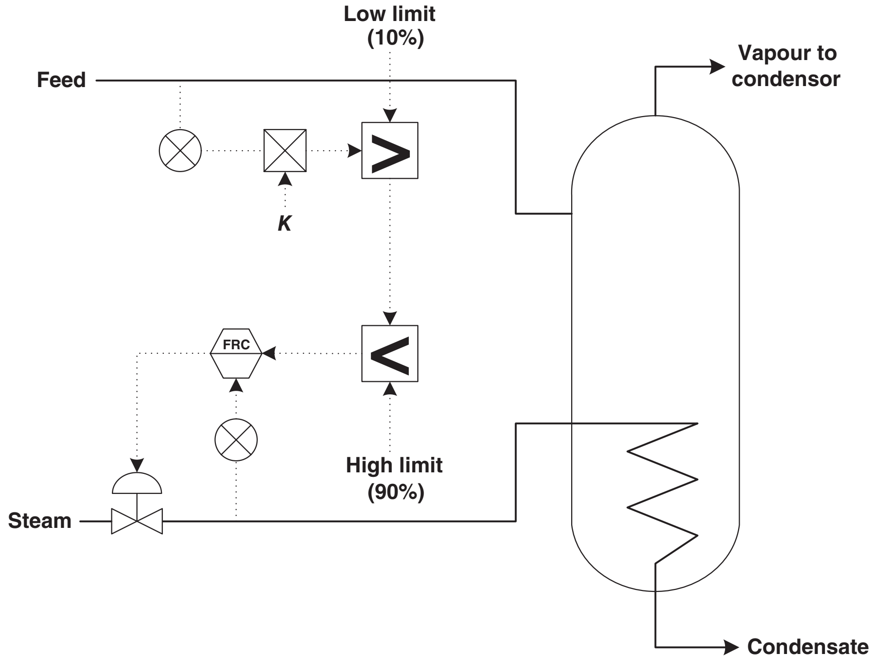

Figure 6.20 *Artificial constraints.*

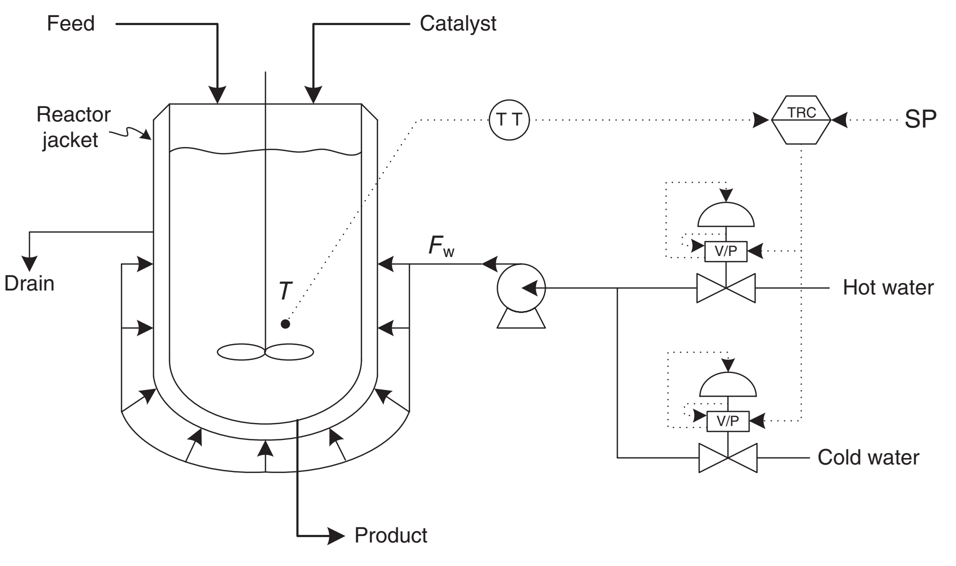

Figure 6.21 *Control scheme for an exothermic reactor.*

The exothermic reactor in Figure 6.21 can be stabilized if the reaction temperature changes fairly slowly when compared to the rate at which the jacket temperature changes. The idea of the split range control loop shown in Figure 6.21 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. This is split range control.

It is important to note that there are several implementation issues when applying split range control. One issue is the tuning of the system. When a different manipulated variable is selected by the controller, the closed loop process dynamics may change. Therefore the controller tuning should remain the same only if the process dynamics remain the same. For the example in Figure 6.21, this would mean that the hot water dynamics would be similar to the cold water dynamics. If the closed loop dynamics are significantly different, the controller tuning should be automatically changed, that is, a different set of tuning parameters should be used.

In practice there is a limit on how accurately a control valve can be adjusted. This has implications for the number of manipulated variables that may be used in split range. Split range control is normally limited to two or three manipulated variables for this reason. As a further consequence of inaccurate valves a split range control system could cycle if there is a dead band in which neither manipulated variable is adjusted. Valves in split range control systems are therefore calibrated to have an overlap, for example, of 10% to prevent this from occurring.

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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 responds very fast 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 non-linear 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, that is, orifice, and the d/p cell are connected together the response is non-linear, as shown in Figure 7.2.

A Real-Time Approach to Process Control, Third Edition. William Y. Svrcek, Donald P. Mahoney and Brent R. Young.

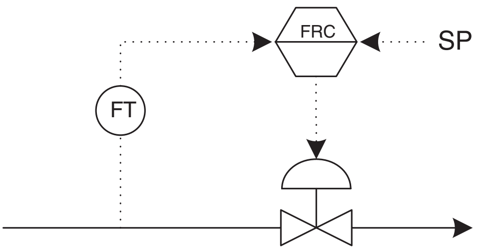

Figure 7.1 Flow control loop.

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 non-linearities 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.

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, 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, 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 effectively used in the controller because the noisy signals cause the loop to become unstable. Flow control is one type of loop where an integral-only (I-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

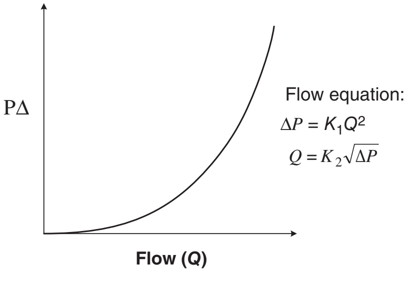

Figure 7.2 Head flow device response curve.

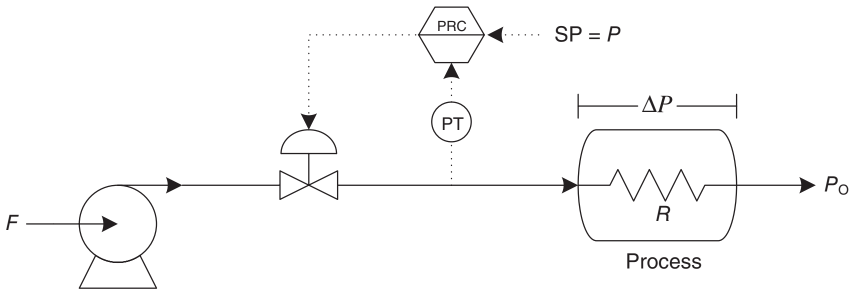

Figure 7.3 *Liquid pressure control loop.*

effective cascade control. Despite this, typically PI controllers are most often employed in flow loops because they are standard in the process control industry.

Tuning a flow loop for a PI controller is rather easy in comparison to other types of loops. The flow loop is so fast that quarter amplitude damping[1] 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 minutes. If there is an instability limit over the operating range due to non-linearities in the loop, 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, that is, 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, 7.2, 7.3 and 7.4:

$$P = \Delta P + P_o. \tag{7.1}$$

P_o is the downstream pressure at zero flow.

Also,

$$\Delta P = \frac{F^2}{R^2}. \tag{7.2}$$

[1] Also called the quarter decay ratio (QDR), see Chapter 5 for more details.

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 as shown in Equation 7.4:

$$K_p = \frac{dP}{dF}. \tag{7.4}$$

Substituting Equation 7.3 into Equation 7.4 results in Equation 7.5:

$$K_p = \frac{dP}{dF} = \frac{2F}{R^2}. \tag{7.5}$$

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 in 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%, and not from 0% to 100% as for the head flow device. For this case, the process gain is somewhat smaller than that for the flow process, and thus, a higher controller gain can be used, that is, between 1 and 2.

Other considerations for the liquid pressure loop are as follows:

- The controller can be proportional plus integral (PI) or 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 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.

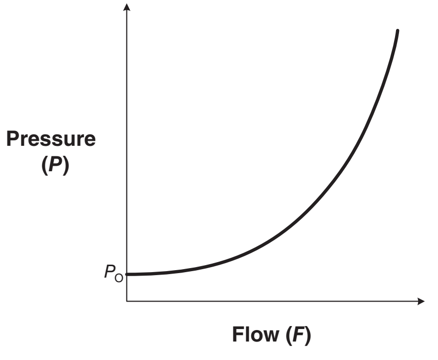

Figure 7.4 *Process gain of pressure flow loop.*

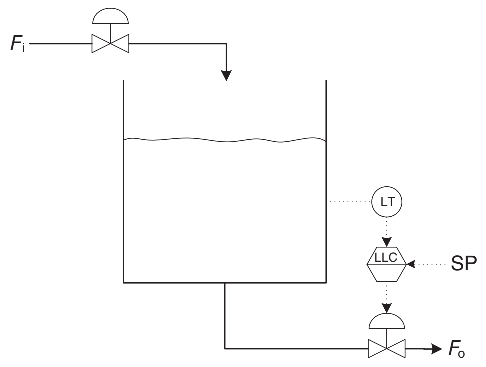

Figure 7.5 *Liquid level control loop.*

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 minutes. 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 and so on. 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 to

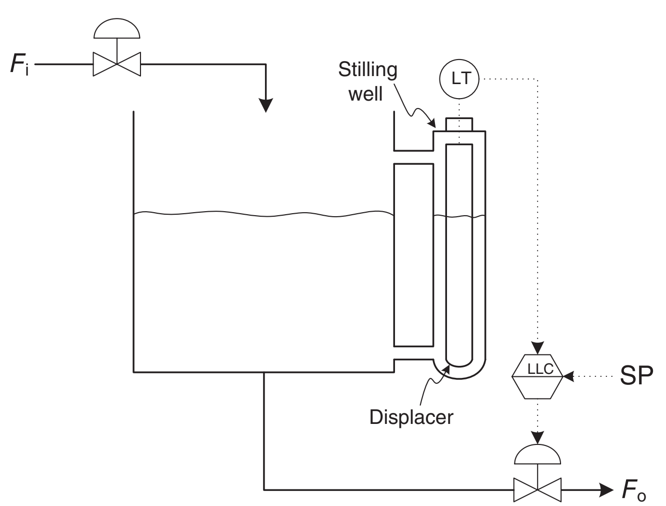

Figure 7.6 *Liquid level measurement with a stilling well.*

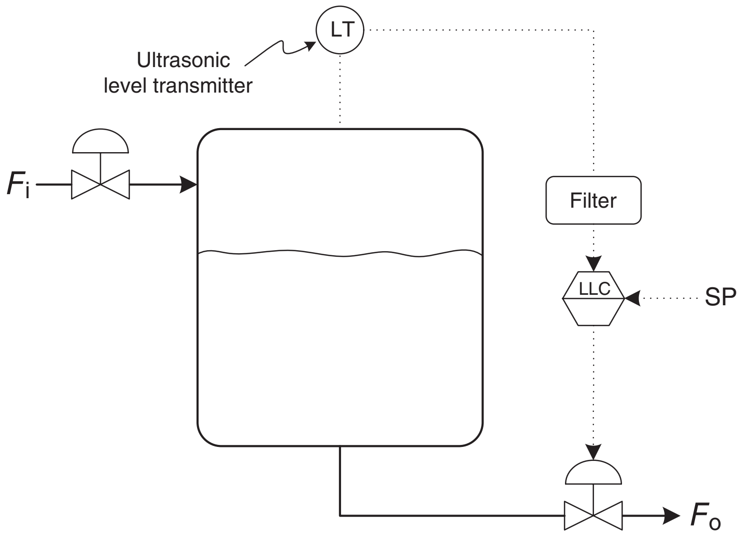

Figure 7.7 *Ultrasonic liquid level measurement.*

effectively 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, 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 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 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, 7.7 and 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 the use of a P-only controller for liquid level control? The equation for the error resulting from P-only control of a process is given in Equation 7.6:

$$e = \frac{\mathrm{mv} - b}{K_c}. \tag{7.6}$$

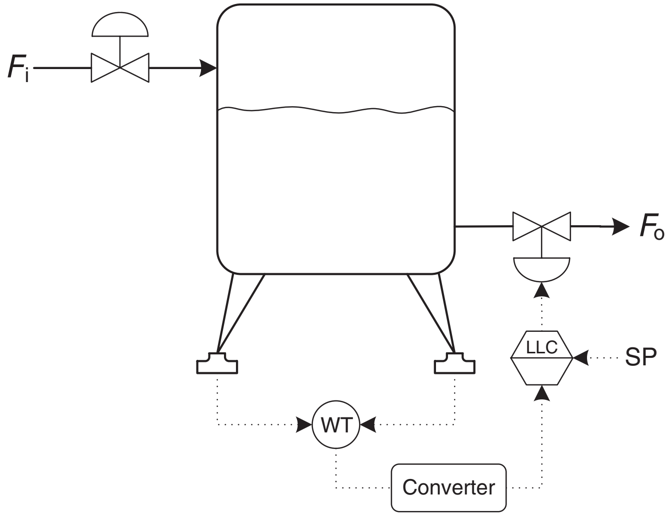

Figure 7.8 *Using load cell to measure mass in tank.*

As seen in the above equation, increasing the controller gain, K_c, will minimize the error. If the process gain is low, 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 percent 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 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 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 Equation 7.7:

$$K_p = \frac{\Delta \text{Out}}{\Delta \text{In}} = \frac{\Delta \text{Level}}{\Delta \text{Inflow}} = \frac{\Delta h}{\Delta F_i} \text{ for } \quad F_o = \text{Constant}. \tag{7.7}$$

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.

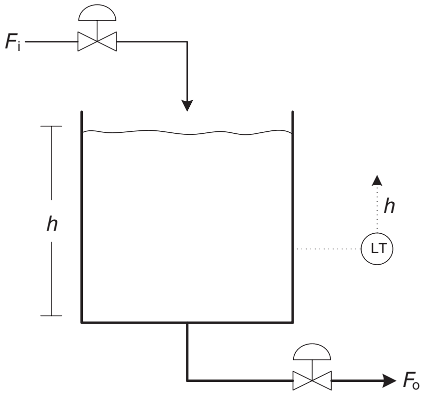

Figure 7.9 *Liquid level process.*

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, that is, F_{02}, F_{03} and

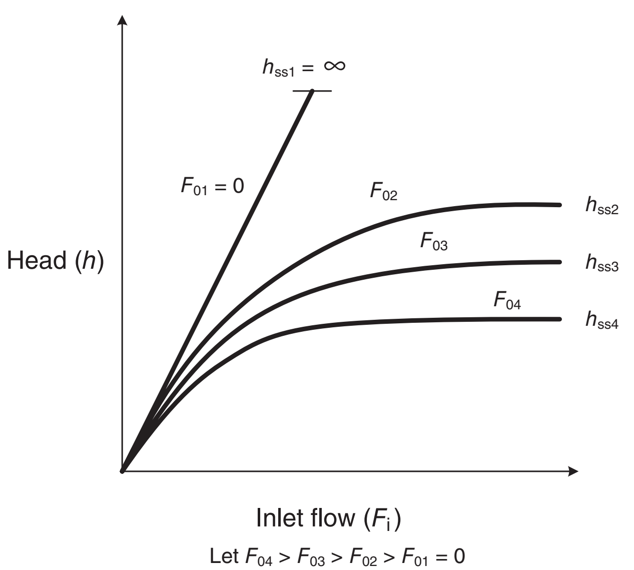

Figure 7.10 *Head versus inlet flow for liquid level process.*

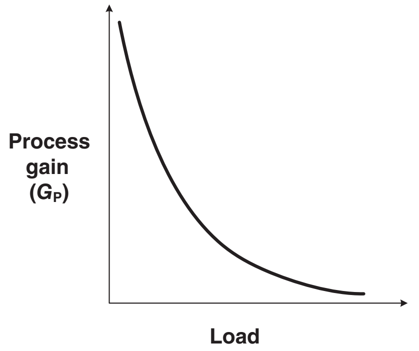

Figure 7.11 Relationship between load and process gain.

so on, 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.

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, that is, fixed value of F_0, 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, the process gain rises and therefore the loop response

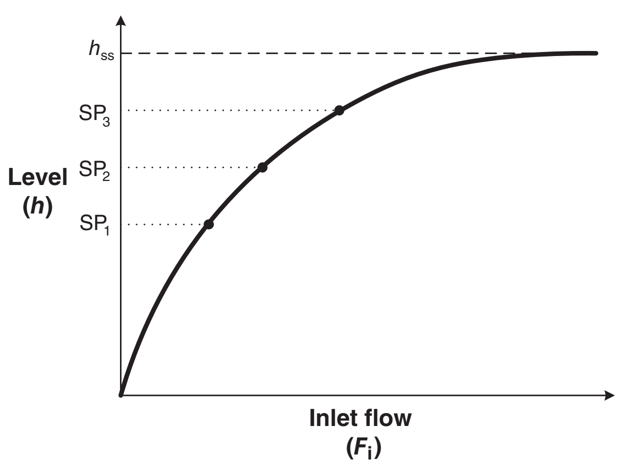

Figure 7.12 Gain and set point relationship.

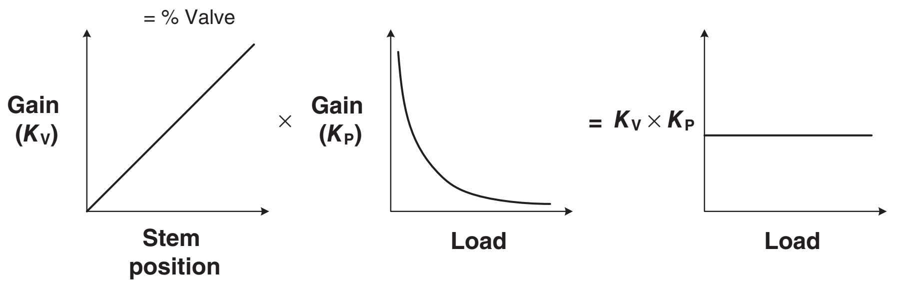

Figure 7.13 Load versus gain.

tends to be more oscillatory. If the process gain increases enough, the loop could become unstable. On the other hand, if the load increases, the process gain decreases, and the loop responds sluggishly. The important fact to remember is that the loop must not become unstable, that is, the loop gain, K_L, must be <1. Therefore, for the situation where the process gain is a function of the load, the simplest thing to do is to 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 ($K_V \times K_p$) 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 a form of adaptive control [5].

Another level control situation where a non-linear element might be introduced into the loop is the case of level control of a cylindrical tank lying on its side, shown in Figure 7.14.

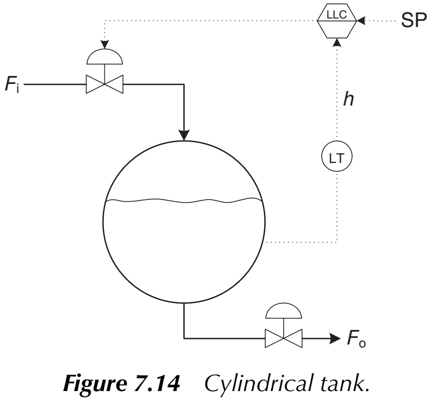

Figure 7.14 Cylindrical tank.

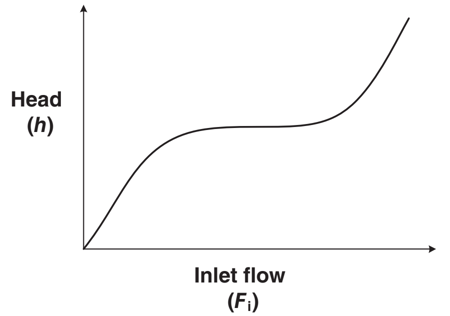

Figure 7.15 *Cylindrical tank level response.*

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 the response for the horizontal, cylindrical tank differs from those previously discussed due to the differences in tank geometry.

Figure 7.16 represents the process gain qualitatively as it varies with load. In this case the load is the height in the tank, *h*. The gain of the tank is high at both low and high levels and is low at normal levels in the 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 which can be varied as shown in Figure 7.17, so that 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.

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.

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, the response might be very slow because the level would change very slowly due

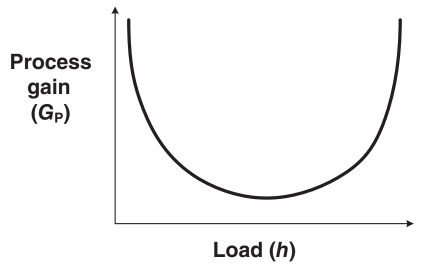

Figure 7.16 *Process gain versus load for a cylindrical tank.*

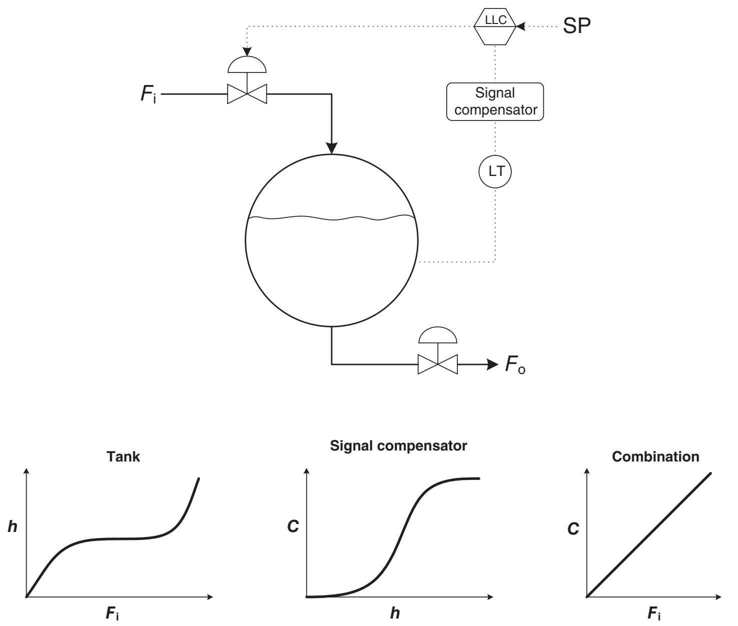

Figure 7.17 *Cylindrical tank level compensator.*

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.

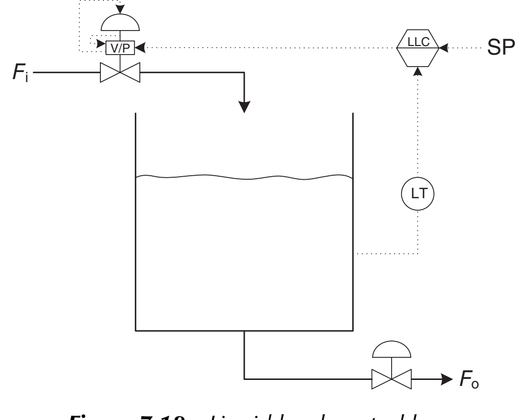

Figure 7.18 *Liquid level control loop.*

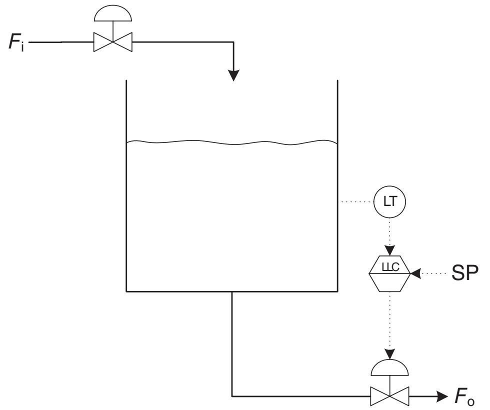

Figure 7.19 *Integrating process – flow smoothing.*

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 and the other uses proportional plus integral action.

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:

$$\begin{aligned} \text{Proportional controller gain} &= 2, \\ \text{Bias} &= 50\%, \\ \text{Set point} &= 50\%. \end{aligned}$$

To set the bias at 50%, put the controller into manual, set the output to 50% and then switch to auto. 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%; *thus the tank cannot run dry.*
2. Output flow reaches the maximum before the level exceeds 75%; *thus the tank level will never exceed 75% for maximum throughput.*

3. Effective buffering is increased when compared to 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% more room, that is, 25/50 or 50%.

Using Equation 4.3 for a proportional-only controller[2] and inserting the above values for the set point, SP, the bias, b, and the controller gain, K_c, and rearranging the results gives

$$\text{mv} = 2(\text{PV}) - 50. \tag{7.8}$$

Therefore, at a 25% level, the output is 0%, and for 75% level, the output is 100%. 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%. For a higher than normal throughput, say mv = 95%, the level would line out at about 72.5%. 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; and
- 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, that is, small hold-ups. 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, 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 hold-up 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

[2] For a direct acting controller, $e = \text{CV} - \text{SP}$.

in Equation 7.9 is the volume of the tank between minimum and maximum controlled levels and not the total tank volume:

$$T_{\mathrm{HU}} = \frac{\text{Volume}[\,\text{ft}^3\,]}{\text{MaxFlow}\left[\dfrac{\text{ft}^3}{\text{min}}\right]} = \frac{V}{Q_{\max}}. \tag{7.9}$$

3. Calculate the integral time using Equation 7.10:

$$T_{\mathrm{i}} = 4\left(\frac{T_{\mathrm{HU}}}{K_{\mathrm{c}}}\right). \tag{7.10}$$

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, due 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. Due 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.

The gain of this process is a function of the load, F_0. 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.

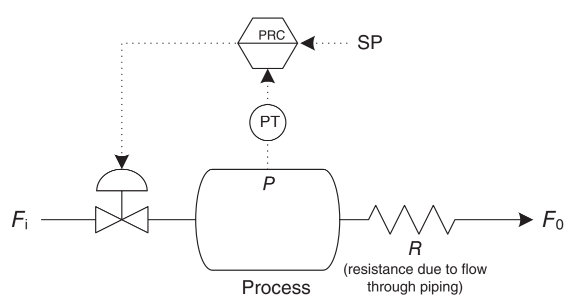

Figure 7.20 *Gas pressure control loop.*

7.5 Temperature Control Loops

Temperature loops may be divided into two main categories:

1. Endothermic – requiring heat energy.
2. Exothermic – generating heat energy.

Both of these processes have similar characteristics in that they are typically comprised of one large and many small capacities, that is, valve actuator, transmitter and so on. 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); and
- Resistance temperature detector (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.

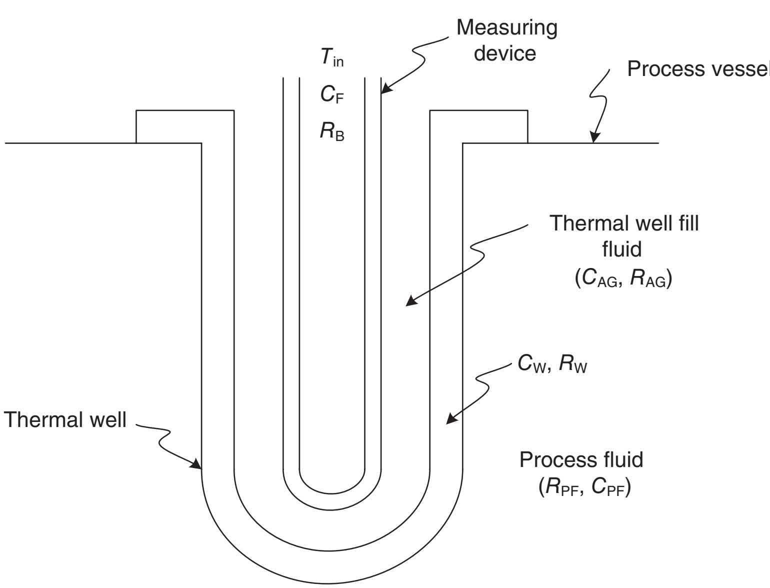

Figure 7.21 *Typical temperature measurement device installation.*

Every component in the measuring systems, 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, $\tau_{PF} = R_{PF}C_{PF}$. The other time constants (thermal well, $\tau_w = R_w C_w$; thermal well fluid, $\tau_{AG} = R_{AG}C_{AG}$; and measuring device, $\tau_B = R_B C_F$) 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 high thermally 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 a 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, while the TC is somewhat faster. For a TC, 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 seconds.

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 due to its greater internal resistance. But for the same size bulb, either a liquid FTS or RTD has a time constant of about 3 seconds.

For an FTS the capillary has a time constant of 0.55 seconds per 10 feet, and thus it is obvious why a transmitter is often used since some capillaries can be up to 100 feet 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 time of RTD or FTS bulbs in a thermal well depends 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 minutes, while 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 due 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, that is, make τ_{AG} as small as possible.

If it is necessary to use a thick metal or ceramic well due to corrosive process fluid, 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 minutes. This increase can be detrimental in certain processes, that is, 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 FTS 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 slow as well as 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 seconds 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 from Equation 7.11:

$$K_{\mathrm{p}} = \left(\frac{\Delta T_2}{\Delta F_{\mathrm{S}}}\right)_{F_{\mathrm{W}}=\mathrm{Constant}}. \tag{7.11}$$

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:

$$F_{\mathrm{S}} = KF_{\mathrm{W}}(T_2 - T_1), \tag{7.12}$$

$$K = \frac{C_{\mathrm{p}}}{\lambda}, \tag{7.13}$$

where C_{p} is the specific heat of F_w; λ is the heat of condensation; F_{w} is the flow of cold fluid (load variable); T_1 is the inlet temperature of cold fluid (load variable); and T_2 is the outlet temperature of cold fluid (load variable).

The heat exchanger energy balance equation can be solved for the heat exchanger gain, K_{p}, as shown in Equation 7.14:

$$K_{\mathrm{p}} = \frac{\mathrm{d}T_2}{\mathrm{d}F_{\mathrm{S}}} = \frac{1}{KF_{\mathrm{w}}} = \frac{K'}{F_{\mathrm{w}}}. \tag{7.14}$$

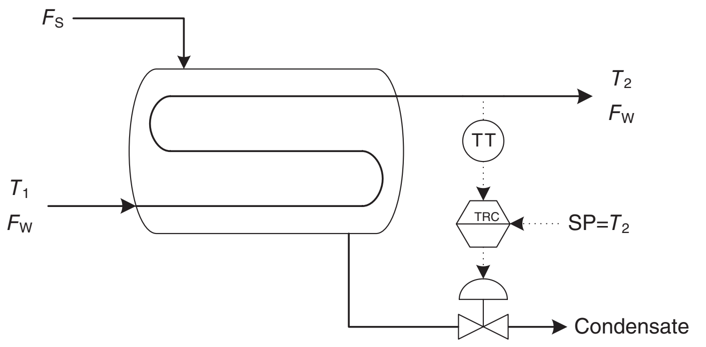

Figure 7.22 *Temperature control via level control (Courtesy of REM Technology Inc -Spartan Controls Ltd).*

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, 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, 7.23, 7.24 and 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, that is, overhead condenser. When the temperature is too high, the valve closes, which causes condensate to cover more tubes and reduce heat transfer to the cold fluid. Because the condensation time is large, the response is slower than for other systems. Also, due to 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 of the control scheme of 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 a uniform T_1. The energy penalty for this scheme is the cooling of the stream the energy of which has been expended in heating up.

The scheme shown in Figure 7.25 throttles F_w, to maintain T_2, and is usually used in 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.

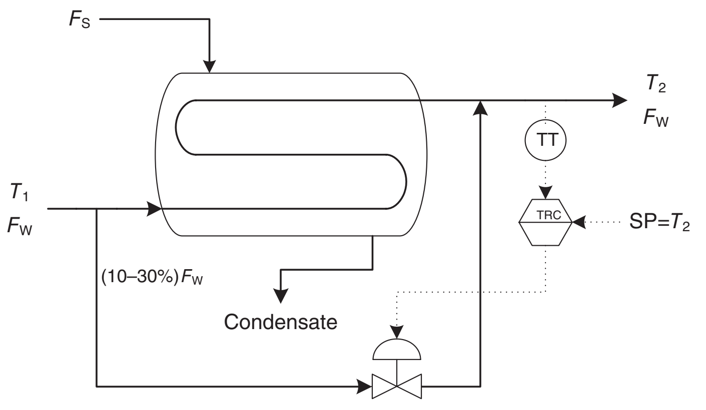

Figure 7.23 *Control scheme for critical temperature control.*

A detailed discussion of tube and shell, aerial coolers and fired heaters is found in the Hydrocarbon Processing articles by W.C. Driedger [9, 10].

7.5.2 The Exothermic Reactor Temperature Control Loop

The exothermic reactor is perhaps the most difficult process to control due to its instability and extreme non-linear response [8]. A chemical reactor is quite often an exothermic process where some feedstock and catalyst are mixed together, and the 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 to 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

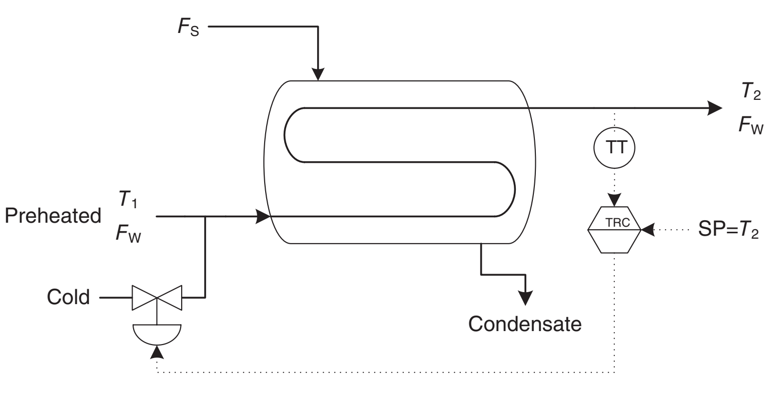

Figure 7.24 *Variation of critical temperature control scheme.*

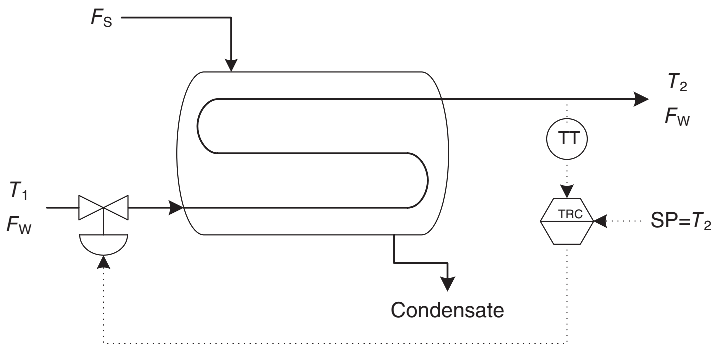

Figure 7.25 *Temperature control scheme for capacity-limited exchangers.*

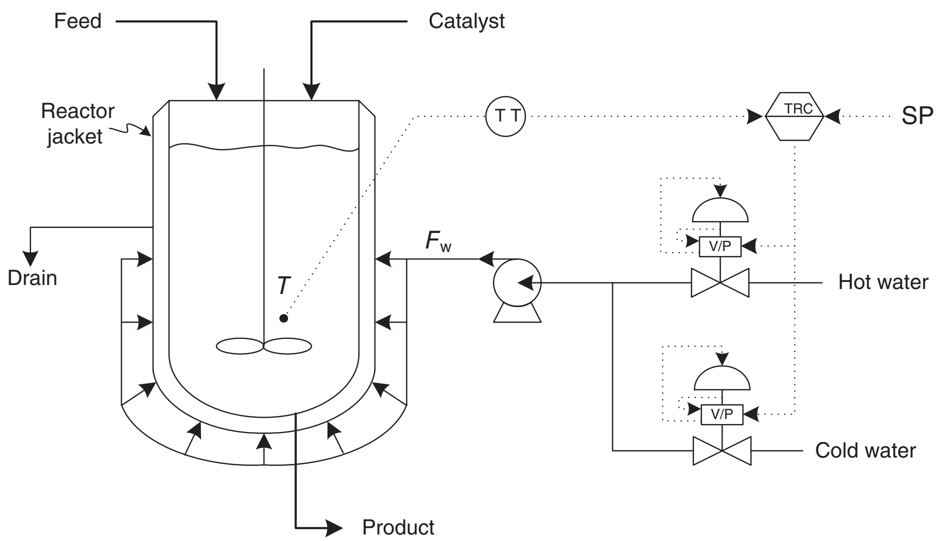

Figure 7.26 *Control scheme for an exothermic reactor.*

hot water valve, which is air to open and opening the coldwater 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, that is, 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.[3]

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

[3] Similar to a gas pressure loop.

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[4] 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 References [3, 8].

In addition to these problems there is also the problem of the extreme non-linear process response. As before, the gain is a function of the temperature operating level. For the control system shown, the gain is defined by Equation 7.15:

$$K_\mathrm{p} = \frac{\Delta T}{\Delta F_\mathrm{w}}. \tag{7.15}$$

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:

$$K_\mathrm{p} \approx \frac{Q}{F_\mathrm{w}^2}. \tag{7.16}$$

Due to this non-linearity as well as to the problems mentioned earlier, some exothermic reactors are controlled with advanced control techniques such as feedforward, 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 and 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 million dollars.

Two major types of compressors are commonly used in chemical and petroleum plants, reciprocating and centrifugal compressors.

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[4] Override controls.

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, *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 outwards. The centrifugal force develops a pressure proportional to the density and to the speed squared. Suction throttling reduces the density and hence ΔP; thus the machine operates as a constant pressure ratio machine.
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 done.

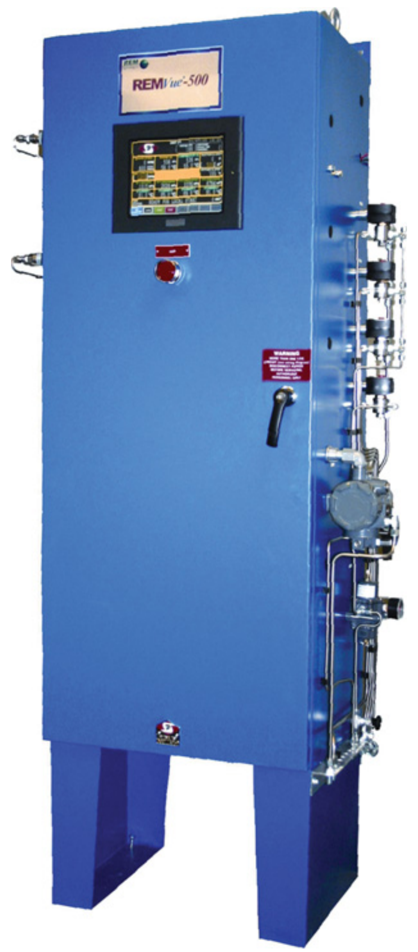

Figure 7.27 *A REMVue Compressor Control Station (Courtesy of REM Technology Inc.-Spartan Controls Ltd.).*

3. *Recirculation*: Recirculation has a much lower efficiency for similar reasons, but is essential for low turndown (low flowrates).
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 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

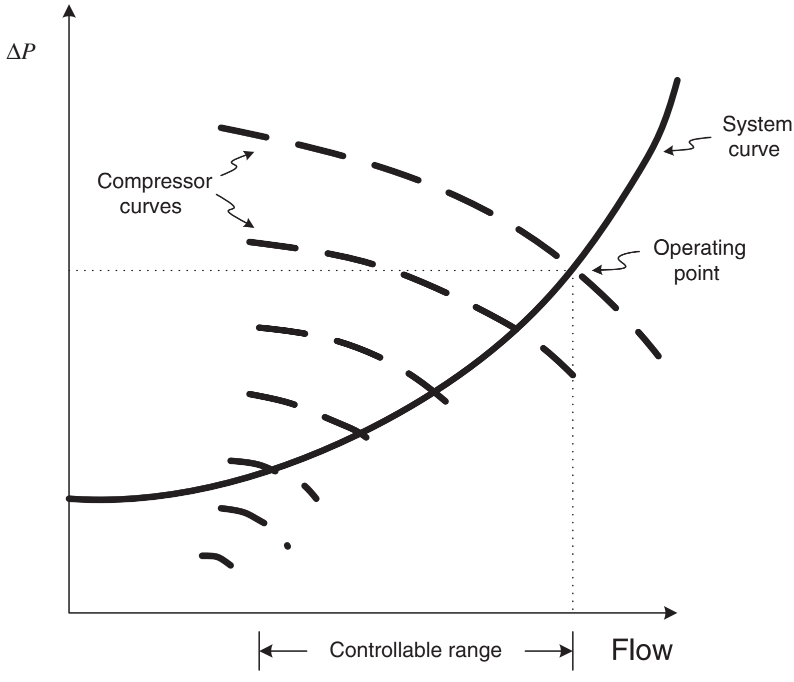

Figure 7.28 *A typical compressor map.*

curve and a system curve, specific operating points occur which collectively establish the controllable range.

If the load on the compressor changes – that is the system curve changes – as shown in Figure 7.29, 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).

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, 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].

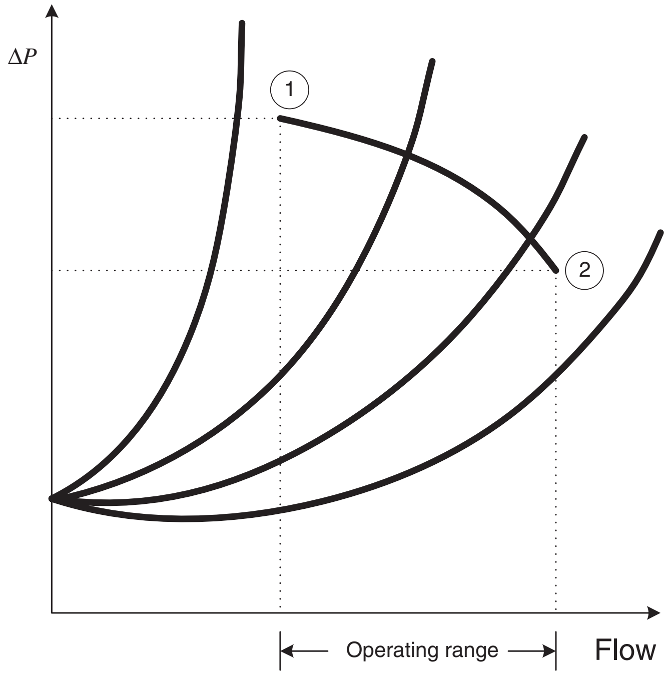

Figure 7.29 Compressor load changes.

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, that is, 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, *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

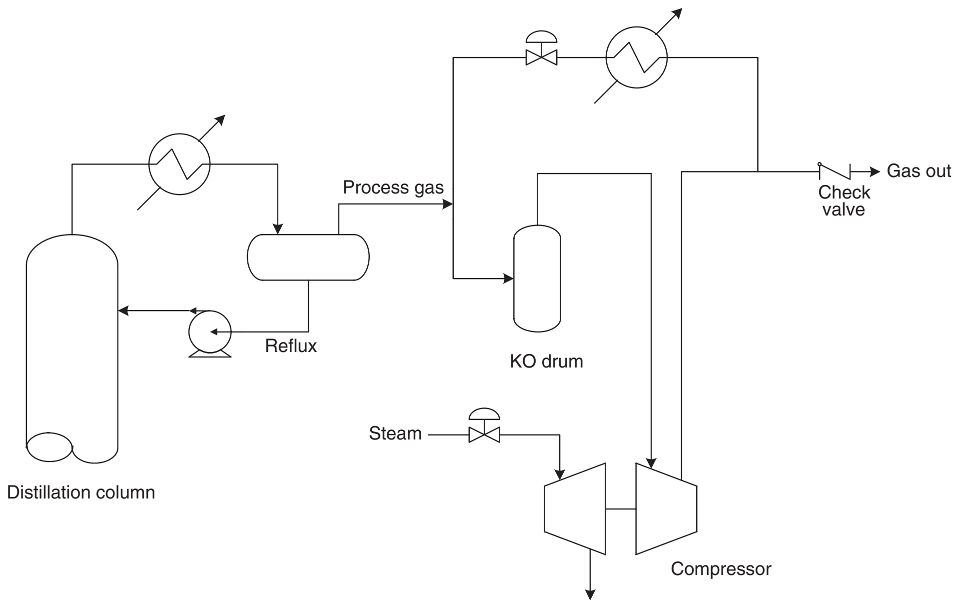

Figure 7.30 *Compressor control example process schematic.*

by speed control on the steam turbine. Steam turbines generally have 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 a 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 to 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 result 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, suction flow and ΔP must be measured. Suction flow must be in terms of actual, not standard volume units at the inlet. 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 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% 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.

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

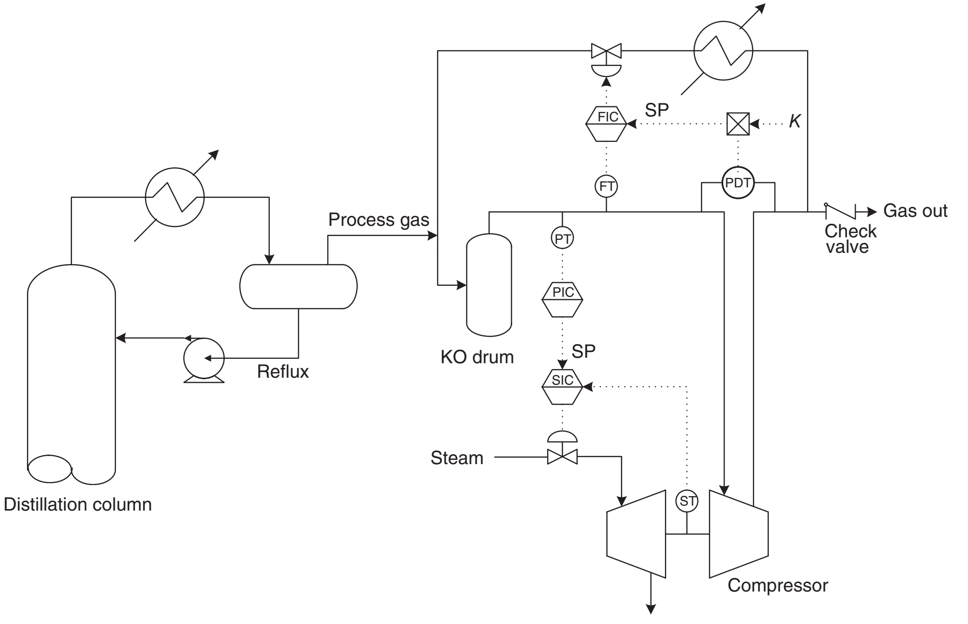

Figure 7.31 *Complete compressor control system schematic.*

[11, 19, 20] include a substantial emphasis on the importance of dynamic simulation for control scheme design validation and performance evaluation.

Compressors' Performance Optimization with Air–Fuel Ratio and Slipstream Control
Compressors are powered by electric motors or reciprocating internal combustion engines (RICEs), which supply the load demanded by the compressor at the desired speed. RICEs require fuel flow control to achieve and maintain the desired speed. Speed is controlled by a governor that measures the rotational speed, compares to the desired speed and increases or decreases the fuel supplied to the engine. Electronic governors are replacing mechanical governors because of greater precision and reduced maintenance needs. In the electronic governor, the speed is measured by pulses generated in an induction coil or Hall-effect pickup by flywheel gear teeth or a flywheel stud. These pulses are converted to an analog signal. This analog signal is compared to an analog voltage proportional to the set point or desired speed by a PID (proportional, integral and derivative) software algorithm in the controller. If the engine speed is less than the set point speed, the PID analog output to a fuel valve actuator is increased to increase the fuel flow to the engine. The reverse occurs for an excess speed. The factors controlling the relative amounts of proportional, integral and derivative functions can be adjusted to ensure an optimum response to changes in compressor load or desired speed.

Air control is necessary for both spark-ignited (SI) engines and compression-ignited (CI) engines to achieve regulatory exhaust emissions control. In the past no air control was used for CI engines and a carburettor was commonly used for air to fuel ratio control. The advent of emissions regulations, particularly for oxides of nitrogen, NOx, has led to the predominant use of electronic control of the engine air. Emissions control can be very complex in detail. In summary, the amount of air supplied to the engine is controlled by a valve actuator. A software algorithm determines an air requirement according to the engine speed, load, air temperature and so on and provides an electrical signal to an air actuator. For many SI engines the NOx emissions are reduced in an exhaust catalyst. To ensure these catalysts achieve the desired emissions reductions, very precise control of the air to fuel ratio is required.

In the natural gas industry the majority of gas compressors are powered by SI RICEs with natural gas fuel. All gas compressors have leaks at rotational or sliding seals. In a novel control application, the natural gas from these leaks is collected and added to the engine air to supplement the normal engine fuel. Since the compressor leak rates are variable in time and magnitude, the addition of this gas to the engine air represents a control challenge. A commercial control system, known as SlipStream, is able to achieve the engine control necessary to both meet emissions regulations and save fuel costs.

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 dangerous as it may explode if the air flow is increased. If there is too much air, 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 0.5–2% 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 build-up in the system [12].

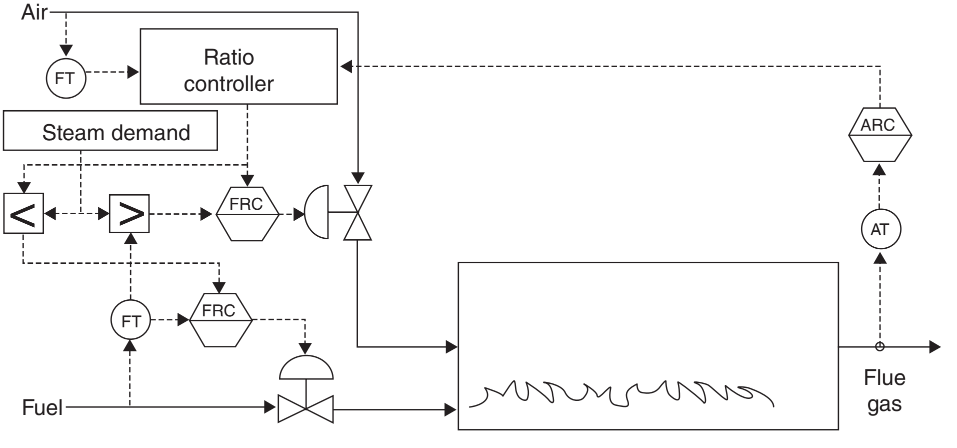

Figure 7.32 Combustion control scheme.

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, 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 lift more liquid into the water drum, increasing the level. As the drum level increases, a controller will add less feed water 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, 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. 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 the demand for steam is low, 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 superheated steam must be maintained at a constant value, feed water is added to keep the temperature from getting too high (Figure 7.33). Although this feed water will decrease boiler efficiency it is important for improved temperature control.

Control must always be balanced to ensure safety and maximize benefits while minimizing losses.

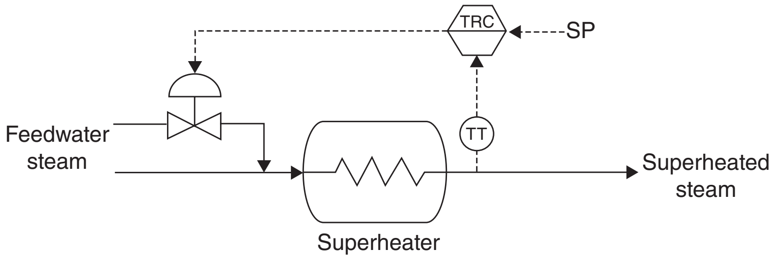

Figure 7.33 *Control scheme for superheated steam.*

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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 [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

A Real-Time Approach to Process Control, Third Edition. William Y. Svrcek, Donald P. Mahoney and Brent R. Young.

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 assumption is that the controller structure will be built up around conventional PID controllers, ratio, feedforward and override control blocks found in all commercial distributed control systems (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, that is, 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, 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

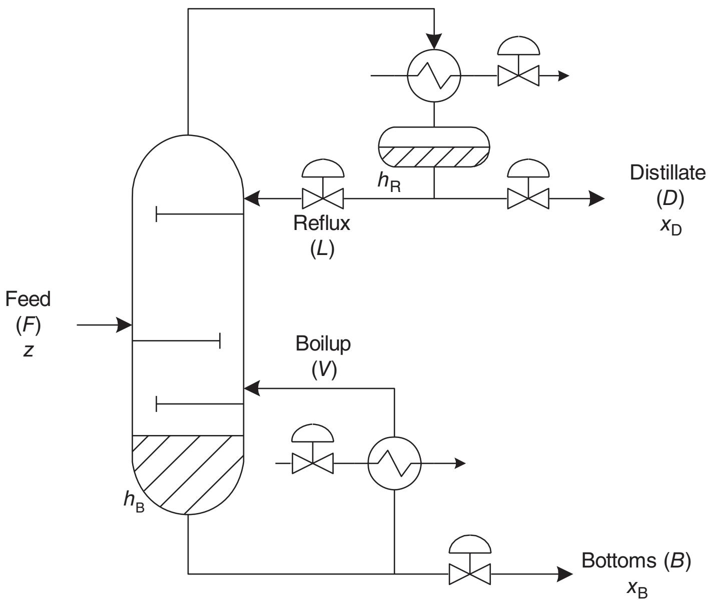

Figure 8.1 *Basic distillation column schematic.*

than composition control. For the same reason, any three of the five manipulated variables can be used to control the column inventories.

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. As well, if it is important to control the inventory of a *trapped* component, such as an entrainer for azeotropic distillation, 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 equal the number of valves available for control in that section of the plant. To find out how many integrating variables, that is, 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 on 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 response 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 change with a steeper slope representing better separation. It is important to realize that fractionation increases the purity of both products simultaneously while changing the feed split will make one product more pure and the other less pure.

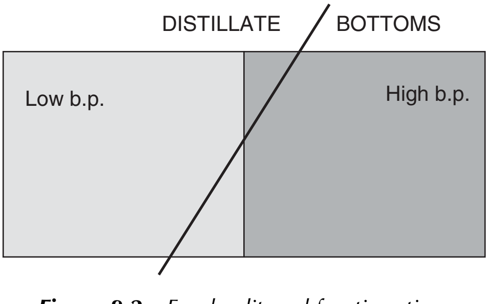

Figure 8.2 Feed split and fractionation.

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 while 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 while 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.

Overall material balance:

$$F = D + B. \tag{8.1}$$

Component balance:

$$F\,x_{\mathrm{f}} = D\,x_{\mathrm{D}} + B\,x_{\mathrm{B}}, \tag{8.2}$$

where F is the feed; B is the bottoms; D is the 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.

Energy balance:

$$FH_{\mathrm{f}} + Q_{\mathrm{reb}} = DH_{\mathrm{d}} + BH_{\mathrm{b}} + Q_{\mathrm{cond}}, \tag{8.3}$$

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, 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 non-linear. 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

$$\left(\frac{\partial x_{\mathrm{D}}}{\partial D}\right)_{x_{\mathrm{B}}} = -\frac{F\,(x_{\mathrm{f}} - x_{\mathrm{B}})}{D^2}. \tag{8.6}$$

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 as 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 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 a 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 single input/single output (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 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 hold-up). 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.

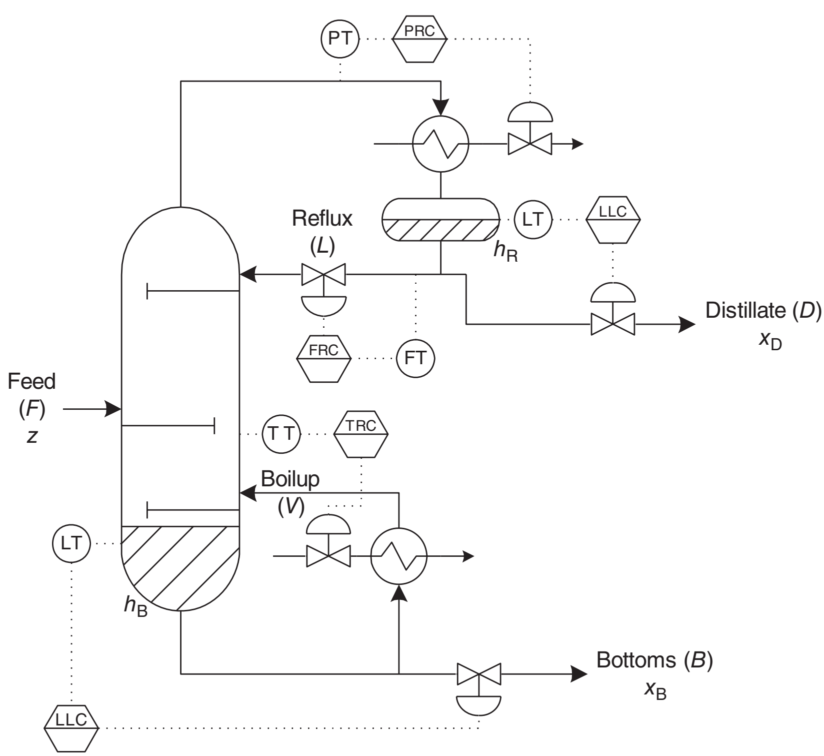

Figure 8.3 *Column basic control scheme.*

The most convenient method of verifying the operability of a proposed multivariable control scheme is through dynamic simulation. However, to effectively use dynamic simulation it is first necessary to define the objectives of the control system, define the nature of the expected disturbances and develop a basic understanding of the process in terms of both 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. It is therefore 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, that is, 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 Chapter 7 in the liquid level control section. Typical hold-up times for condenser accumulators and reboilers are of the order of 5–10 minutes and 20 minutes (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% 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.

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, 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 of this scheme to changes in the coolant flow rate is inherently slow in comparison to 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 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

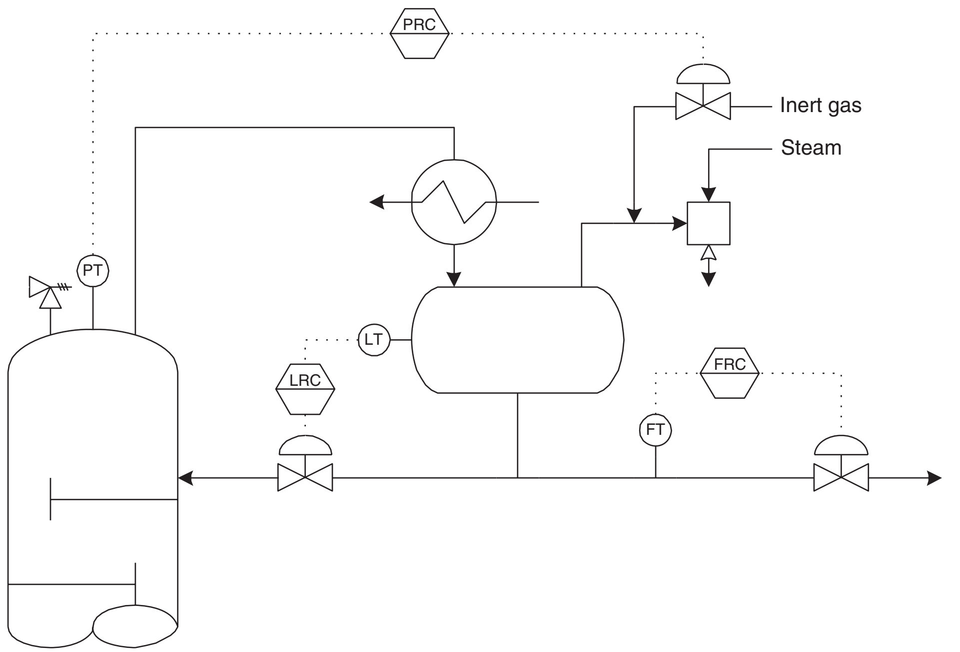

Figure 8.4 *Pressure control for sub-atmospheric operations.*

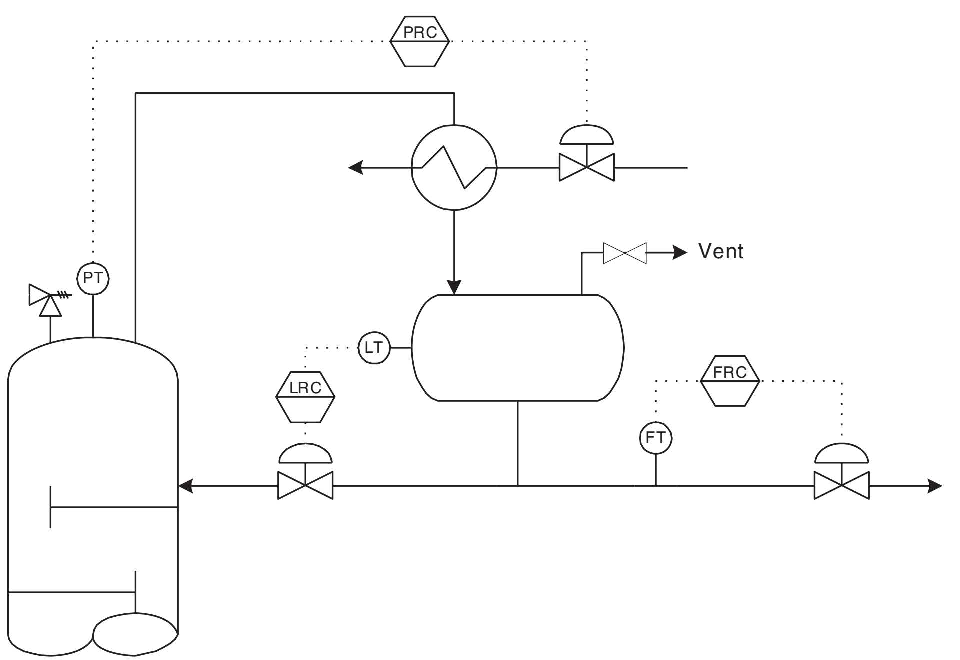

Figure 8.5 *Pressure control for above atmospheric operation.*

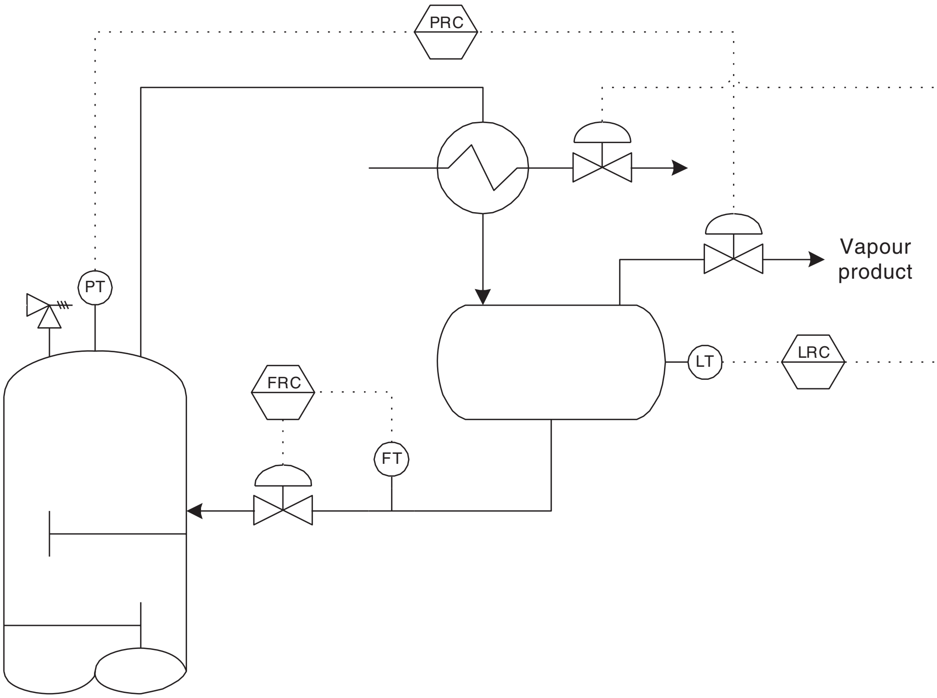

Figure 8.6 *Pressure control by control of overhead product vapour flow.*

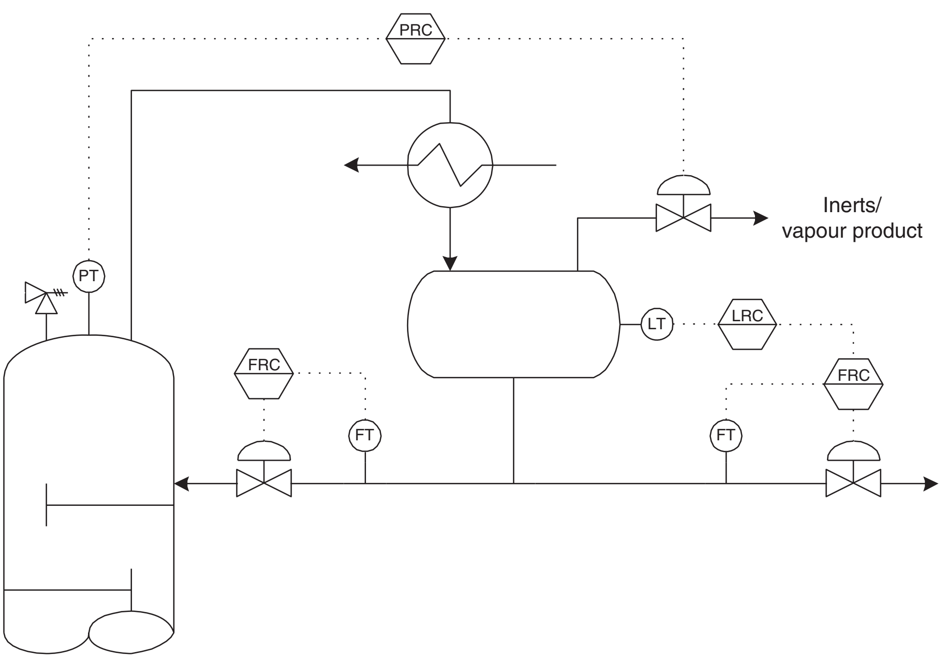

Figure 8.7 *Pressure control by venting of inerts.*

subject to change. A flow controller fixes the reflux rate while 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 8.9. The column pressure is controlled by either injecting or venting blanketing gas from the accumulator/condenser.

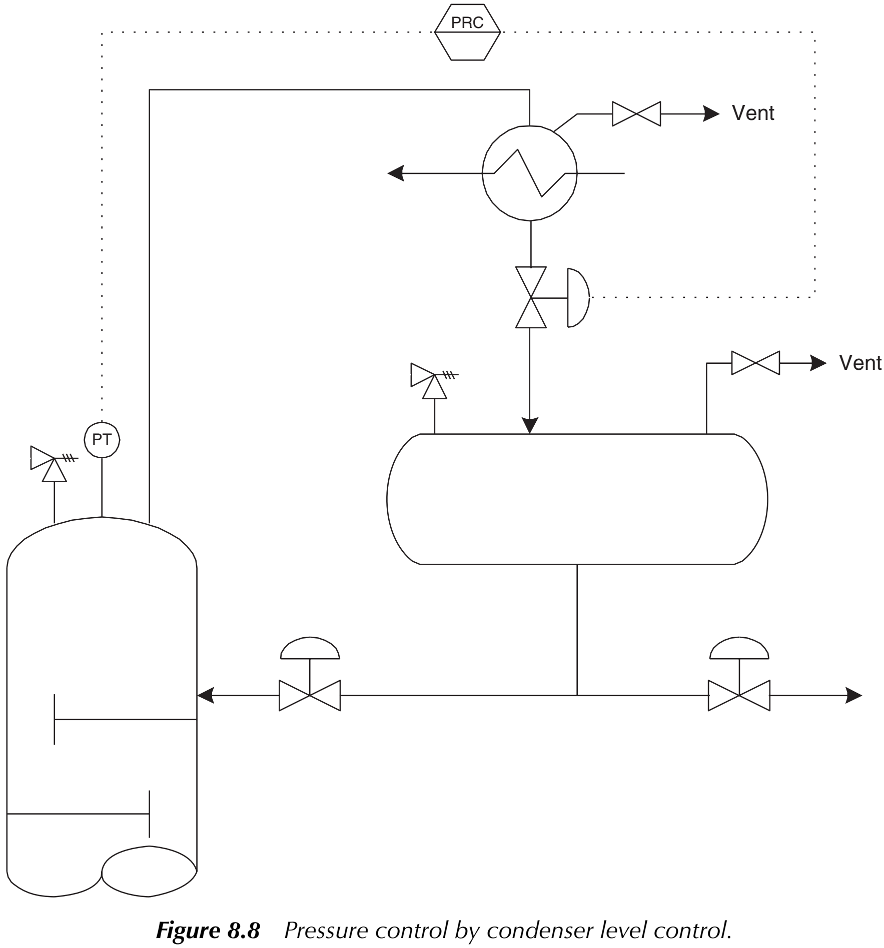

Figure 8.8 *Pressure control by condenser level control.*

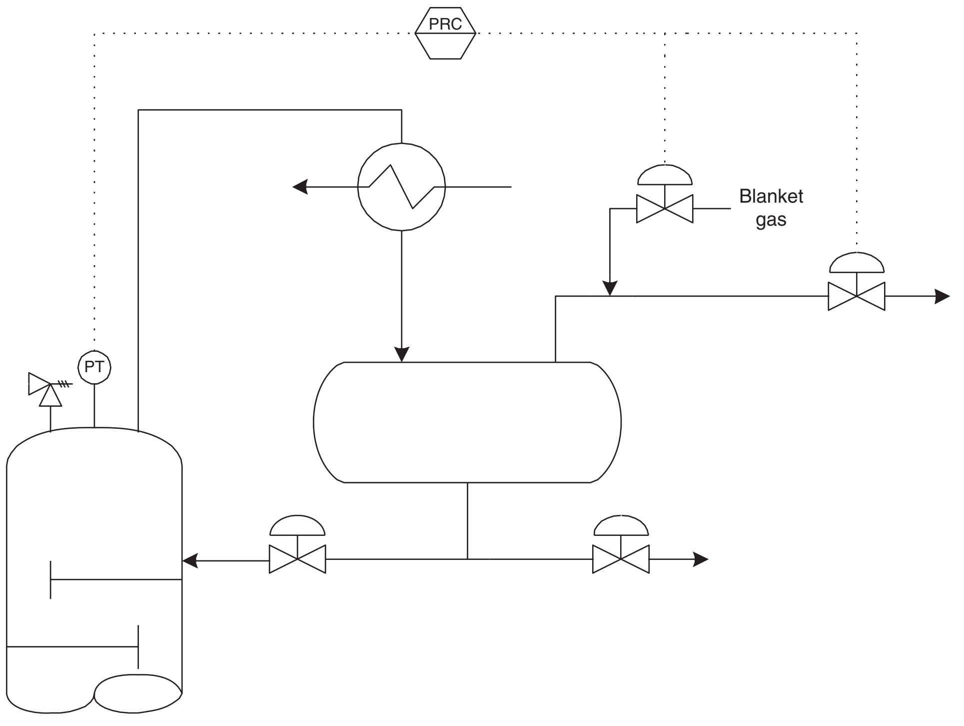

Figure 8.9 *Pressure control by regulating condenser surface area with blanket gas.*

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].

Some of the factors that should be considered in locating the pressure measuring point are as follows:

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 to 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 5–10 minutes. Infrared analysers that produce continuous composition estimates are seeing increased use and response times are of the order of a minute [12]. 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 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 [13]. 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, that is, column feed composition changes. If the magnitude of these changes can be estimated, 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 to 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 to 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 the 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 Sidestream

To better illustrate how the described control strategy design method 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,

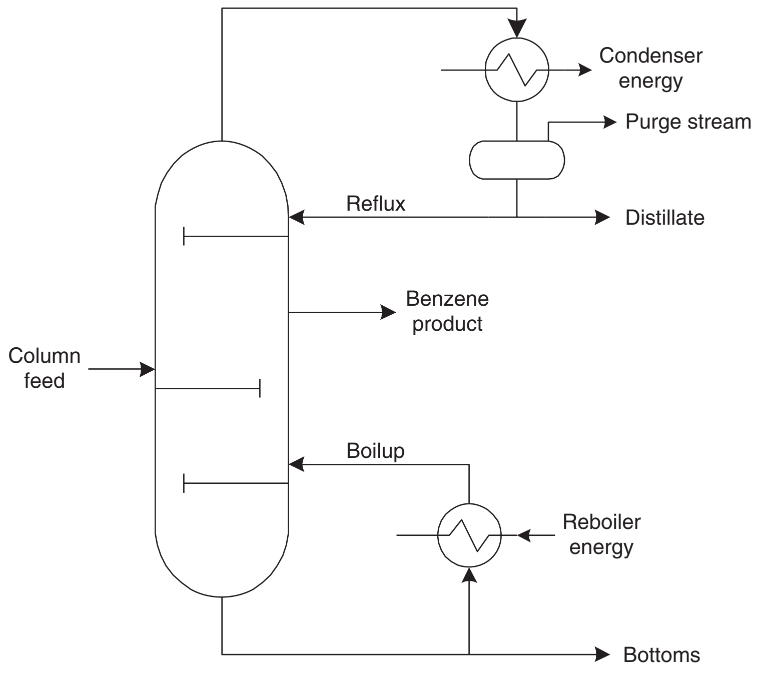

Figure 8.10 *Liquid side draw benzene column.*

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, 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 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.

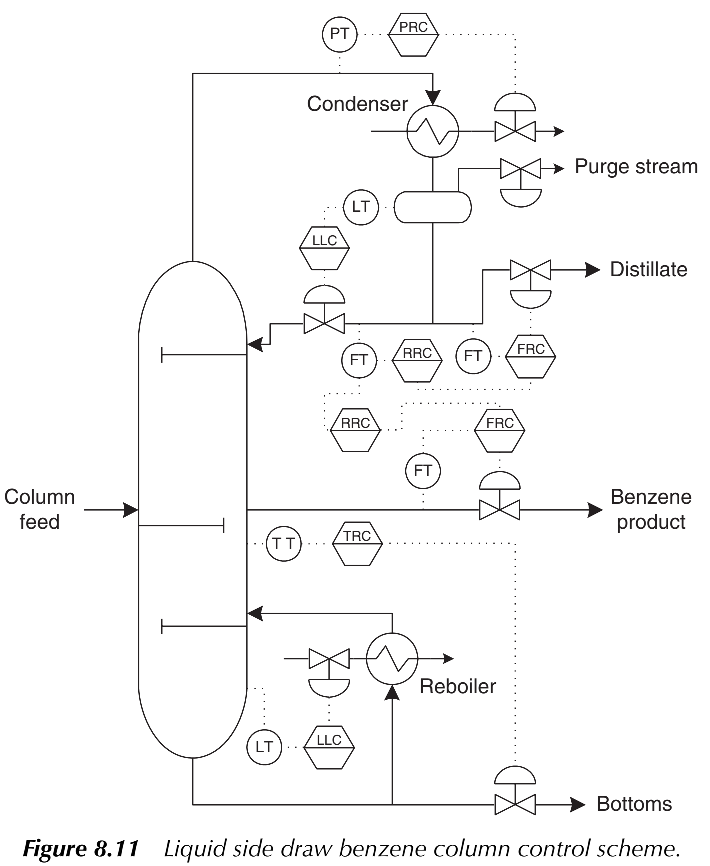

Figure 8.11 *Liquid side draw benzene column control scheme.*

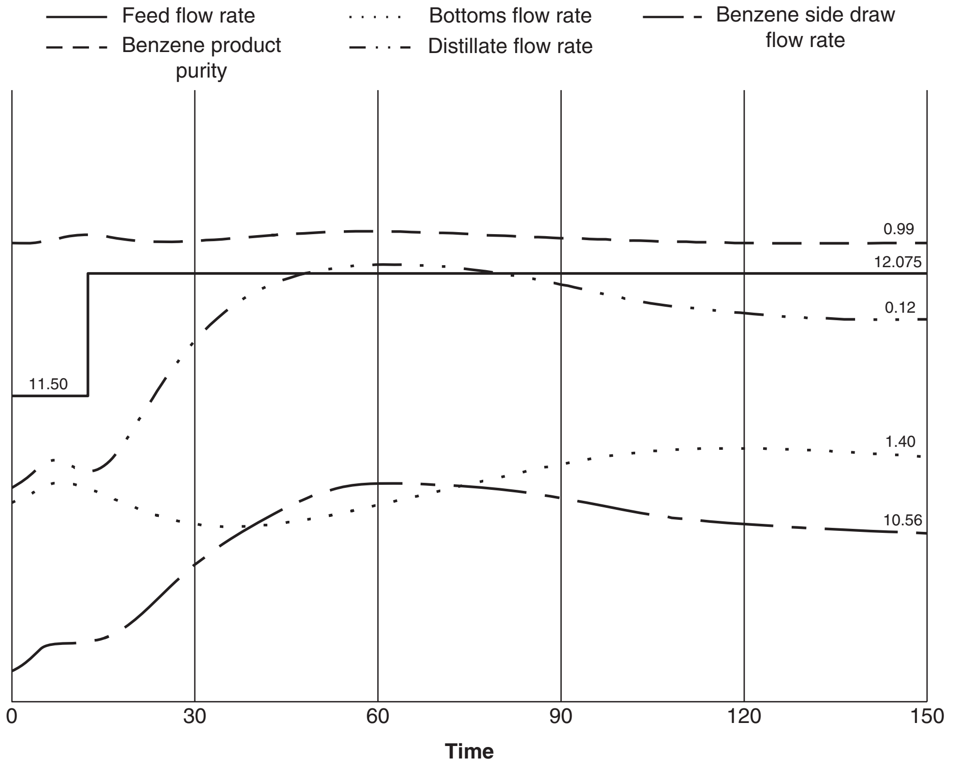

Figure 8.12 *System response to a step change in feed flow rate.*

How does this control scheme respond to disturbances? The major control objective is to produce a side stream of essentially pure benzene, approximately 99%. To test this control scheme, the column was subjected to two different types of disturbances. The first disturbance was that of an increase in the total volumetric flow rate of the feed introduced into 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 to 12.075 m^3/h. This corresponds to an increase of 5%.

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? 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.

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 $\pm 10\%$ over periods ranging from 20 to 30 minutes. 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.

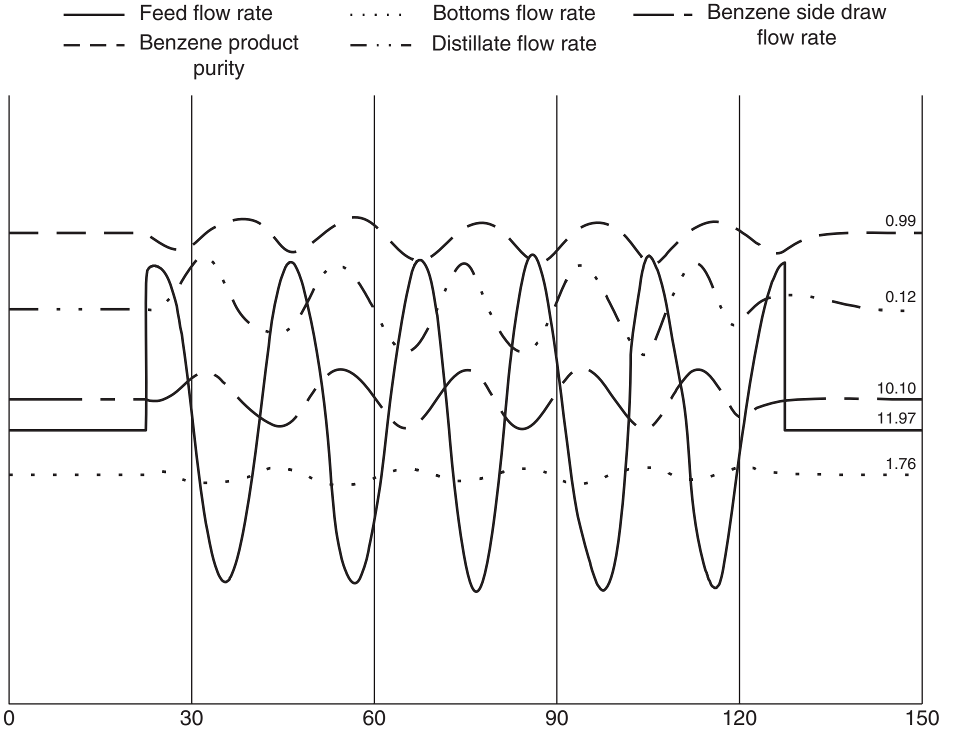

Figure 8.13 *System response to sinusoidal disturbances in the component flow rates.*

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. Fruehauf and Mahoney [11, 14, 15] have shown that steady-state calculations [16] 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 configuration. 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, feed split and fractionation. The resulting best feed split control schemes are shown in Figures 8.14 and 8.15.

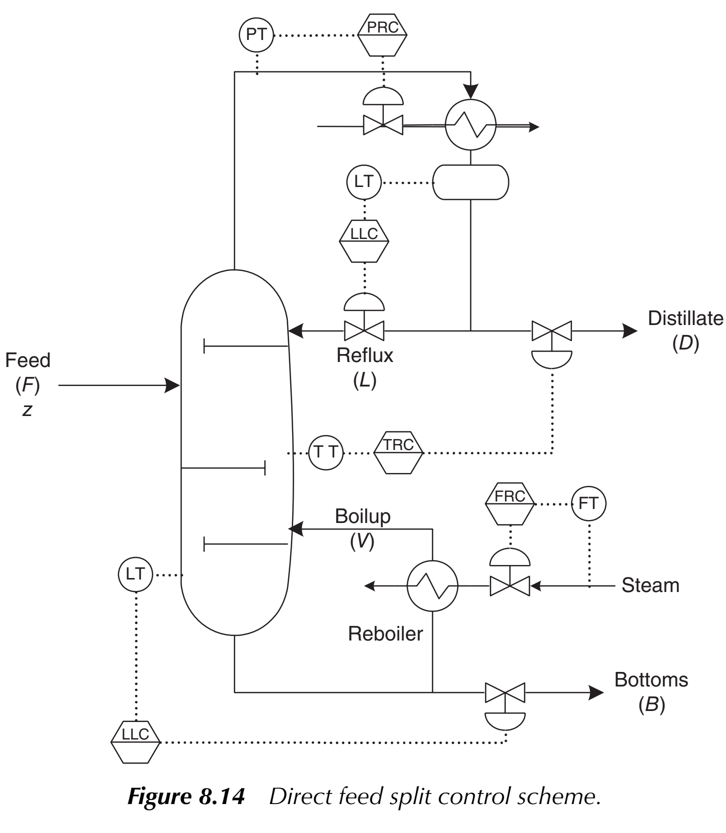

Figure 8.14 *Direct feed split control scheme.*

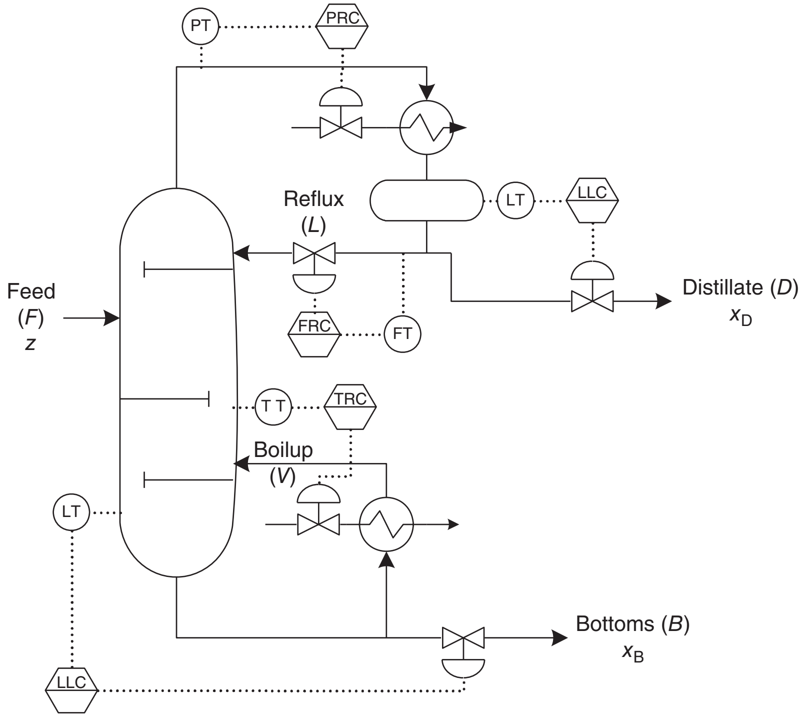

Figure 8.15 *Indirect feed split control scheme.*

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[1] 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 downstream unit operations. To achieve flow smoothing, the level controller must have averaging level controller tuning.

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.

[1] Shorter natural period.

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 varying, in this case, the distillate flow and observing the change in column temperature profile. A good tray is one on which the temperature change was 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 the procedure is complete. If not, the procedure is repeated with another candidate control scheme.

A case study that demonstrates the application of this design scheme is available in the literature [17].

8.7.1 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 (B.J. Cott, Private communication, 2 January 2003). 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 lab 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 a reduced engineering cost against a potential drop in operational flexibility.

8.7.2 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% propane, 40% isobutane, 40% *n*-butane and 15% isopentane. The total flow rate is 40,000 bbl/day 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.

8.7.3 Respecifying Simulation Specifications

Figure 8.16 shows the column specifications in the VMGSim Simulation file as received from the process design engineer. Note that the designer has included several potential specifications for the column, of which only two can be active at any given time. Not shown are extra draw specs that can be specified on other windows within VMGSim. 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 and not in molar units. Because the specifications of 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

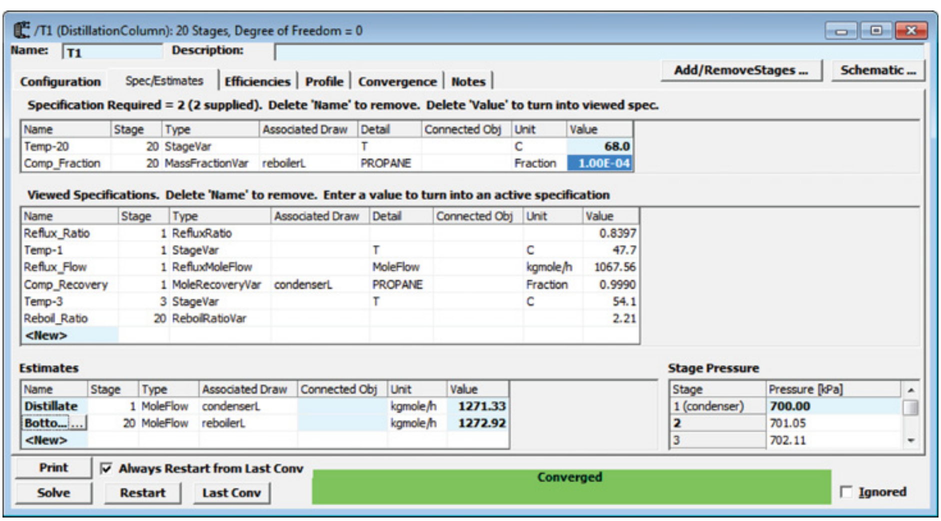

Figure 8.16 *Typical column specifications in VMGSim from process design (With permission from Virtual Materials Group Inc.).*

specifications are also expressed in molar units respectively 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; and
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 flow rate of reflux (in m^3/h) 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 result 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.

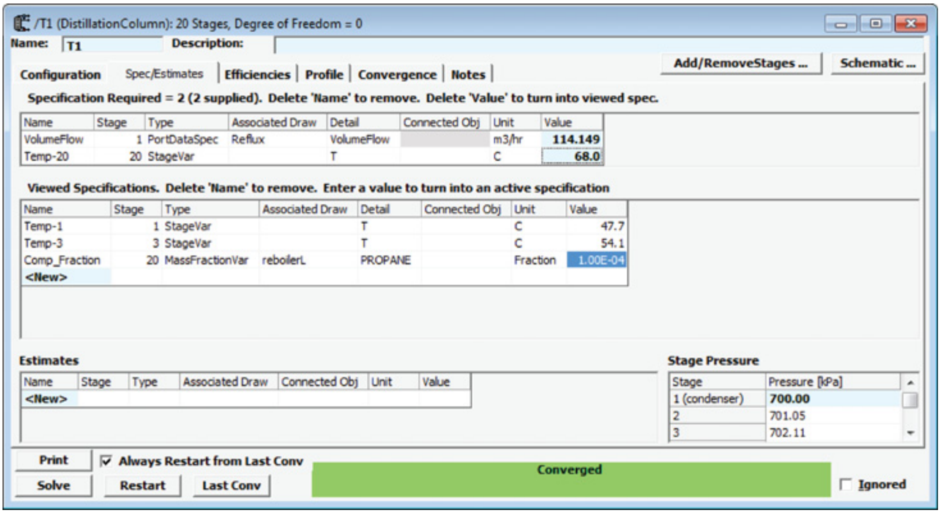

Figure 8.17 *Column respecified to simulate one potential control strategy (With permission from Virtual Materials Group Inc.).*

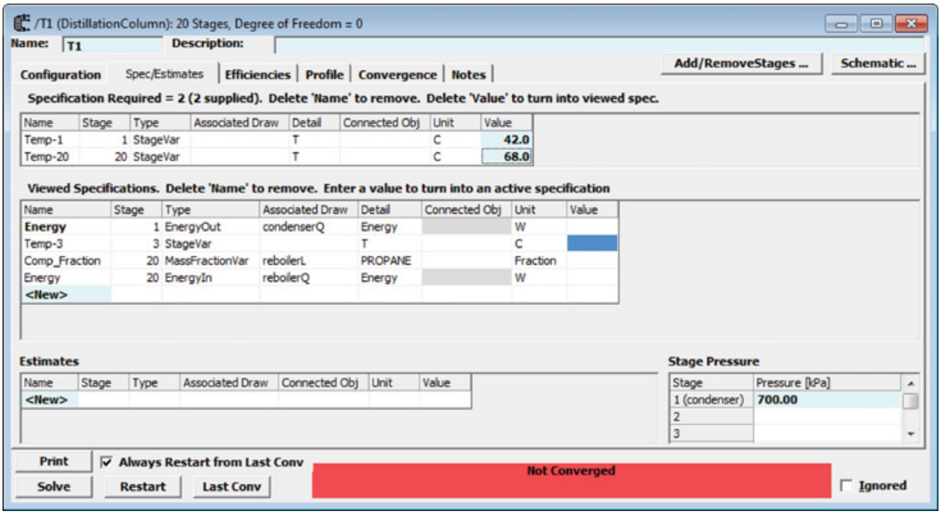

Figure 8.18 *Example of conflicting specifications (With permission from Virtual Materials Group Inc.).*

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.

8.7.4 Mimicking the Behaviour of Analysers or Lab Analyses

The process designers used the component fraction C3 in the bottoms to design the stabilizer. While 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 VMG (RVP Methods Extension to the Special Properties Unit Operation) was used to create a TVP measurement for the stabilizer bottoms stream.

8.7.5 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 bottoms stream, which will be valued at regular gasoline prices. The value of the distillate product is significantly less than that 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, 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 $\times 10^{-6}$/kJ/h.

Computing the change in profitability from the base case operation is the most straightforward way of using the profitability function.

8.7.6 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 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

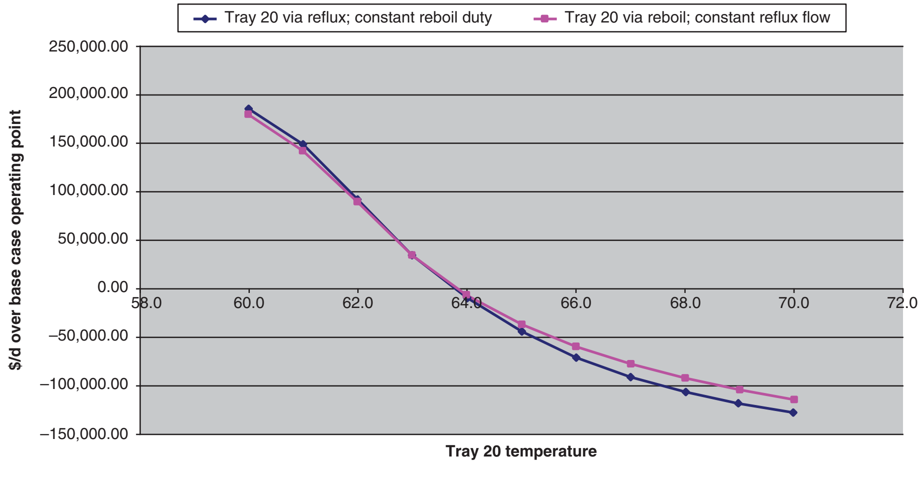

Figure 8.19 Delta profitability curves for candidate control strategies.

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.

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, while the constant reflux flow strategy is preferred at high tray 20 temperatures. However, when the bottoms stream flow constraint of 200 m^3/h 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 of bottoms material (62.1°C vs. 91.9°C). This shows the importance of determining all the important process constraints affecting the operation of the process under investigation.

8.7.7 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 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% propane, 41% isobutane, 40% *n*-butane and 14% isopentane. Remember that the base case feed was 5% propane, 40% isobutane, 40% *n*-butane and 15% 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^3/h of bottoms product (a drop of $18,000/day in profitability) while the constant reflux flow strategy only produced 183.3 m^3/h for the same disturbance (a drop of 26,500$/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.

8.7.8 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 directly control it? 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 $12,6729/day while the profitability of the TVP objective is $12,8611/day. Therefore, this indicates that, for this disturbance, there is a positive economic driver to control the TVP directly of about $1880/day. 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.

8.7.9 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 the following:

1. an accurate economic profitability function for the process and
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 to then 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 [17].

The basic dynamic design procedure consists of the following five steps (which follow upon the steady-state procedure):

1. *Provide information on material holdup*: 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.
3. *Implement control structures*: Tune these controllers.
4. *Practice with dynamic simulation*: Test for stability at start-up, shutdown, runtime, typical upsets and so on.
5. *Repeat steps 8 and 9 for each control strategy*: Compare ease of start-up and shutdown, disturbance rejection, what would happen if there were a change in the rate of production, complexity and interaction between the controllers [15].

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

A case study that demonstrates the application of this design scheme is available in the literature [17].

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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 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.

A Real-Time Approach to Process Control, Third Edition. William Y. Svrcek, Donald P. Mahoney and Brent R. Young.

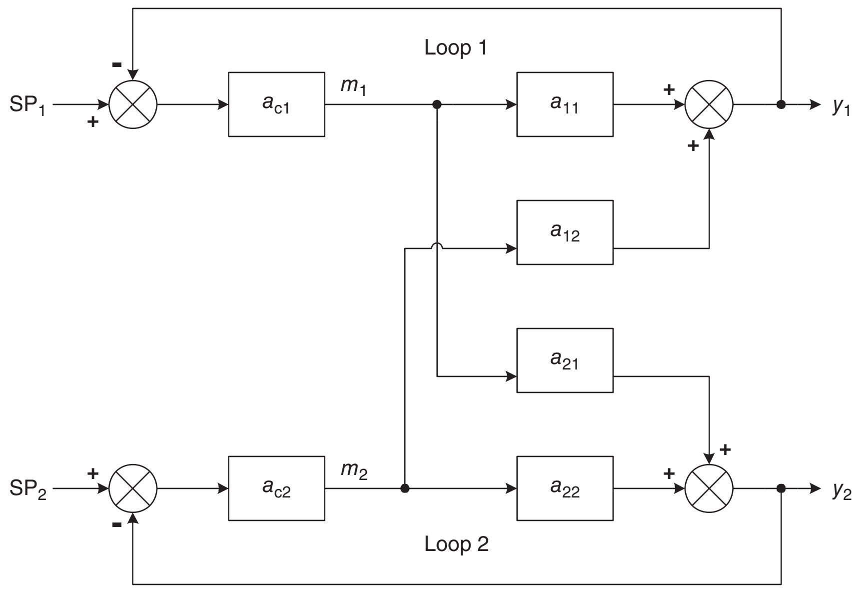

Figure 9.1 *Loop interactions for a 2 × 2 system.*

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 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 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 a quantitative comparison of how one control loop will affect others. Since it compares the effect 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, that is, during process disturbances.

9.2.1 Calculating the RGA 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, as shown in Equation 9.3.

$$g_{11} = {}^{\partial y_1}\!/_{\partial m_1}\Big|_{m_2 = \text{constant}} = \frac{\Delta y_1}{\Delta m_1} = \text{gain}(y_1 - m_1)\,\text{loop with all loops open.} \tag{9.3}$$

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:

$$g_{11}^{*} = {}^{\partial y_1}\!/_{\partial m_1}\Big|_{y_2 = \text{constant}} = \frac{\Delta y_1}{\Delta m_1} = \text{gain}(y_1 - m_1)\,\text{loop with all other loops closed.} \tag{9.4}$$

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):

$$\begin{aligned}\lambda_{ij} &= \frac{\left(\dfrac{\partial y_i}{\partial m_j}\right)_{\text{all loops open}}}{\left(\dfrac{\partial y_i}{\partial m_j}\right)_{\text{all loops closed and in perfect control except the } m_j \text{ loop}}} = \frac{g_{ij}}{g_{ij}^{*}} \\ &= \left(\frac{\textit{open loop gain}}{\textit{closed loop gain}}\right)_{\text{for loop } i \text{ under control of } m_j}.\end{aligned} \tag{9.5}$$

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 RGA 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, $\hat{G}$, of the process can be derived based on the process model. The steady-state gain matrix is defined as follows:

$$\hat{y} = \hat{G}\,\hat{m}, \tag{9.7}$$

where $\hat{y}$ is the vector of output or controlled variables, $\hat{m}$ is the vector of the input or manipulated variables and $\hat{G}$ 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:

$$\hat{G} = \begin{bmatrix} g_{11} & g_{12} & \cdots & g_{1n} \\ g_{21} & g_{22} & \cdots & g_{2n} \\ \vdots & & & \\ g_{n1} & g_{n2} & \cdots & g_{nn} \end{bmatrix}, \tag{9.8}$$

where g_{ij} has been defined in Equation 9.3 as

$$g_{ij} = {}^{\partial y_i}\!/_{\partial m_j} = \text{steady-state gain of } (y_i - m_j) \text{ with all loops open.}$$

Now, let $\hat{R}$ be the transpose of the inverse of the steady-state gain matrix, $\hat{G}$, that is,

$$\hat{R} = \left(\hat{G}^{-1}\right)^{\mathrm{T}}. \tag{9.9}$$

The elements of the RGA can be obtained as follows:

$$\lambda_{ij} = g_{ij} \circ r_{ij}. \tag{9.10}$$

It is important to note that Equation 9.10 indicates an element by element multiplication of the corresponding elements of the two matrices $\hat{G}$ and $\hat{R}$. 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}, 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,

$$g_{ij} = {}^{\Delta y_i}\!/_{\Delta m_j} = \text{steady-state gain of } (y_i - m_j) \text{ with all loops open.}$$

Once the steady-state gain matrix is known it can be manipulated to generate the RGA as described previously.

9.2.3 Interpreting the RGA

The RGA is a useful tool if its properties and limitations are recognized. 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 1:

$$\sum_{i=1}^{n} \lambda_{ij} = \sum_{j=1}^{n} \lambda_{ij} = 1. \tag{9.11}$$

   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 $\lambda_{ij} = 0$, 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 $\lambda_{ij} = 1$, 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 $\lambda_{ij} < 0$, 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 $0 < \lambda_{ij} < 1$, the other control loops are interacting with the (m_j–y_i) control loop. If $\lambda_{ij} = 0.5$, the control pair effect is equal to the retaliatory effect of the other loops. $\lambda_{ij} < 0.5$ indicates the other control loops have a greater influence on the control pair than it has on itself. $\lambda_{ij} > 0.5$ indicates 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 $\lambda_{ij} > 1$, the open loop gain of the pair (m_j–y_i) is greater than the gain when all other loops are closed. This indicates 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 $\lambda_{ij} \gg 1$.

RGA Pairing Rules

There are some basic rules that should be followed to obtain optimal pairing in control loops:

RGA rule 1: Pair 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.

9.3 Niederlinski Index [6]

The NI 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, 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 described in Equation 9.7:

$$\hat{y} = \hat{G}\,\hat{m},$$

where each element in $\hat{G}$ is rational and open loop stable [3]; the system will definitely be unstable if the NI is negative, that is, if

$$\mathrm{NI} = \left. \frac{|\hat{G}|}{\prod_{i=1}^{n} g_{ii}} \right|_{\mathrm{SS}} \tag{9.12}$$

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). Grosididier [4] 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.

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 include restructuring the pairing of variables, detuning the offending control loops to minimize interactions, opening some loops (manual control) and using linear combinations of manipulated and/or controlled variables [2].

SVD [2, 3] is a useful tool to determine if 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, that is, m_1 controls y_1, m_2 controls y_2 and so on, 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, 5] matrix:

$$G = U\Sigma V^{\mathrm{T}}, \tag{9.13}$$

where U is the matrix of normalized eigenvectors of GG^{T}, V is the matrix of normalized 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:

$$\hat{G} = \begin{bmatrix} g_{11} & g_{12} \\ g_{21} & g_{22} \end{bmatrix}. \tag{9.14}$$

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]:

$$\mathrm{CN} = {}^{s_1}\!/_{s_2}. \tag{9.21}$$

For example, if the CN number were equal to 100, this would indicate that one manipulative variable has 100 times more effect on the system than the other manipulative variable. The higher the CN number, the more difficult it becomes to decouple a control loop interaction. A rule of thumb is 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 properly tune the system. 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 minimize interactions between control loops. This entails decreasing the controller gains and increasing the integral times [3]. Dynamic simulation can then be used to drastically reduce the time required 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, a 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 the SVD will be used to test if the control loop interactions are overly sensitive to slight errors in process gains.

9.6.1 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.

The equations, or process model, to describe this system are

$$y_1 = m_1 + m_2, \tag{9.22}$$

$$y_2 = \frac{T_1 m_1 + T_2 m_2}{y_1}, \tag{9.23}$$

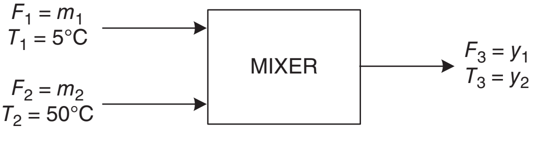

Figure 9.2 *Mixer control.*

where T is in Kelvin and the specific heat capacity of the streams are assumed constant.

At steady state the following values are maintained:

$T_1 = 5°\text{C} = 278.15$ K,
$m_1 = 200$ kg/h,
$T_2 = 50°\text{C} = 323.15$ K,
$m_2 = 400$ kg/h,
resulting in
$y_1 = 600$ kg/h,
$y_2 = 35°\text{C} = 308.15$ K.

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:

$$g_{11} = {}^{\partial y_1}\!/_{\partial m_1}\Big| m_2 = \text{constant} = \frac{\partial}{\partial m_1}[m_1 + m_2] = 1, \tag{9.24}$$

$$g_{21} = {}^{\partial y_2}\!/_{\partial m_1}\Big| m_2 = \text{constant} = \frac{\partial}{\partial m_1}\left[\frac{T_1 m_1 + T_2 m_2}{m_1 + m_2}\right] = \left[\frac{(T_1 - T_2)m_2}{(m_1 + m_2)^2}\right], \tag{9.25}$$

$$g_{12} = {}^{\partial y_1}\!/_{\partial m_2}\Big| m_1 = \text{constant} = \frac{\partial}{\partial m_2}[m_1 + m_2] = 1, \tag{9.26}$$

$$g_{22} = {}^{\partial y_2}\!/_{\partial m_2}| m_1 = \text{constant} = \left[\frac{(T_2 - T_1)m_1}{(m_1 + m_2)^2}\right]. \tag{9.27}$$

Equations 9.24, 9.25, 9.26 and 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, 9.25, 9.26 and 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

$$\begin{bmatrix} y_1 \\ y_2 \end{bmatrix} = \begin{bmatrix} 1 & 1 \\ -.05 & 0.025 \end{bmatrix} \begin{bmatrix} m_1 \\ m_2 \end{bmatrix}, \tag{9.29}$$

which is of the form

$$\hat{y} = \hat{G}\,\hat{m}$$

as described in Equation 9.7.

RGA Calculation

To calculate the RGA, first the inverse of the steady-state matrix must be found:

$$G^{-1} = \begin{bmatrix} {}^1\!/_3 & -13{}^1\!/_3 \\ {}^2\!/_3 & 13{}^1\!/_3 \end{bmatrix}. \tag{9.30}$$

The transpose of G^{-1} is

$$\hat{R} = \left(G^{-1}\right)^{\mathrm{T}} = \begin{bmatrix} 1/3 & 2/3 \\ -13\,1/3 & 13\,1/3 \end{bmatrix}. \tag{9.31}$$

Now, the Hadamard product of the two matrices must be calculated where $\lambda_{ij} = g_{ij} \circ r_{ij}$. For example:

$$\lambda_{21} = (-0.05) \circ \left(-13\,1/3\right) = 2/3. \tag{9.32}$$

So, the resulting RGA is

$$\Lambda = \begin{bmatrix} 1/3 & 2/3 \\ 2/3 & 1/3 \end{bmatrix}. \tag{9.33}$$

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:

$$\frac{\partial y_1/\partial m_1}{\partial y_1/\partial m_1}\Bigg|_{\substack{m_2=\text{constant}\\ y_2=\text{constant}}} = \frac{1}{1 + \partial m_2/\partial m_1 \big|_{y_2=\text{constant}}}. \tag{9.34}$$

$\partial m_2/\partial m_1 \big|_{y_2=\text{constant}}$ will now be evaluated by substituting Equation 9.22 into Equation 9.23 and solving for m_1:

$$m_1 = \left(\frac{T_2 - y_2}{y_2 - T_1}\right) m_2. \tag{9.35}$$

Substituting Equation 9.35 into Equation 9.22:

$$y_1 = \left(\frac{T_2 - y_2}{y_2 - T_1}\right) m_2 + m_2. \tag{9.36}$$

Now this equation must be differentiated at constant y_2 and set equal to the denominator in Equation 9.34:

$$\partial y_1/\partial m_1 \Big|_{y_2=\text{constant}} = \left(\frac{T_2 - y_2}{y_2 - T_1}\right) \partial m_2/\partial m_1 + \partial m_2/\partial m_1 = 1 + \partial m_2/\partial m_1. \tag{9.37}$$

Evaluating to solve for $\partial m_2/\partial m_1$ at steady state:

$$\partial m_2/\partial m_1 = \left(\frac{y_2 - T_1}{T_2 - y_2}\right) = \left(\frac{308.15 - 278.15}{323.15 - 308.15}\right) = 2. \tag{9.38}$$

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 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 9.3, point 6 states that one should avoid pairing m_j with y_i when $\lambda_{ij} \leq 0.5$.

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

$$\Lambda = \begin{bmatrix} 2/3 & 1/3 \\ 1/3 & 2/3 \end{bmatrix}.$$

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:

$$\mathrm{NI} = \left.\frac{|\hat{G}|}{\prod_{i=1}^{n} g_{ii}}\right|_{ss} = \frac{(1)(-0.05) - (1)(0.025)}{(1)(-0.05)} = 1.5.$$

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 '$\hat{G}$' as follows:

$$\hat{G} = \begin{bmatrix} 1 & 1 \\ 0.025 & -0.05 \end{bmatrix},$$

where

$$b = g_{11}^2 + g_{12}^2 = 1^2 + 1^2 = 2,$$
$$c = g_{11}g_{21} + g_{12}g_{22} = (1) \times (0.025) + (1) \times (-0.05) = -0.025,$$
$$d = g_{21}^2 + g_{22}^2 = (0.025)^2 + (-0.05)^2 = 0.003125,$$

resulting in

$$\lambda_1 = s_1^2 = \frac{(b+d) + \sqrt{(b-d)^2 + 4c^2}}{2}$$
$$= \frac{(2 + 0.003125) + \sqrt{(2 - 0.003125)^2 + 4(-0.025)^2}}{2} = 2,$$
$$\lambda_2 = s_2^2 = \frac{bd - c^2}{s_1^2} = \frac{(2) \times (0.003125) - (-0.025)^2}{2} = 0.00281$$

and

$$\Sigma = \begin{bmatrix} s_1 & 0 \\ 0 & s_2 \end{bmatrix} = \begin{bmatrix} 1.414 & 0 \\ 0 & 0.053 \end{bmatrix},$$
$$\mathrm{CN} = s_1/s_2 = 1.414/0.053 \approx 27.$$

The condition number, CN, is less than 50 and, therefore, this system will not be prone to sensitivity problems [2].

9.6.2 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), the following variables can be used as manipulating 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 manipulating 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 manipulating 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 the three viable alternative pairings 4, 10 and 18, which are shown in boldface in Table 9.1. The reasons for discarding other pairings are as follows:

- Schemes 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.
- Schemes 6, 8, 14 and 19 are discarded since they involve manipulating the flow rate of bottom product or reboiler heat to control the liquid level in reflux drum.
- Schemes 21 and 22 are discarded since they do not regulate the material balance.
- Schemes 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 – 4, 10 and 18 – will be found through RGA analysis of a water–ethanol distillation column. A common approach is to use a process

Table 9.1 *Pairings in dual composition control.*

Case	Reflux Drum Level	Column Base Level	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

simulation software package to determine the necessary gains for the RGA analysis and the NI. 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.

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 (Table 9.5) for the column yield the following results.

Table 9.2 *Characteristics of the column feed.*

Conditions and Composition	
Temperature (°C)	20.0
Pressure (kPa)	101.3
Comp. molar flow of water (kmol/h)	60.00
Comp. molar flow of ethanol (kmol/h)	40.00

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)	30

RGA 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. 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. 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.

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)	4.2×10^6

Table 9.6 Specifications for case 4 open loop.

Specification	Value
Reboiler duty (kJ/h)	4.2×10^6
Distillate rate (kmol/h)	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)	4.2×10^6

The open loop gain is then calculated as follows:

$$g_{11} = \frac{\Delta x_{\mathrm{D}}}{\Delta D} = \left(\frac{0.7890 - 0.8165}{40 - 30}\right) = -2.75 \times 10^{-3}. \tag{9.39}$$

The closed loop gain, g_{11}^*, can be calculated from the steady-state solution by closing the bottoms composition control loop, that is, 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.

Now, the closed loop gain is calculated as follows:

$$g_{11}^* = \frac{\Delta x_{\mathrm{D}}}{\Delta D} = \left(\frac{0.6680 - 0.8165}{40 - 30}\right) = -1.48 \times 10^{-2}. \tag{9.40}$$

At this point, the RGA matrix can be calculated because this is a 2×2 system:

$$\lambda_{11} = \frac{g_{11}}{g_{11}^*} = \left(\frac{-2.75 \times 10^{-3}}{-1.48 \times 10^{-2}}\right) = 0.185. \tag{9.41}$$

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

$$\begin{bmatrix} \lambda_{11} & 1 - \lambda_{11} \\ 1 - \lambda_{11} & \lambda_{11} \end{bmatrix} = \begin{bmatrix} 0.949 & 0.051 \\ 0.051 & 0.949 \end{bmatrix}$$

Table 9.8 *Specifications for case 4 closed loop.*

Specification	Value
Mole fraction of ethanol in bottoms	0.2214
Distillate rate (kmol/h)	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)	2.6×10^6

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	$B = 70$	$L = 70$	0.8239	0.2180	$g_{11} = 7.4 \times 10^{-4}$	
	Closed loop	$B\ x_{EtOH} = 0.2214$	$L = 70$	0.8243	0.2214	$g_{11}^{*} = 7.8 \times 10^{-4}$	
18		$L = 60$	$L = 70$				2.8590
	Open loop	Reboiler duty $= 4.2 \times 10^6$		0.8388	0.2880	$g_{11} = 2.23 \times 10^{-3}$	
	Closed loop	$B\ x_{EtOH} = 0.2214$	$L = 70$	0.8243	0.2214	$g_{11}^{*} = 7.8 \times 10^{-4}$	

and for case 18 is

$$\begin{bmatrix} \lambda_{11} & 1-\lambda_{11} \\ 1-\lambda_{11} & \lambda_{11} \end{bmatrix} = \begin{bmatrix} 2.859 & -1.859 \\ -1.859 & 2.859 \end{bmatrix}.$$

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 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

$$\hat{G} = \begin{bmatrix} 7.4\times10^{-4} & -3.4\times10^{-4} \\ 1.7\times10^{-3} & 7.0\times10^{-3} \end{bmatrix}.$$

The NI can then be calculated from

$$\mathrm{NI} = \frac{|G|}{\prod_{i=1}^{n} g_{ii}} = \frac{\left|\left(7.4\times10^{-4}\right)\left(7.0\times10^{-3}\right) - \left(-3.4\times10^{-4}\right)\left(1.7\times10^{-3}\right)\right|}{\left(7.4\times10^{-4}\right)\left(7.0\times10^{-3}\right)} = 1.11.$$

Because the NI is >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:

$$\begin{aligned}
b &= g_{11}^2 + g_{12}^2 = (7.4\times10^{-4})^2 + (-3.4\times10^{-4})^2 = 6.63\times10^{-7}, \\
c &= g_{11}g_{21} + g_{12}g_{22} = (7.4\times10^{-4})\times(1.7\times10^{-3}) + (-3.4\times10^{-4})\times(7.0\times10^{-3}) \\
&= -1.12\times10^{-6}, \\
d &= g_{21}^2 + g_{22}^2 = (1.7\times10^{-3})^2 + (7.0\times10^{-3})^2 = 5.19\times10^{-5},
\end{aligned}$$

resulting in

$$\begin{aligned}
\lambda_1 &= s_1^2 = \frac{(b+d)+\sqrt{(b-d)^2+4c^2}}{2}, \\
s_1^2 &= \frac{(6.63\times10^{-7}+5.19\times10^{-5})+\sqrt{(6.63\times10^{-7}-5.19\times10^{-5})^2+4(-1.12\times10^{-6})^2}}{2}, \\
s_1^2 &= 5.19\times10^{-5}, \\
\lambda_2 &= s_2^2 = \frac{bd-c^2}{s_1^2} = \frac{(6.63\times10^{-7})\times(5.19\times10^{-5})-(-1.12\times10^{-6})^2}{(5.19\times10^{-5})} = 6.39\times10^{-7}
\end{aligned}$$

and

$$\Sigma = \begin{bmatrix} s_1 & 0 \\ 0 & s_2 \end{bmatrix} = \begin{bmatrix} 7.20 \times 10^{-3} & 0 \\ 0 & 7.99 \times 10^{-4} \end{bmatrix},$$

$$\mathrm{CN} = {}^{s_1}\!/_{s_2} = {}^{7.20 \times 10^{-3}}\!/_{7.99 \times 10^{-4}} \approx 9.$$

The condition number, CN, is less than 50 and, 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 if they are definitely unstable with the NI. SVD may be used to determine if the control loops are overly sensitive to errors in process gain, as well as if 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.

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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 and 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 which reduce or eliminate either the variations or the problems caused by variations.

To better illustrate 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 which varies noisily. Tank C is fed by streams which 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.

A Real-Time Approach to Process Control, Third Edition. William Y. Svrcek, Donald P. Mahoney and Brent R. Young.

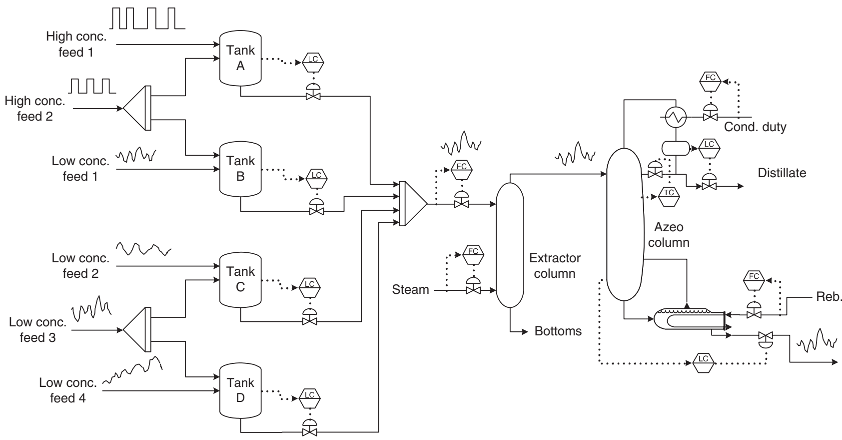

Figure 10.1 *Acid recovery plant control scheme (Reproduced from [3]. Copyright © 1991 by Greenwood Publishing Group, Inc., Wesport, CT. Reproduced with permission of ABC-CLIO, LLC).*

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.

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, while 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.

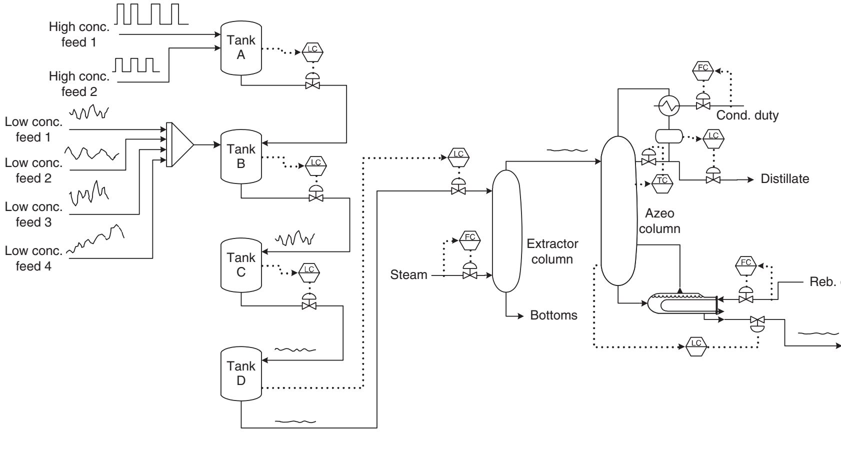

Figure 10.2 *Revised acid recovery plant control scheme (Reproduced from [3]. Copyright © 1991 by Greenwood Publishing Group, Inc., Wesport, CT. Reproduced with permission of ABC-CLIO, LLC).*

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, that is, 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 proportional integral 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, that is, cooling water, steam, refrigerant and so on. 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.

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, reflux and vapour boilup, so some combination of two variables can be controlled in each column, that is, two compositions, two temperatures or one temperature and one flow. Vapour boilup changes require changes in steam flow to the reboiler and also in cooling water flow indirectly through the pressure controller. Both vapour boilup 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,

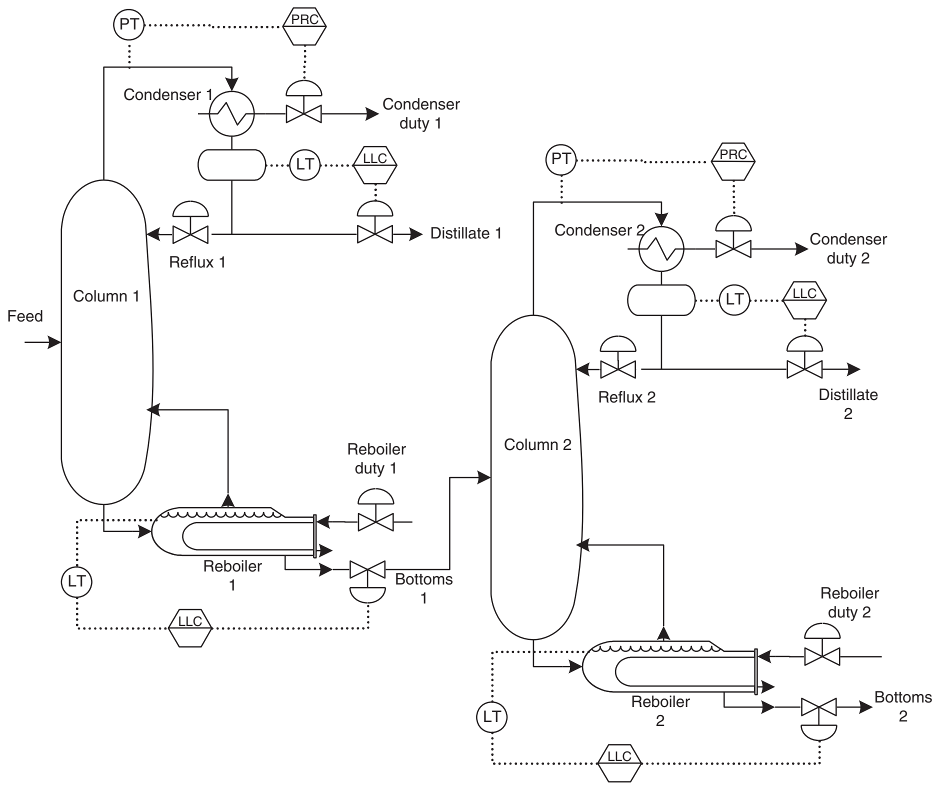

Figure 10.3 *Cascade system with two distillation columns in series.*

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, that is, products come off due to level control, or in the opposite direction, that is, feed is brought in on level control. The same design procedure applies.

10.3 Recycle Streams

If recycle streams exist in the plant, the procedure for designing an effective plant-wide control scheme becomes more complicated. Processes with recycle streams are quite common, but their dynamics are poorly understood at present.

The typical approach in the past for plants with recycle streams has been to install large surge tanks. This isolates the sequences of units and permits the use of conventional cascade process design procedures. However, this practice can be very expensive in terms of capital

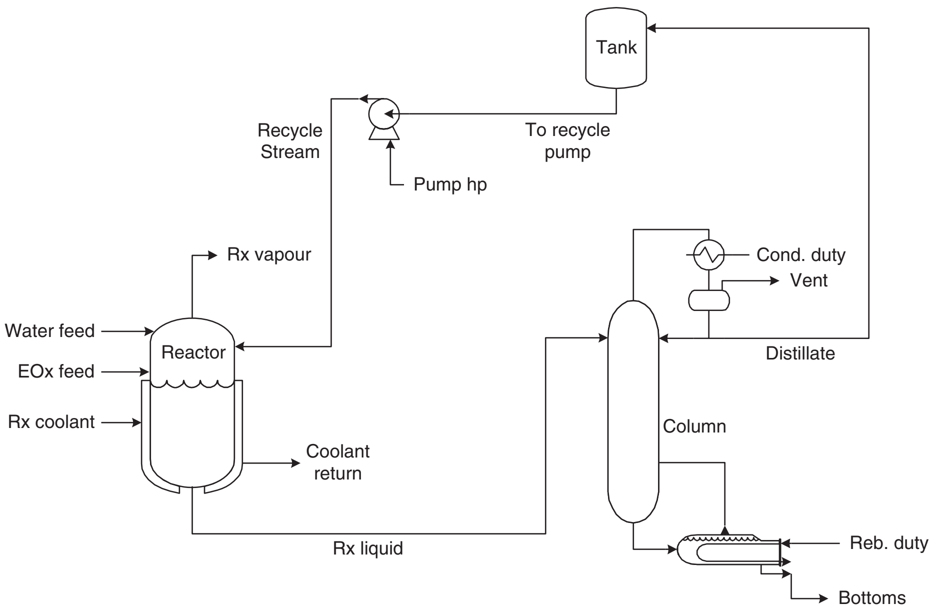

Figure 10.4 *Ethylene glycol plant.*

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 ethylene glycol. The unreacted feed is sent back through a recycle loop to the reactor.

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

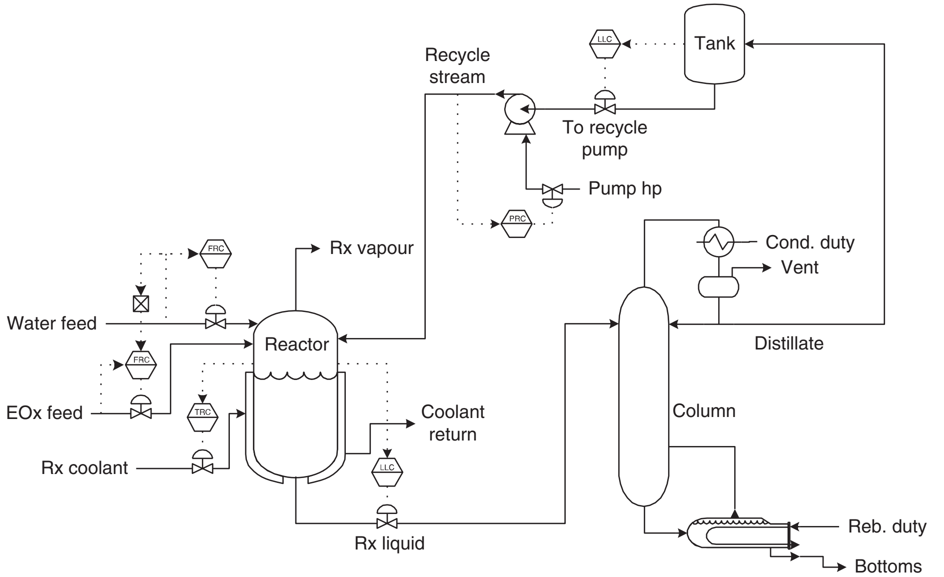

Figure 10.5 *Ethylene glycol plant control scheme 1.*

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 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 better illustrate this concept of snowballing, 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% and with a valve size on the stream to the recycle pump which results in almost saturated flow, that is, 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% point where the tank begins to overflow. The plant must shut down.

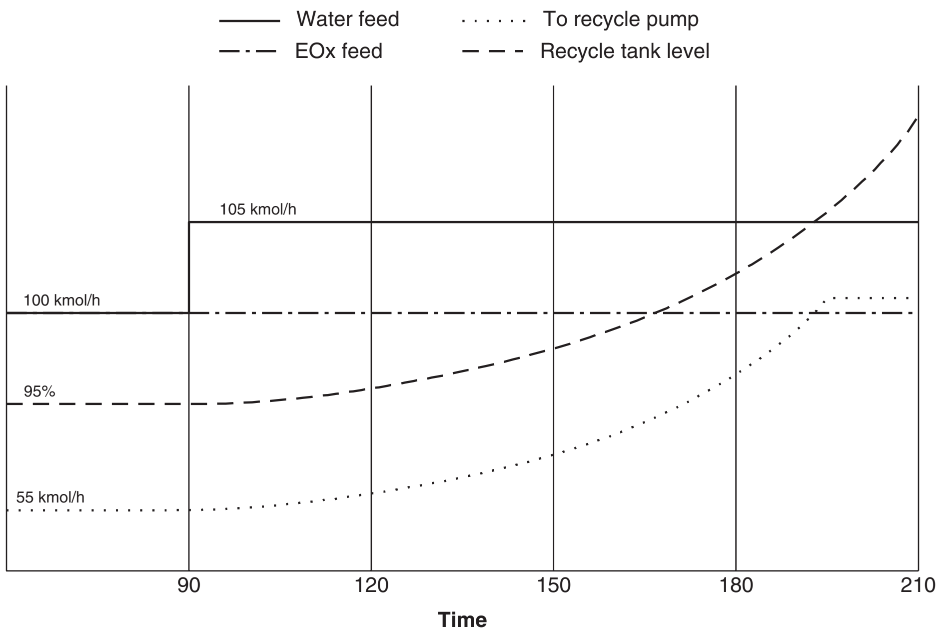

Figure 10.6 *Control scheme 1 response to a measurement error.*

If a positive measurement error is supplied, 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, 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.

While 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].

Luyben's Rules:

1. Only flow control a feed if it is sure to be fully consumed in the reaction.
2. Always put 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 level.

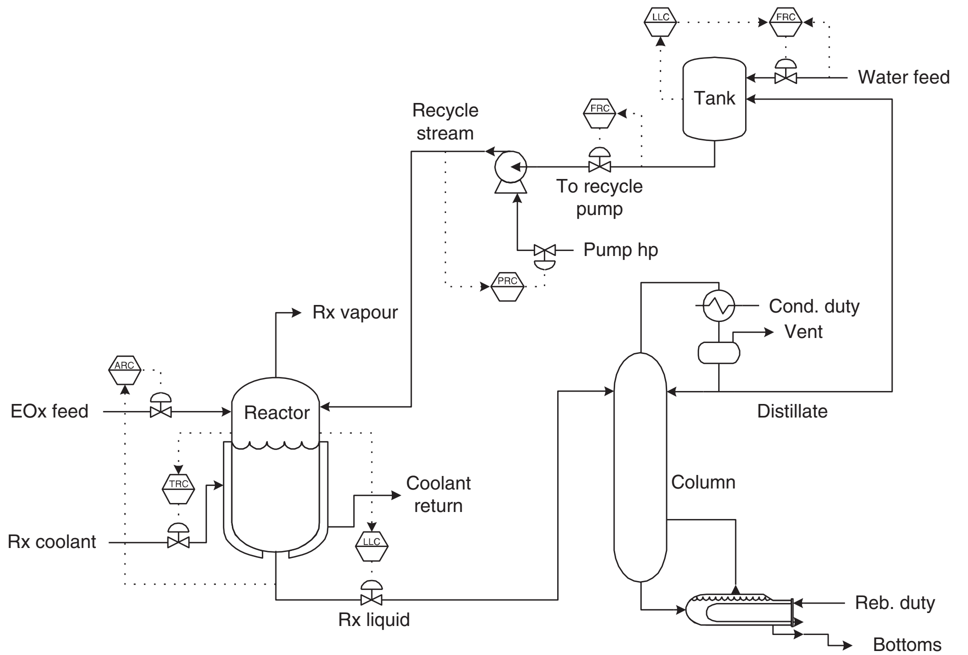

Figure 10.7 *Ethylene glycol plant control scheme 2.*

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 following strip chart (Figure 10.8) shows an introduction of −5 kmol/h error into the sensor transmitting a flow measurement to the water feed flow controller.

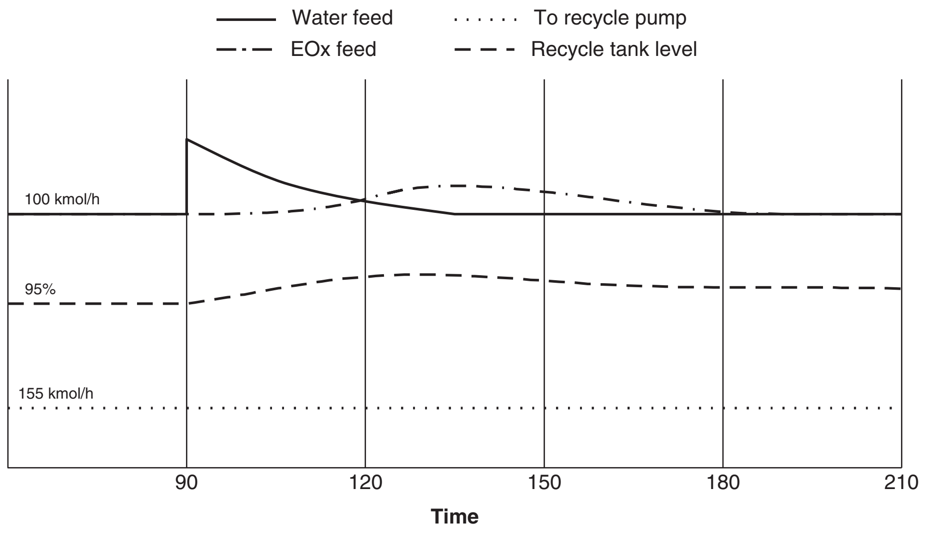

Figure 10.8 *Control scheme 2 response to measurement error.*

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 control for the level control. 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 continually developing area for research. As such, it cannot be summarized simply in one paragraph. For a more in-depth discussion of this topic, refer to the series of papers authored by W.L. Luyben and B.D. Tyreus [9, 11–14] and the book entitled *Plant Wide Process Control* by W.L. Luyben, B.D Tyreus and M.L. Luyben [10]. More recently Skogestad and co-workers [15, 16] have developed a systematic plant-wide control design procedure inspired by Luyben's approach which explicitly incorporates economics in a seven-step procedure that starts with a top-down approach to steady-state economics followed by a bottom-up approach dealing with stabilization and loop pairing. A practical article providing guidelines to ensure smooth plant operation is given by Lieberman [17], and a book on *Plantwide Process Control* is presented by Erickson and Hedrick [18].

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11

Advanced Process Control

In this final chapter we tackle the subject of advanced process control (APC) as is widely practiced in industry. Over the past 20 years or so, APC has become synonymous with model predictive control (MPC). Thus we will review why this is so by first considering what the term advanced process control actually has come to mean and why, and then introduce the reader to the fundamental concepts of MPC and industrial implementation of MPC.

11.1 Advanced Process Control

The reasons that APC may be necessary include the multivariable nature of processes, significant variations in process disturbances, variations in process dynamics and the ability to handle constraints which means that the performance of single-loop conventional control systems is not sufficient. APC can offer solutions to these issues and substantial economic benefits (savings of the order of 2–6% of annual operating costs typically (and reported up to 15%) and generate an extra 1% in revenue [1, 2]).

The term advanced process control means different things to different people. It is a relative term, to both the process and technology used. It is usually multivariable in nature, typically involves computer-based mathematical models of the process and can involve manipulation of the tuning constants (adaptive control) or model parameters (MPC) and on-line optimization; see, for example, [3].

Although admittedly subjective, Seborg [4] presented the following useful classification of process control strategies that fits with these considerations in Table 11.1. This classification still largely applies, despite being published almost 20 years ago.

As a consequence, thus far in this book we have concentrated on the first two categories of control strategies and have dealt with most of these topics extensively in the preceding chapters. In this chapter we will deal with advanced control as widely used in industry.

According to the classification, statistical quality control, internal model control and adaptive control are APC techniques that are widely applied, but we will not address them

A Real-Time Approach to Process Control, Third Edition. William Y. Svrcek, Donald P. Mahoney and Brent R. Young.

Table 11.1 Classification of process control strategies according to the degree of use in industry.

Category 1 – Conventional control
Manual, PID, ratio, cascade and feedforward control
Category 2 – Advanced control: classical techniques
Gain scheduling, time delay compensation, decoupling and selective/override control
Category 3 – Advanced control: widely used in industry
Model predictive control, statistical quality control, internal model control and adaptive control
Category 4 – Advanced control: with some industrial application
Optimal control, expert systems, artificial neuro-controllers and fuzzy logic control
Category 5 – Advanced control: proposed, with few (if any) applications

in any detail here beyond the following comments for the following reasons. Statistical process control [5, 6], while very useful, is not strictly a closed loop process control technique and is covered extensively elsewhere [5–7]. Internal model control (IMC) is not discussed further because it is really a general concept and a specific approach that has mainly been used industrially for model-based tuning (as described already in Chapter 5 [8]). Adaptive control has been widely applied but mainly as auto-tuners which provide PID controller tuning on a user-generated, on-demand basis using the ATV tuning technique (but this was also covered in Chapter 5 [9]).

The most widely used advanced control strategy is a general approach referred to as model predictive control, because it is a multivariable control strategy that addresses the process issues identified at the start of this section, there are literally thousands of applications worldwide and there are a number of vendors who market MPC and install it on a turnkey basis. We will therefore deal solely with this topic in this chapter.

11.2 Model Predictive Control

MPC is a class of multivariable computer control scheme that uses a process model for explicit prediction of future plant behaviour and involves computation of the appropriate control action required to drive the predicted output as close as possible to the desired target value. It is the most widely industrially used of all advanced control methods.

As mentioned previously, major industrial control challenges where advanced control should be considered include multivariable/interacting processes (where some variables may not be measurable), difficult dynamics (such as time delays, inverse responses, instabilities, non-linearities) and constraints (on absolute values and rates of change).

MPC has the capability for handling multivariable/interacting processes and difficult dynamics with ease by using a process model that can be simple. It controls measured and unmeasured variables (through inferential control). Posed as an optimization problem it optimizes control effort to meet objectives and is potentially capable of handling all constraints (hard, soft, equality and inequality constraints).

The general principles of MPC may be illustrated by considering first how the process variable will behave in the future if no further action is taken and target control action to rectifying what is left to be corrected after the full effects of the previously implemented control action.

The four basic elements of MPC are as follows:

1. A reference or set point trajectory specification, $y^*(k)$, where the argument (k) represents the value of reference trajectory time series, y^*, at the kth sampling or time instant.
2. Process variable prediction, $\hat{y}(k + i)$, where the argument $(k + i)$ represents the value of the predicted process variable, $\hat{y}$, at the $(k + i)$th sampling instant.
3. The model $\boldsymbol{M}$ predicts $\hat{y}(k + i)$, based on the manipulated variable or control action sequence, $u(k + j)$, where $u(k + j)$ is the value of the control action sequence at the $(k + i)$th time instant. The model $\boldsymbol{M}$ calculates the control action, to satisfy an optimization objective, subject to pre-specified constraints. This is akin to using the model inverse, $\boldsymbol{M}^{-1}$, which is mostly carried out numerically as the solution of an optimization problem.
4. The error prediction update

$$e(k) = y_{\mathrm{m}}(k) - \hat{y}(k), \tag{11.1}$$

where $e(k)$ is the error prediction update at the kth sampling instant and $y_{\mathrm{m}}(k)$ is the measured value of the process variable at the kth time instant.

Some of these basic elements of MPC are illustrated in Figure 11.1.

Standard MPC uses linear models and these MPC schemes are implemented on digital computers as linear discrete-time models [10, 11]. Three linear discrete-time models are mostly used:

1. Finite convolution models
2. Discrete state space models
3. Transfer function models

Finite convolution discrete-time models are expressed in either impulse response or step response forms. The impulse response model form shows the process variable sequence as a function of the control action sequence as a series of impulses, so

$$y(k) = \Sigma g(i)u(k - i), \quad \text{for } i = 0 \text{ to } k, \tag{11.2}$$

where the parameters, $g(i)$, comprise the impulse response function.

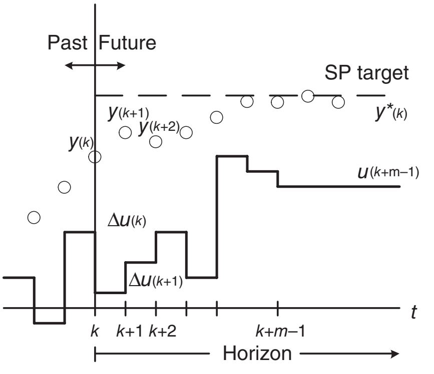

Figure 11.1 *Some of the basic elements of model predictive control.*

The step response model form expresses the process variable output sequence as a function of the series of control action changes or steps as in the following equation:

$$y(k) = \Sigma b(i)\Delta u(k-i), \quad \text{for } i = 0 \text{ to } k, \tag{11.3}$$

where $\Delta u(k) = u(k) - u(k-1)$ is the step or change in the control action between each sampling instant and the parameters, $b(i)$, constitute the step response function.

The impulse and step response functions are related as follows:

$$g(i) = b(i) - b(i-1), \tag{11.4}$$

$$b(i) = \Sigma g(j), \quad \text{for } j = 1 \text{ to } i. \tag{11.5}$$

For real, causal systems because of a mandatory one-step delay:

$$g(0) = b(0) = 0. \tag{11.6}$$

The remainder of the response coefficients, $g(i)$ or $b(i)$, are obtained from noise-free data or other models. The obtaining of these coefficients is via careful system identification of the plant and is a key aspect of MPC. As introduced elsewhere in the book (Chapter 3 and Workshop 3), experimental model building includes the construction, estimation of parameters and validation of mathematical models of dynamic systems based on observed data. The classical text on lumped parameter system identification is that by Ljung [12].

Discrete state space models, also known as auto-regressive with exogenous or external input models (more simply denoted as 'ARX' models), are

$$y(k) = \Sigma a(i)y(k-i) + \Sigma b(i)u(k-i-m), \quad \text{for } i = 0 \text{ to } k, \tag{11.7}$$

where $a(i)$ and $b(i)$ are the model parameters and m is the time delay.

Again, $a(0) = b(0) = 0$ for real, causal systems and the other parameters are obtained from fitting plant data via careful system identification experiments.

Transfer function models are also used by some MPC schemes, where the process variables and the manipulated variables are related by transfer functions, for example, first or second order plus dead time. Yet again, these transfer function models are obtained from system identification of the plant.

Many factors contribute to discrepancy between actual data and model predictions. These include un-modelled or unmeasured disturbances, fundamental errors in model structure and/or unavoidable errors in model parameter estimates. The fundamental MPC strategy is pragmatic. It is assumed that the discrepancy is caused by unmeasured disturbances and is constant. So the current discrepancy is applied to all predictions as an error prediction update.

The first MPC techniques were developed in the 1970s independently by two industrial groups. Dynamic Matrix Control (DMC) was developed by Shell in the United States [13], while a similar technique called Model Predictive Heuristic Control (MPHC) was developed in France [14]. Both these approaches were similar and share many common features. Since then these and many other related MPC algorithms have been developed and there have been thousands of instantiations across the world. Qin and Badgwell [10, 11, 15] and Morari and Lee [16] have presented excellent reviews of MPC technology. DMC and its descendants remain the most popular of the MPC technologies and will be described hereon in the next section.

11.3 Dynamic Matrix Control

In this section we will first introduce the concepts of DMC by considering the unconstrained case for a single process variable and then extend the treatment to include multiple variables and constraints. We then finish the chapter with a brief consideration about some of the practical implementation aspects of MPC generally.

DMC uses a step response model to represent the process:

$$y'(t) = y(t_i) - y(t_0) = a_i \Delta u(t_0). \tag{11.8}$$

We will illustrate this model with a first-order plus dead time example where the process gain $K_p = 1$, time constant $\tau = 1$ and dead time $L = 1$ (arbitrary units). A step change $\Delta u(t)$ is introduced at time $t = 0$, that is, $\Delta u(0) = 1$. For a sampling time of $h = 0.2$, the step response coefficients can then be easily calculated, the first 25 of which are shown in Table 11.2.

Table 11.2 *Example step response and step response model coefficients.*

Time	i	$\Delta u(t)$	$y'(t)$	a_i
0.0	0	1	0	0
0.2	1	0	0	0
0.4	2	0	0	0
0.6	3	0	0	0
0.8	4	0	0	0
1.0	5	0	0	0
1.2	6	0	0.18	0.18
1.4	7	0	0.33	0.33
1.6	8	0	0.45	0.45
1.8	9	0	0.55	0.55
2.0	10	0	0.63	0.63
2.2	11	0	0.70	0.70
2.4	12	0	0.75	0.75
2.6	13	0	0.80	0.80
2.8	14	0	0.83	0.83
3.0	15	0	0.86	0.86
3.2	16	0	0.89	0.89
3.4	17	0	0.91	0.91
3.6	18	0	0.93	0.93
3.8	19	0	0.94	0.94
4.0	20	0	0.95	0.95
4.2	21	0	0.96	0.96
4.4	22	0	0.97	0.97
4.6	23	0	0.97	0.97
4.8	24	0	0.98	0.98
5.0	25	0	0.98	0.98

We can use the model of Equation 11.8 to predict the process variable $y(t)$ given the control moves $\Delta u(t)$ as follows:

$$\begin{aligned}
y(t_1) - y(t_0) &= a_1 \Delta u(t_0), \\
y(t_2) - y(t_0) &= a_2 \Delta u(t_0) + a_1 \Delta u(t_1), \\
\vdots \qquad & \qquad \vdots \\
y(t_n) - y(t_0) &= a_n \Delta u(t_0) + a_{n-1} \Delta u(t_1) + a_{n-2} \Delta u(t_2) + \cdots,
\end{aligned} \tag{11.9}$$

that is, in vector notation:

$$y'(t) = y(t_i) - y(t_0) = \Sigma a_i \Delta u(t_{n-i}), \quad \text{for } i = 1 \text{ to } n, \tag{11.10}$$

or, in matrix form, we get the so-called dynamic matrix that DMC is named after:

$$\begin{bmatrix} y'(t_1) \\ y'(t_2) \\ y'(t_3) \\ \vdots \\ y'(t_n) \end{bmatrix} = \begin{bmatrix} a_1 & 0 & 0 & \cdots & 0 \\ a_2 & a_1 & 0 & \cdots & 0 \\ a_3 & a_2 & a_1 & \cdots & 0 \\ & \vdots & & & \vdots \\ a_m & a_m & a_m & \cdots & a_1 \end{bmatrix} \begin{bmatrix} \Delta u(t_0) \\ \Delta u(t_1) \\ \Delta u(t_2) \\ \vdots \\ \Delta u(t_{n-1}) \end{bmatrix}, \tag{11.11}$$

that is,

$$\boldsymbol{y}' = \boldsymbol{A} \Delta \boldsymbol{u}. \tag{11.12}$$

Note that we are assuming here that $n > m$, where n is the prediction horizon and m is the model horizon. This is often the case.

Using the dynamic matrix, we can predict the process variable y for a series of manipulated variable moves Δu. This is then implemented in DMC as a moving horizon algorithm. All past y and u values are known and future u values are chosen to regulate y using the step response model and past u values. This is then repeated after another time interval, that is, only first Δu is actually implemented.

The prediction vector $\boldsymbol{y}^{\mathrm{p}}$ ($y(t)$ for $t > t_0$) is the predicted value of the process variable assuming no future manipulated moves, that is, $\Delta u(t) = 0$, $t \geq t_0$, and it is obtained from the following calculation:

$$\begin{bmatrix} y^{\mathrm{p}}(t_1) \\ y^{\mathrm{p}}(t_2) \\ \vdots \\ y^{\mathrm{p}}(t_n) \end{bmatrix} = \begin{bmatrix} y(t_{-m}) \\ y(t_{-m}) \\ \vdots \\ y(t_{-m}) \end{bmatrix} + \begin{bmatrix} a_m & a_m & a_{m-1} & \cdots & a_3 & a_2 \\ a_m & a_m & a_{m-1} & \cdots & a_4 & a_3 \\ & & & \vdots & & \\ a_m & a_m & a_m & \cdots & a_m & a_m \end{bmatrix} \begin{bmatrix} \Delta u(t_{-m}) \\ \Delta u(t_{-m+1}) \\ \vdots \\ \Delta u(t_{-1}) \end{bmatrix}, \tag{11.13}$$

that is,

$$\boldsymbol{y}^{\mathrm{p}} = \boldsymbol{y}(t_{-m}) + \boldsymbol{A}^{\mathrm{p}} \Delta \boldsymbol{u}^{\mathrm{p}}, \tag{11.14}$$

and when combining with future control moves we get the following equation:

$$\boldsymbol{y} = \boldsymbol{y}^{\mathrm{p}} + \boldsymbol{A} \Delta \boldsymbol{u}. \tag{11.15}$$

The prediction of the process variable from Equation 11.15 will result in errors due to the following:

1. Errors in step response model coefficients
2. Unmeasured disturbances
3. Non-linear behaviour
4. Not being at steady state at $t = t_{-m}$

DMC uses a simple prediction error to account for these errors as follows:

$$e = y(t_0) - y^{\mathrm{P}}(t_0), \tag{11.16}$$

resulting in a revised prediction of the process variable, as follows:

$$\boldsymbol{y} = \boldsymbol{y}^{\mathrm{p}} + \boldsymbol{A}\Delta\boldsymbol{u} + \boldsymbol{e}^{\mathrm{T}}, \tag{11.17}$$

where $\boldsymbol{e} = [e\ e \cdots e]$.

DMC is based on minimizing the error from the set point trajectory. The objective function is the sum of the square of the errors from the set point trajectory n steps into the future:

$$\Phi = \Sigma[y_{\mathrm{sp}} - y^{\mathrm{P}}(t_i)]^2, \quad \text{for } i = 1 \text{ to } n, \tag{11.18}$$

where y_{sp} is the set point trajectory and n is the prediction horizon. If we then define

$$E(t_i) = y_{\mathrm{sp}} - y^{\mathrm{P}}(t_i) - e \tag{11.19}$$

and substitute Equation 11.19 into Equation 11.18 we arrive at the following revised form of the objective function:

$$\Phi = \Sigma[E(t_i) - y_{\mathrm{c}}(t_i)]^2, \quad \text{for } i = 1 \text{ to } n, \tag{11.20}$$

where

$$\boldsymbol{y}_{\mathrm{c}} = \boldsymbol{A}\Delta\boldsymbol{u} \tag{11.21}$$

The objective of the DMC controller is to choose $\Delta u(t_i)$, for n moves into the future to minimize Φ, the sum of the squares of the errors from set point. The control law is obtained by differentiating Φ with respect to Δu:

$$\mathrm{d}\Phi/\mathrm{d}\Delta\boldsymbol{u} = \boldsymbol{A}^{\mathrm{T}}(\boldsymbol{A}\Delta\boldsymbol{u} - \Phi) = 0/ \tag{11.22}$$

and solving for $\Delta\boldsymbol{u}$ we obtain

$$\Delta\boldsymbol{u} = (\boldsymbol{A}^{\mathrm{T}}\boldsymbol{A})^{-1}\boldsymbol{A}^{\mathrm{T}}\boldsymbol{E}. \tag{11.23}$$

Note that if $\boldsymbol{A}$ is square, $(\boldsymbol{A}^{\mathrm{T}}\boldsymbol{A})^{-1}\boldsymbol{A}^{\mathrm{T}} = \boldsymbol{A}^{-1}$.

However, we are not quite yet done with the control law. It turns out that Equation 11.23 results in an overly aggressive response and $(\boldsymbol{A}^{\mathrm{T}}\boldsymbol{A})^{-1}$ can well be mathematically ill-conditioned due to process model mismatch or incorrect identification of process dead time. As a result a diagonal move suppression matrix is employed to overcome these issues:

$$\boldsymbol{Q} = q\boldsymbol{I}, \tag{11.24}$$

where the move suppression factor q is positive. A large value penalizes Φ more for control moves and is also used for large non-linearities/disturbances. The resulting control law is

$$\Delta \boldsymbol{u} = (\boldsymbol{A}^{\mathrm{T}}\boldsymbol{A} + \boldsymbol{Q}^2)^{-1}\boldsymbol{A}^{\mathrm{T}}\boldsymbol{E}. \tag{11.25}$$

From the above discussion the reader will have noted that the DMC controller has a large number of tuning parameters – the model horizon (m), the prediction horizon (n), the number of control actions, the model parameters (a_i) and the move suppression matrix ($\boldsymbol{Q}$). The prediction horizon (n) must be chosen so results of control actions can be seen in it. The number of control actions calculated is dictated by process dead time and prediction horizon.

Once these tuning parameters are defined, DMC is implemented as follows. The controller gain matrix $\boldsymbol{K}_{\mathrm{c}}$ is first calculated off-line via

$$\boldsymbol{K}_c = (\boldsymbol{A}^{\mathrm{T}}\boldsymbol{A} + \boldsymbol{Q}^2)^{-1}\boldsymbol{A}^{\mathrm{T}}. \tag{11.26}$$

Then the uncontrolled output predictions are calculated on-line via

$$\boldsymbol{y}^{\mathrm{p}} = \boldsymbol{y}(t_{-m}) + \boldsymbol{A}^{\mathrm{p}}\Delta \boldsymbol{u}^{\mathrm{p}} \tag{11.27}$$

And then finally the next control action (only) from the control law is calculated on-line via

$$\Delta \boldsymbol{u} = \Sigma \boldsymbol{k}_{ii}[\boldsymbol{y}_{\mathrm{sp}}(t_i) - \boldsymbol{y}^{\mathrm{P}}(t_i)], \tag{11.28}$$

where the $\boldsymbol{k}_{ii}$ are the elements of the first row of $\boldsymbol{K}_{\mathrm{c}}$.

Thus far we have developed the case for DMC of a single process variable without constraints. However, as mentioned by several authors, the value of MPC (and therefore DMC) is its ability to handle multiple interacting variables and multiple constraints on variables. Therefore we will consider extension to multiple variables and constraints next.

Extension of DMC to multiple input, multiple output (MIMO) processes is handled easily by using augmented vectors and matrices. For example, the augmented dynamic matrix and control moves vector are as follows:

$$\boldsymbol{A} = \begin{bmatrix} \boldsymbol{A}_{11} & \boldsymbol{A}_{12} & \cdots & \boldsymbol{A}_{1j} \\ \boldsymbol{A}_{21} & \boldsymbol{A}_{22} & \cdots & \boldsymbol{A}_{2j} \\ \vdots & & & \\ \boldsymbol{A}_{k1} & \boldsymbol{A}_{k1} & \cdots & \boldsymbol{A}_{kj} \end{bmatrix} \quad \text{and} \quad \Delta \boldsymbol{u} = \begin{bmatrix} \Delta u_1 \\ \Delta u_2 \\ \vdots \\ \Delta u_j \end{bmatrix}, \tag{11.29}$$

where j is the number of manipulated variables and k is the number of process variables.

As a consequence, we also need to prioritize the control objectives in the control law. This is done as follows by adding an augmented diagonal weighting matrix, $\boldsymbol{W}$, that contains diagonal matrices which provide the relative weighting for the controlled variables:

$$\Delta \boldsymbol{u} = (\boldsymbol{A}^{\mathrm{T}}\boldsymbol{W}^2\boldsymbol{A} + \boldsymbol{Q}^2)^{-1}\boldsymbol{A}^{\mathrm{T}}\boldsymbol{W}^2\boldsymbol{E}. \tag{11.30}$$

The augmented diagonal weighting matrix $\boldsymbol{W} = w_i \cdot \boldsymbol{I}$ and therefore becomes an additional set of tuning parameters for the multivariable DMC controller (i.e. the relative weighting factors for the controlled variables).

Thus far we have assumed that the control moves $\Delta \boldsymbol{u}$ are unbounded. However, this is rarely the case in real process control. The first approach to solving a constrained DMC case was a modified unconstrained approach to DMC which is as follows:

1. Solve the unconstrained DMC problem.
2. Check for constraint violations.
3. Fix any infeasible variables from the above solution at their limits, remove them from the DMC problem and re-solve.
4. Repeat steps 2 and 3 until a feasible solution is found.

The problems with this modified unconstrained DMC approach are that whenever constraints are violated, the dynamic matrix is modified and the entire DMC problem must be solved on-line, it is an iterative approach and a solution is not guaranteed.

The solution to these problems was the development of the quadratic DMC approach (QDMC) [10, 11]. This is an optimization-based approach with the following objective function:

$$\min \Phi = [y_{\text{sp}} - y^{\text{P}}(t_i)]W^2[y_{\text{sp}} - y^{\text{P}}(t_i)] + \Delta u^{\text{T}} Q^2 \Delta u, \tag{11.31}$$

subject to the following:

$$\begin{array}{ll}
y = y^{\text{P}} + A\delta u & \text{(model)},\\
g(y,\ \Delta u,\ u) \geq 0 & \text{(constraints)},\\
u_{\text{L}} \leq u \leq u_{\text{H}} & \text{(control move bounds)},\\
\Delta u_{\text{L}} \leq \Delta u \leq \Delta u_{\text{H}} & \text{(control move rate bounds)},\\
y_{\text{L}} \leq y \leq y_{\text{H}} & \text{(process variable bounds)}.
\end{array}$$

The major advantages of QDMC are that any constraint can be explicitly represented in the control calculations and it determines the optimal, feasible moves in the prediction horizon. However, there is no analytical solution – the optimization must be solved at each control interval. This is done by a quadratic program, hence the name quadratic DMC.

Thus far we have largely, by implication, considered the number of inputs and outputs in the DMC formulation to be the same, that is, the case of square DMC. However, the case when the number of inputs and outputs are not equal occurs frequently in real processes. This is termed non-square DMC. In the case of non-square DMC, steady-state 'economic' optimization is used to specify good compromises among variables. For excess inputs, the optimization is often specified as a linear program. The solution to this problem provides steady-state targets to which input variables should be driven. Steady-state targets are then introduced into QDMC. The problem is then solved on-line and is computationally heavy.

11.4 General Considerations for Model Predictive Control Implementation

When considering MPC, a number of questions must be answered:

- What is the basic plant instrumentation like? Are the instruments well maintained and the control valves working effectively, for example, non-stiction?

- What is the fundamental regulatory control system working like? Have we got the optimal plant-wide control strategy (Chapter 10)? Are all the controllers in auto and well-tuned?
- What are the costs and potential benefits of installing an MPC system?
- What are the control objectives?
- Can the control objectives be defined in terms of the available process variables? Do we need to build 'soft' sensors to infer process variables?
- Are there constraints on the process variables which are likely to be violated during normal controller operation?
- If the problem is square, is there some natural, independent pairing between input and output variables?
- If the problem is non-square, is the additional economic optimization justified and is accurate economic information available?
- Are personnel sufficiently capable to be trained to maintain the system once it is implemented?

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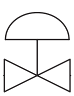
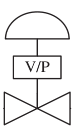
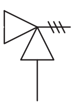
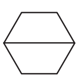
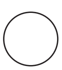
Appendix A
P&ID Symbols

Symbol	Description
	Control valve
	Valve
	Control valve with valve positioner
	Check valve
	Pressure relief valve
	Controller
	Transmitter/sensor
	Transmitter/sensor
	Controller summer

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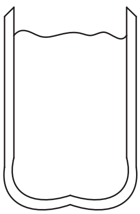
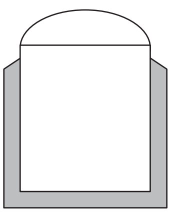
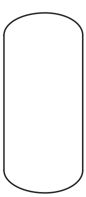
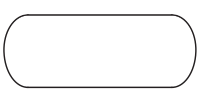
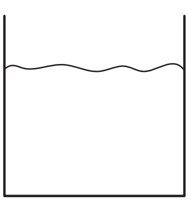
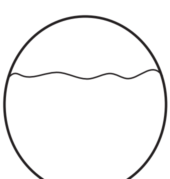
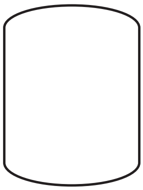
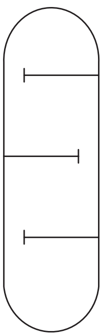
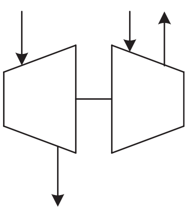
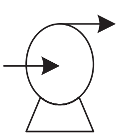
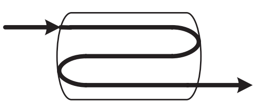
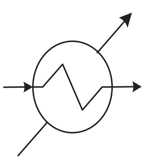
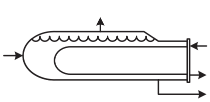
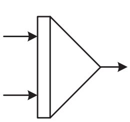
Symbol	Description
	Summer or multiplier
	Divider
	Selector
	Reactor with cooling jacket
	Reactor with cooling jacket
	Knock-out drum
	Reflux drum
	Tank
	Horizontal tank

Symbol	Description
	Vertical tank
	Distillation column tray section
	Compressor
	Pump
	Heat exchanger
	Heat exchanger
	Kettle reboiler
	Stream mixer

Appendix B
Glossary of Terms

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 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.
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.

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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* (ζ) 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 non-linear 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.

Non-linear: 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*.

Oscillation: See cycling.

Output: The variable that is chosen to describe the condition of a system; the dependent variable in the dynamic equation.

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% 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.

Reference

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Appendix C

New Capabilities with Control Technology Hardware and Software

C.1 Foundation Fieldbus for Instrumentation and Control

In the mid-1990s, significant advances in microprocessor technologies made possible the development of field-based electronic devices that led to a revolution in field instrument communications. The move to an all-digital communication protocol was pushed by the need to access the wealth of diagnostic data and advanced functions now possible in these enhanced field devices. Several technologies came together in what is now called Foundation Fieldbus (FF).

The primary benefit of a completely digital communication bus is the ability to access multiple variables available from more powerful devices and to use these for control applications, reducing the number of devices needed to implement various control strategies. The FF supports a deterministic communication that allows devices to communicate directly to each other in order to perform closed loop control, independently of the control system to which it is connected. With multiple devices connected to a common bus, or segment, the installed wiring costs of the instrumentation are reduced, along with cabinet footprint. Each segment can host a mix of input and output devices, removing the need for rationalizing I/O signals to dedicated card types, thus reducing the design costs of the physical I/O.

Since its introduction, FF has seen a steady increase in adoption as users realized the benefits of this technology over conventional field instruments and the associated reduction in installation costs.

C.2 Hardware Specifications

FF is based on a hardware specification that defines the physical parameters for the wiring and the number of devices that can be installed together on a segment. The bus can

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Table C.1 *FF specifications (Compiled from information in AG-140 Revision 1.0 (www.fieldbus.org)).*

Baud rate	31.25 kbits/s
Signal type	Manchester encoding (alternating polarity at 800 mV peak to peak)
Wire media	Twisted pair with shield
Max. device per segment	32
Minimum voltage	9 V (measured at the device)
Maximum voltage	30 V (typical)
Maximum trunk distance	1900 m
Maximum spur length	120 m
Terminators	2 per segment

supply transmitter power as well as digital communication. FF was designed to operate on the same type of field wiring recommended for traditional 4–20 mA instrument wiring (Table C.1).

The physical wiring installation rules for FF are very flexible to accommodate both new installations and retrofitting of brown field installations. Figure C.1 shows a typical layout of a segment trunk with a number of spurs.

Figure C.1 shows how multiple devices can be connected together on a segment. The trunk cable is typically the longer one and replaces a traditional multi-core cable that would be required for conventional two-wire devices. As a standard technology, many manufacturers provide devices and supporting hardware such as couplers, power conditioners and terminators. The Foundation also defines a validation protocol for all manufacturers to have their devices tested for compliance to the standard, ensuring inter-operability of devices. Although all devices must meet this inter-operability testing, not all devices are equal. Each device manufacturer can add to the minimum requirements to make their devices more suitable for certain markets or applications. From a customer perspective, it is important to

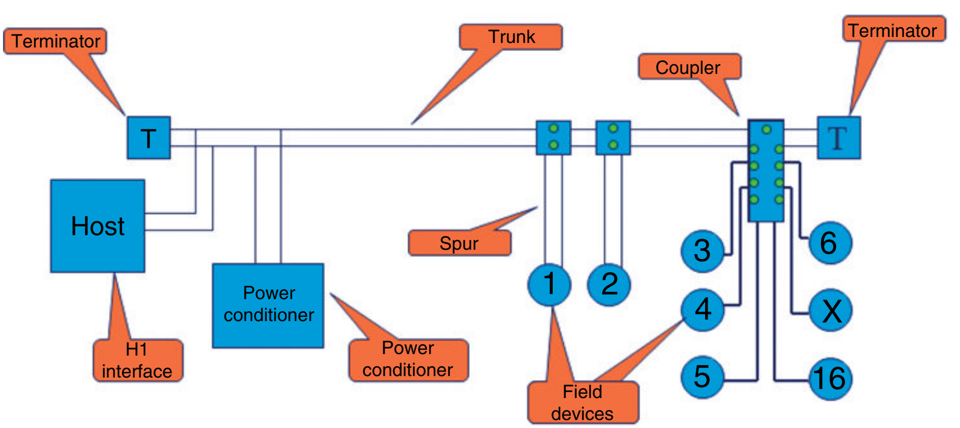

Figure C.1 *Segment diagram (Source: Spartan Controls PowerPoint presentation, author unknown).*

Table C.2 *Wiring type and length.*

Type	Description	Size	Max. Length
A	Shielded, twisted pair	#18 AWG (0.8 mm^2)	1900 m (6232 ft)
B	Multi-twisted-pair with shield	#22 AWG (0.32 mm^2)	1200 m (3963 ft)
C	Multi-twisted-pair without shield	#26 AWG (0.13 mm^2)	400 m (1312 ft)
D	Multi-core, w/o twisted pairs and having overall shield	#16 AWG (1.25 mm^2)	200 m (656 ft)

select the FF devices to meet the process requirements, keeping in mind that with FF, the devices are more than simply I/O – they form an integral part of the control system.

Table C.2 provides a breakdown of wiring types and maximum length.

Like any digital communication bus, the signal is impacted by the electrical characteristics of the wire and devices. Type A wire provides the best environment and therefore the least amount of signal attenuation per linear foot. The wire distance is reduced significantly with type D wire because of the impact on the signal due to the capacitance of the wire. For new installations, type A cable should be used to avoid physical layer issues and restrictions in segment topology.

In addition to selecting the correct wire type, the segment is also conditioned using a pair of terminators installed at the ends of the trunk wire. The terminators absorb the energy of the signals and prevent reflections from traversing back down the wire. Spurs are shorter distances and are not terminated. Having too many terminators can cause excessive cycling (Figure C.2).

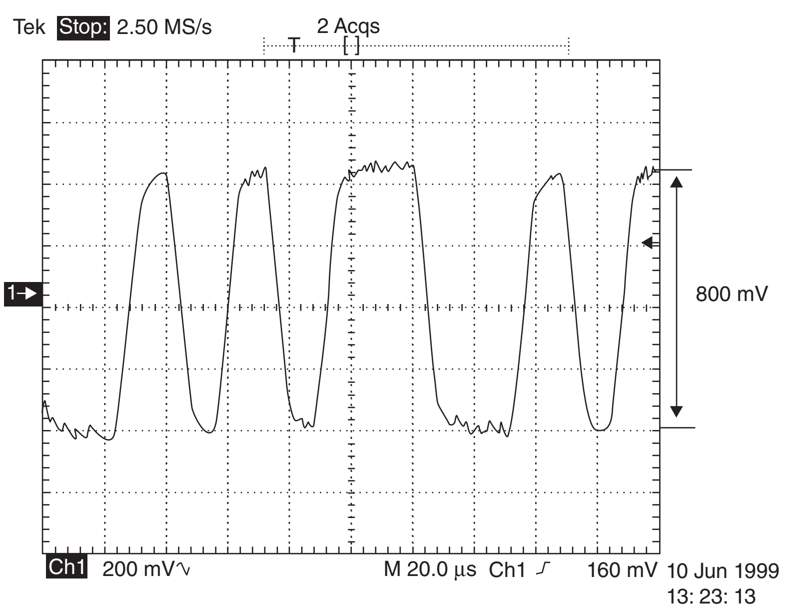

Figure C.2 *Typical waveform of Manchester encoding.*

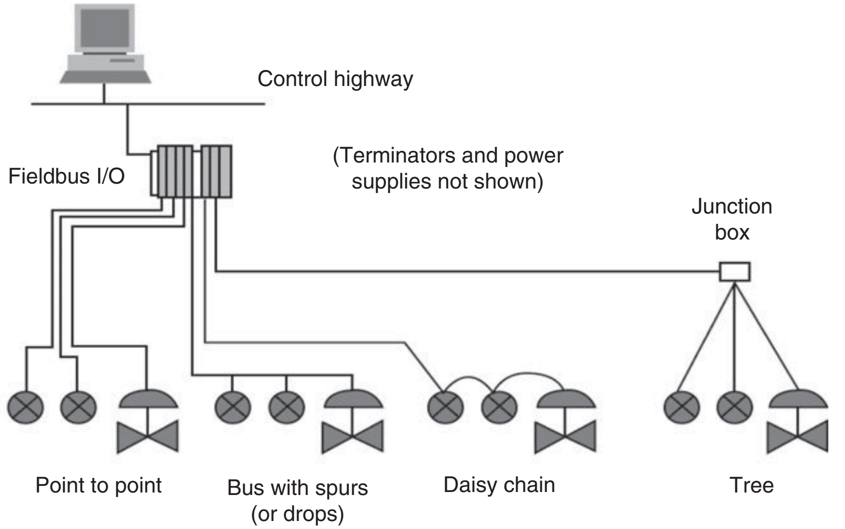

Figure C.3 *Segment wiring architectures.*

The design of FF segments can vary greatly, from point to point, daisy chain, bus with spurs, tree or combinations of these (Figure C.3).

The overall length of the segment can be calculated to avoid problematic designs due to wire length. Standard terminal blocks can be used in junction boxes to connect spurs to the trunk. There are also segment hardware components from various manufacturers that are designed specifically for FF H1 segments that facilitate the installation while providing additional protection from wiring errors on the spurs.

C.3 DeltaV Implementation

The DeltaV system was one of the first major control systems to adopt FF. The H1 interface card was developed and supports 2 H1 segments, with up to 16 devices per segment, for a total of 32 devices per card. Although the FF specification supports a maximum of 32 devices per segment, the DeltaV implementation uses some of these addresses for device commissioning and to support diagnostic devices. There is a limitation on the number of segment-powered devices that effectively limits the number of devices as well. By supporting 16 devices, many design issues are eliminated, simplifying the engineering effort for each segment.

FF was designed for control in the field. This means the entire control loop can be implemented using field devices. A simple PID loop requires an analog input (AI) block and an analog output (AO) block in order to connect the PID block to the process. (See Figure C.4.)

The function block editor allows the user to graphically link the AI to the PID input and the PID output to the AO block. A third connection connects the AO to the PID on what is

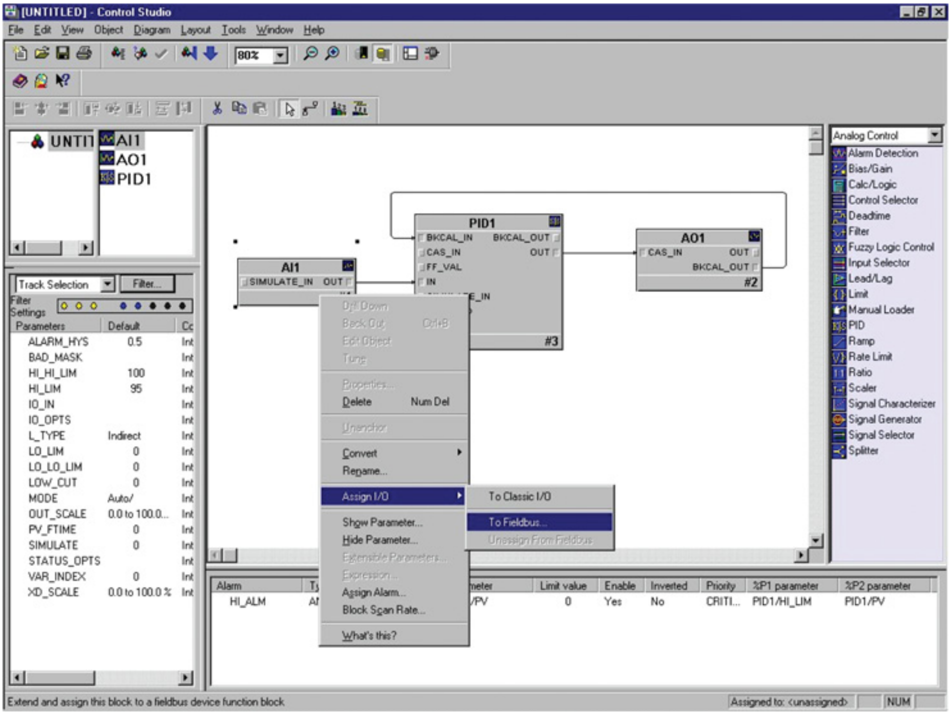

Figure C.4 *Typical PID loop.*

called the 'back calculation' signal. This signal synchronizes the PID block to the current AO output when the AO block is being operated directly by the operator.

In addition to a value, each signal in FF is also accompanied by a status. The status can be BAD, Uncertain, Good or Good Cascade. With traditional instrumentation, a transmitter fault may cause the 4–20 mA signal to be driven to a minimum or maximum value or may hold steady, no longer representing the process measurement. The control algorithm may respond to this 'faulty' signal, assuming it represents the process. Without the status of the transmitter, the process may be disrupted as the control algorithm drives its output to saturation. With FF, a fault in the transmitter will immediately cause the status of the signal to reflect the usability of the value. This status is used by the PID block to suspend automatic control and hold the process steady. Enhanced diagnostics in FF devices help identify many conditions that could impact the sensor reading and use the signal status to tell the control system about the usability of the signal.

C.4 Segment Design

FF systems are structured around the physical segment that directly connects a group of instruments and allows peer-to-peer communication used for closed loop control in the field. There are two limiting factors that must be considered in designing the segment. On the physical layer, the power conditioner must be able to handle the number of devices installed, based on the physical wiring required by the process piping. The second is the segment execution period, or the macrocycle must be fast enough to support the control algorithm based on the process dynamics.

Most systems today support a maximum of 16 devices per FF segment. During design, a maximum of 8–12 devices may be specified, allowing for expansion during and after the project is completed. The required power conditioner must be able to provide sufficient power for all the devices.

The other consideration is the process dynamics, which determine the frequency of execution for PID control blocks. Typically, PID blocks can be run at once a second, easily allowing 12–16 devices on a segment with the associated function blocks. Faster loops may require as fast as 250 millisecond execution, at which time, the segment may be limited to fewer devices.

DeltaV supports a feature called module-driven macrocycle, which allows a faster loop to be scheduled multiple times within the overall slower macrocycle. This allows more devices to be supported on a segment that runs both slow and fast loops. The vast majority of control loops will require no more than 1 second execution rates.

C.5 Control Strategy Design

FF devices now provide a wide variety of function blocks that allow great flexibility in the design of control algorithms executing in the field. The basic PID control block is supported in many transmitters and valve positioners. In most cases, the PID block should be assigned to the valve positioner where the AO block resides. This reduces the segment communications as the PID output and the AO back calculation signals are performed internally to the device. The AI will be executed in the sensing device. In some cases, signal characterization or math calculations are needed to condition the signals for use by the PID. These are often available within the transmitter. Instrumentation blocks such as control selectors, input selectors, splitters and signal characterizers are common in many devices. Math blocks and special function blocks are supported by major brands of transmitters and positioners.

The control strategy is developed using a configuration tool like DeltaV's Control Studio (Figure C.5). A graphical function block design tool allows the various blocks to be assembled to satisfy the control needs and each block is assigned to the appropriate device

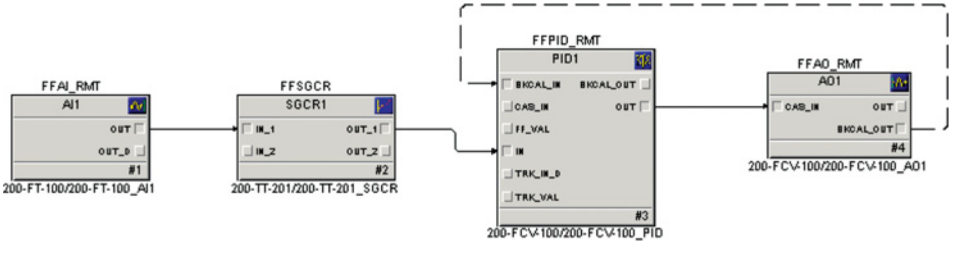

Figure C.5 *Typical closed loop control in the field (Screen image captured from Emerson Process Management's DeltaV Control Studio, v11.3.1).*

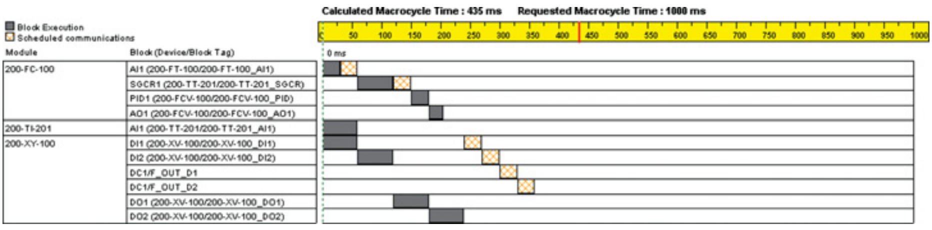

Figure C.6 *Example macrocycle schedule (Screen image captured from Emerson Process Management's DeltaV Control Studio, v11.3.1).*

for execution. The order of execution is defined by this diagram. By assigning all the function blocks to fieldbus devices, the order of execution is synchronous from the AI to the AO block, ensuring the algorithm is executed in its entirety every macrocycle.

C.6 Macrocycle

The FF segment is based on a synchronous execution of function blocks and intercommunications. The schedule of execution (Figure C.6) is managed by the Link Active Scheduler (LAS) and is typically the host system H1 interface card. The schedule ensures that function blocks in a given sequence are executed in order, with the inter-device communication scheduled between these blocks. In Figure C.6, function blocks are represented by dark grey boxes with hatched boxes representing the scheduled 'Compel Data' needed in the next block. The hatched boxes represent the connection lines in the module diagram above that connect blocks in different devices. When the blocks are in the same device, they are directly connected internally and do not require segment traffic.

Function blocks in different devices can execute in parallel, or at the same time if they are not interconnected in a control module execution order. In this example, the frequency of execution has been determined to be 1 second, though the actual execution time of the segment blocks and communications is about 300 ms. The remaining time is available for unscheduled communications, such as diagnostic requests or for genera function block data. All these additional communications are managed automatically and require no communication. In many applications, the segment macrocycle is set in the H1 interface properties.

In Figure C.7, two different control modules use devices on the same segment. One module requires an execution frequency of 1 second based on the process dynamics. The second module only requires a 2-second execution rate. The option to use the module execution frequency to set the macrocycle allows both modules to execute at their optimum frequency, ensuring the best performance of the system and allowing more devices to share the same segment.

C.7 Asset Management

As mentioned earlier, FF provides enhanced diagnostics that improve the quality of the control signal and performance of the control loop. Another benefit of these intelligent devices

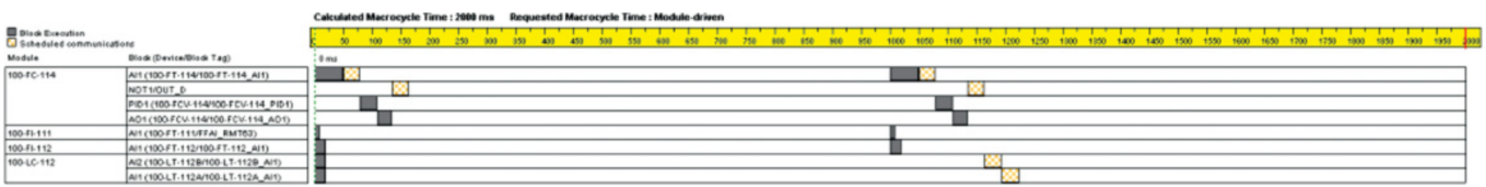

Figure C.7 *Module-driven macrocycle with two different module scan times.*

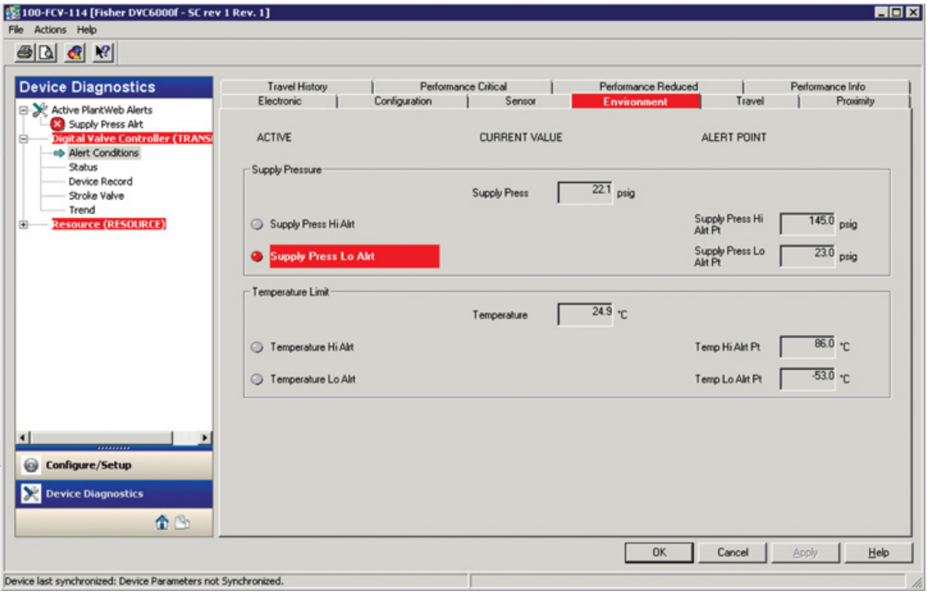

Figure C.8 *Device diagnostic screen showing loss of supply pressure to valve.*

is in the level of diagnostic information available for predictive maintenance (Figure C.8). Each device manufacturer is able to add key diagnostics to their devices to facilitate identification of faults before they actually affect the control level signals. Some devices perform statistical calculations on the process signal to derive blocked sensor lines, or they measure multiple variables on a valve actuator to determine wear, stiction and hysteresis.

All this information is available to Asset Management software that helps record abnormal events in the fieldbus devices and helps manage maintenance activities, including calibration schedules, and repair activities, including configuration changes. By moving from schedule or preventive maintenance to predictive maintenance, intelligent FF devices (Figure C.9) can reduce maintenance costs by reducing unnecessary maintenance and avoiding costly outages from unexpected failures.

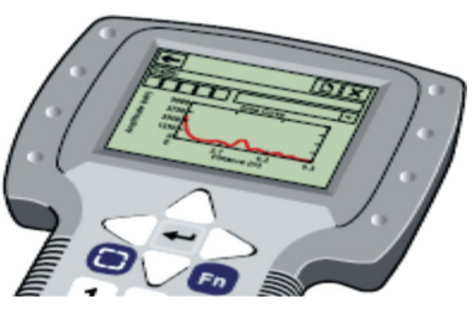

Figure C.9 *Handheld device diagnostic tool (Screen image captured from Emerson Process Management's AMS Intelligent Device Manager, v11).*

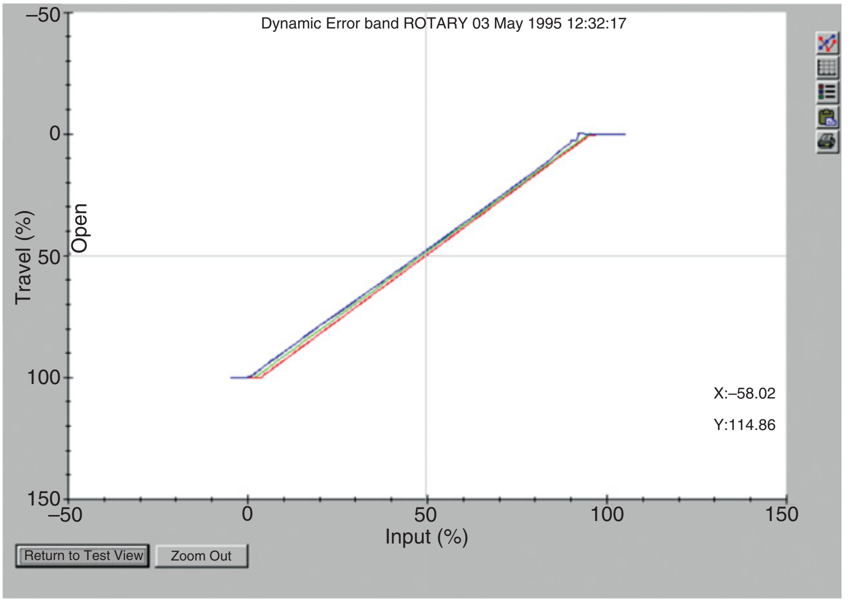

Figure C.10 Valve signature from smart FF positioner.

C.7.1 Valve Signatures

In situ valve signatures (Figure C.10) can confirm installed performance at start-up and comparable performance at turn-around to determine if the valve needs maintenance. Only valves that need to be removed from the line are replaced. Knowledge of the nature of a repair can ensure needed parts are in stock.

C.7.2 Statistical Data and Fault Detection

Another advantage of the intelligent capabilities of FF devices is made possible by their powerful microprocessors. Because the bus is entirely digital, power to the devices does not vary with the control signal. As a result, the processor can perform other tasks with this extra capacity.

While the transmitter may send the process signal to the control host once a second, the transmitter can monitor the process at a much higher rate to determine specific characteristics of the measurement. Statistical analysis of these data is done in the transmitter to determine standard deviations and even noise frequencies. This information can then be used in diagnostic ways to determine whether the sensing lines of a DP cell are plugged or whether there is cavitation occurring in a pump upstream of the sensor (Figure C.11).

C.8 Wireless Technology for Instrumentation and Control

From the I/O card in the panel, a wireless gateway (Figure C.12) can be used to communicate with up to 100 field devices. An example is shown in Figure C.12.

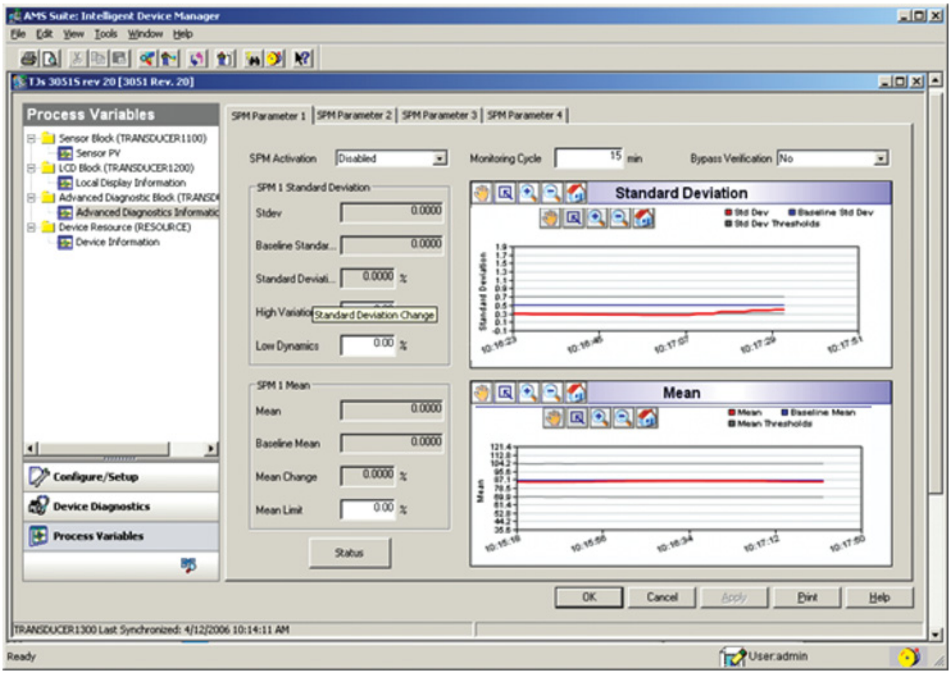

Figure C.11 Statistical data from transmitter over-sampling of a process (Screen image captured from Emerson Process Management's AMS ValveLink SnapOn).

Figure C.12 Wireless gateway (Screen image captured from Emerson Process Management's AMS Intelligent Device Manager, v11).

(a) (b) (c)

Figure C.13 *Wireless field instruments: (a) wireless temperature transmitter assembled to RTD thermowell sensor; (b) wireless pressure transmitter assembled to integral coplanar manifold; (c) 8705 magnetic flowmeter with supplementary wireless antennae (THUM) for diagnostics (Reproduced with permission of Emerson Process Management).*

A range of instruments can be used to communicate through each other to the gateway. The best communication path present will be used unless it is interrupted. In that case, an alternate route will be found by the devices to communicate the latest update. Multiple attempts will be made if needed to ensure the information is successfully communicated. Examples of wireless field devices are shown in Figure C.13. Some devices are available with an integral wireless antenna and are battery operated, while others can be retrofit with a wireless antenna that will run on power from the hardwired loop to communicate diagnostics wirelessly.

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 through the course.
3. Develop a working knowledge of the VMGSim Steady State and Dynamic Simulation Package.
4. Understand the fundamentals of steady-state and dynamic process simulation.

Course Coverage

This course will deal with the fundamental and underlying principles of automatic process control and simulation. The course covers the theory associated with single-input/single-output (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 and pressure, except

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measurement as it affects the control loop. In the course note, Chapter 2 provides a simplified summary of control loop hardware.

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 text book and the VMGSim Process Simulator (including user manual) are the only materials required for the course. The text book is 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 and Fisher. This vibrant area of chemical engineering is represented by the International Society of Automation (ISA), P.O. Box 12277, Research Triangle Park, NC 27709, USA (www.isa.org).

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 responses are characteristic of specific types of processes. 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 modelled through the use of a first-order differential equation. Shown in Figure 2.1 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.

Unlike first-order processes, second-order processes can have several different types of responses. Second-order processes are more complex than first order, and hence the mathematical models used to describe these processes are also more complex. There are three types of second-order systems 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 2.2. 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 $\xi = 1$, the system is deemed critically damped and has

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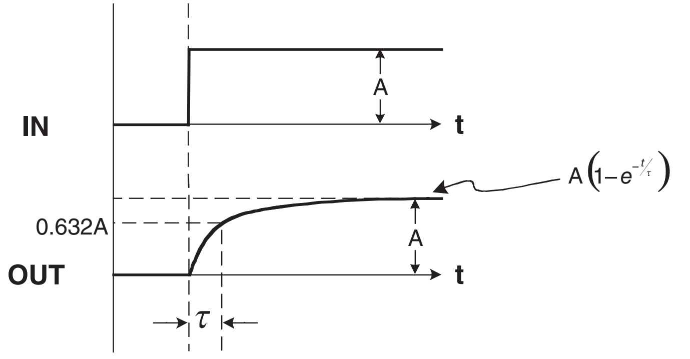

Figure 2.1 First-order process response to a step change.

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, 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. The larger the damping coefficient, 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 while a first order 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 systems that exist. There are higher order systems, such as third- or fourth-order systems. However, these higher order systems will not be discussed.

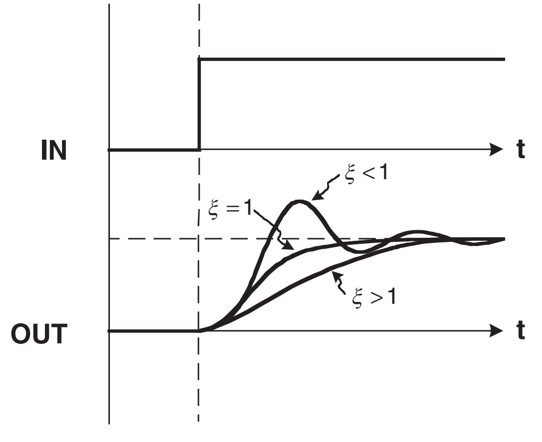

Figure 2.2 Second-order process responses to a step change.

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 single-input single-output (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 ordinary differential equation (ODE) for a single tank has been presented in the notes in Chapter 3 and is implemented in VMGSim as the Separator unit operation.

Build a simulation in VMGSim as follows:

In the Basis environment add a fluid package with the following:

Property Package: Advanced Peng–Robinson;

Components: water, N_2 and O_2 (N_2 and O_2 are used to simulate the air displacing water when the separator level drops).

Switch to the Simulation environment add a feed stream with the following data: Name = Feed to Separator, Flow = 20 kmol/h, Temperature = 15°C and Pressure = 1 atm, Composition: water = 100%, O_2 = 0% and N_2 = 0%.

Add a separator to your flow sheet. Complete the simulation by adding the two product streams as indicated in Figure 2.3.

Now the model should be totally solved in steady state.

Before switching to dynamics, you need to understand some basic rules for converting from steady-state model to a dynamic model in a simulation tool.

1. Save your case for the steady-state model.
2. Enter dynamic mode.
3. Size all relevant process equipments.
4. Add the valves and controllers to control the simulation:
   (a) Valves must be sized and controllers must be properly set up.
   (b) Use Selector Blocks to simulate disturbances, if desired.
5. Add the strip charts to monitor the process.
6. Set up the pressure/flow specification on all boundary streams.
7. Run the dynamics assistant.

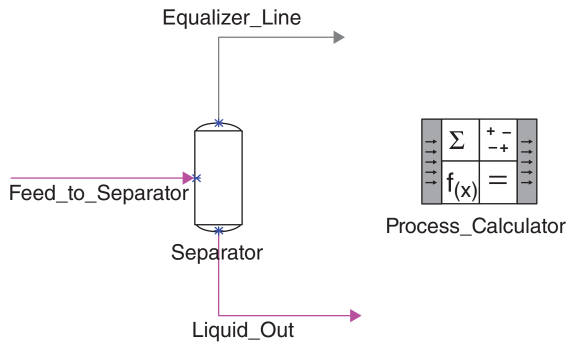

Figure 2.3 *Separator level response.*

8. Save your case (it is a good idea to save your case quite often when working in dynamics as the simulator is not able to go backwards in simulation time; the only way to go back is by using a saved case).
9. Start the integrator.

Now switch to dynamics, add Dynamic Pressure/Flow specifications to the boundary streams. For the feed and liquid product streams use flow specs (however, the flow of liquid product stream will be calculated by the Process Calculator later on), and for the Equalization Line use pressure spec. Also, specify the volume of the separator as 2 m^3.

Before we start running the model we need to take care of the equalization line that simulates the 'open to atmosphere' configuration of separators. Open the 'Equalization Line' stream form, on the mole composition part, enter the following composition specification: water = 0%, N_2 = 80% and O_2 = 20%. This product block will supply composition for the equalization line in case of flow reversal (which will be the case if the level in the separator starts to drop and it needs to be replaced by air).

Since you just switched from steady state, the flow specs will represent the steady-state condition (flow in = flow out), so the level in the separator will stay at 100%. Increase the outlet flow spec value and you will notice the level in the separator dropping.

Calculate the flow out of the separator using Equation 2.1, which describes a linear valve, and then Equation 2.2, which describes a non-linear valve. In both cases, the outlet flow rate is a function of the liquid head on the separator only:

$$\text{Flow} = K \times \text{head}, \tag{2.1}$$

$$\text{Flow} = K \times \sqrt{}(\text{head}). \tag{2.2}$$

In Equations 2.1 and 2.2, SI units are flow in m^3/h and head in metres. Use the Process Calculator unit operation to incorporate Equations 2.1 and 2.2 into the simulation. Import the Separator liquid level into the spreadsheet, this will be the head. Use this value of the head to calculate the liquid outlet flow. A K value of 20 would work well with SI units. Then export the calculated flow back to the 'Molar Flow' of the 'Liquid Out' stream. 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 separator in steps, up and down, and then observing the separator 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 the capacity-dominated process consisting of

(a) a single separator,
(b) two separators in series,
(c) two separators in series with a pipe segment between them (use only linear valves) and
(d) three separators in series (use only linear valves).

[Hints: Use a pipe segment and calculate a volume that should give a dead time of around 10 minutes. The number of increments in the pipe segment will affect the dead time – the more the increments, the closer the simulated dead time will be to the value calculated by fluid velocity and pipe length.]

- What is the open loop response of the separator level to a step change in the feed rate in a process with a single separator only?
- What effect does the addition of a second separator have on the level response in Separator 1? And in Separator 2?
- What effect does the addition of a third separator have on the level response in Separator 1? And in Separator 2? What is the open loop response of Separator 3?
- What effect does the volume of the separator 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 separators have on the response of the level in Separator 1 and Separator 2? Did the pipe segment properly simulate dead time in this situation? Why or why not?

2. Temperature Response

The next exercise in this workshop requires that you set up a mixing separator to heat water directly using live steam, as illustrated in Figure 2.4. Use the Multi-feed Separator volume of 2 m^3. Add a valve on the Liquid Out stream and a level controller (Process Variable = Multi-feed Separator Liquid Level Percent and Output Target Object = The liquid out valve) with a 50% set point. $K_{\mathrm{i}} = 1$ and $T_{\mathrm{i}} = 60$ minutes would be a standard configuration to the liquid level controller parameter. The feed water stream to the separator enters with a flow of 100 kmol/h at 15°C and 1 atm. The steam to the separator enters as saturated steam at 1 atm (since this stream will only bare a flow spec in dynamics, the pressure will float on the separator pressure and temperature will change with pressure to satisfy the saturated vapour condition).

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. Then, switch to the dynamic mode of operation and perform step response testing by varying the inlet water flow rate and feed temperature or steam flow rate 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 separator as in the previous exercise. Calculate a pipe segment volume to give approximately 10 minutes of dead time

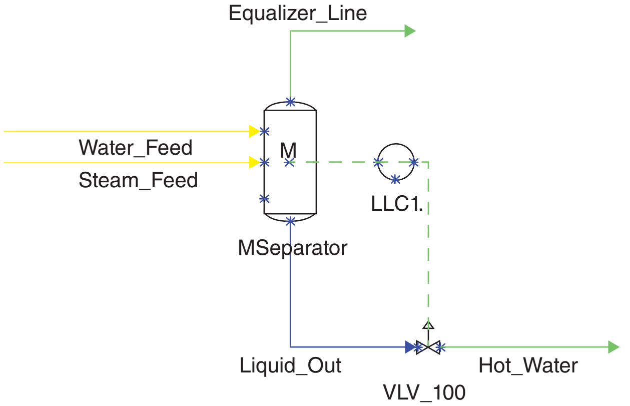

Figure 2.4 *Mixing separator process.*

and specify the volume in the 'Holdup' tab. Make sure that the number of sections in the pipe segment is set to at least 50 instead of the default of 1. Note that the holdup volume is for a single tray, so the total holdup volume should be 50 times of the holdup volume for a single tray. Repeat your analysis of step disturbances noting the relationship between the separator temperature and the temperature at the outlet of the pipe segment. Repeat the analysis again but with a different volume for the pipe segment and note any differences on 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 pipe segment properly simulate dead time in this situation? Why or why not?

Present your findings on CD, DVD or thumb drive in a short report using MS-Word. Also include on the submitted media a copy of the VMGSim 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 downside is that it results in very slow and long response times.
6. Dead time has no effect on the filtering capability of the process.
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.

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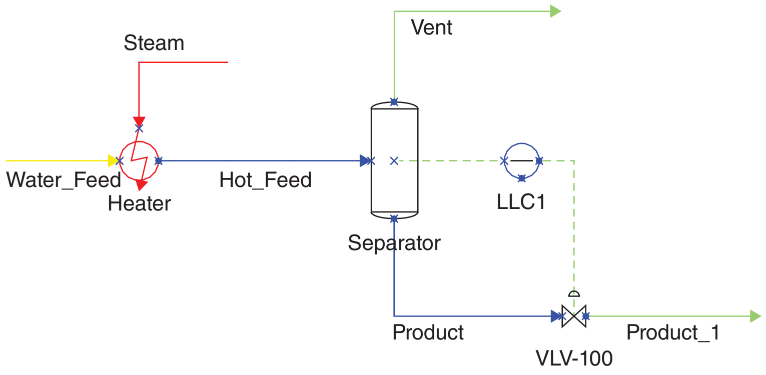


Figure 3.1 *Illustrative capacity-dominated process.*

Tasks

1. System Identification

The process used for this workshop is shown in Figure 3.1. Build a simulation of this system with the following basis:

Property Package: Wilson.

Components: water, methanol, N_2 and O_2 (N_2 and O_2 are used to simulate the air in the separator that displaces liquid when the separator level drops).

Switch to the simulation environment and add a feed stream with the following data: Flow = 100 kmol/h, Temperature = 30°C and Pressure = 200 kPa, Composition: water = 50%, methanol = 50%, O_2 = 0% and N_2 = 0%.

Add a heater and a separator to your flow sheet and specify the volumes of 0.1 m^3 for the heater and 2 m^3 for the separator. The duty stream 'Steam' should heat the stream to about 70°C. The hot mixture is then stored in the separator for future use.

'System identification' is the term used to define a procedure to characterize the process response. In this case, system identification can be accomplished by adding a level controller to the separator (add a controller and flow control valve on the liquid outlet stream), with a set point of 50%, adjusting the steam flow to the heater in steps, up and down, and then observing the temperature response on the strip chart. This is termed 'step response testing'.

Figure 3.2 illustrates a typical step-test response for a first-order system. The relevant process parameters of gain, K_p, time constant, τ, and dead time, t_{DT}, for this first-order process are shown and can be calculated as follows:

$$K_p = \text{Output}/\text{Input}, \tag{3.1}$$

where Output is the separator temperature change/temperature transmitter span and Input is the steam rate change/steam valve span.

τ = The time it takes for the separator temperature to reach 63.2% of its final value.

t_{DT} = The delay between the change in steam rate and the initial change in separator temperature.

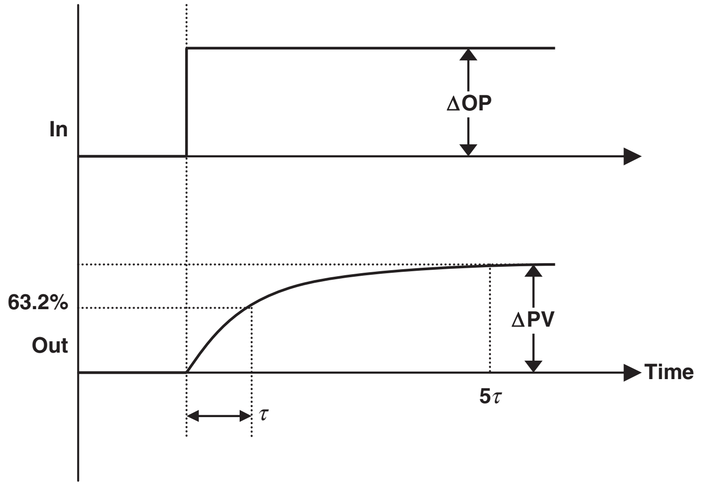


Figure 3.2 System step response.

- In order to check how linear the process is it is necessary to determine if the gain is the same regardless of the steam rate and to see if the magnitude of the gain is unchanged for increases and decreases in steam duty. Do this by using the step-testing method described above and Equation 3.1.

2. Capacitance

Now we will examine how the gain, time constant and dead time vary for different separator levels or different amounts of capacitance in the process shown in Figure 3.1.

- Make step changes in the steam rate for three different separator level set points of 5%, 50% and 95%. Calculate the gain, time constant and dead time for the three different process capacities. Present your results in Table 3.1 and then plot a graph of each of the time constant and gain versus separator level.

3. Attenuation

As has been indicated in the book, an object of both the process and the control system is to reject or at least minimize the effect of disturbances. In order to quantify disturbance

Table 3.1 Summary of process parameters.

Tank level (%)	Process gain	Time constant (minutes)	Dead time (minutes)
5			
50			
95			

Table 3.2 *Attenuation for level at 5%.*

Disturbance period (minutes)	Frequency (1/minute)	ΔT – product temperature (°C)	Attenuation
5			
10			
20			
30			
40			

rejection the term 'Attenuation' has been borrowed from the electrical/mechanical engineers and is defined as follows:

$$\text{Attenuation} = 1 - [(\text{Disturbance Amplitude Out})/(\text{Disturbance Amplitude In})]. \quad (3.2)$$

For example, if the separator temperature varies with an amplitude of 5°C and the input temperature disturbance has an amplitude of 25°C, the attenuation is (1 – (5/25)) = 0.8 or 80%.

In VMGSim, to generate a sinusoidal feed temperature a selector block unit operation is used. From the PFD tab, drag the selector block onto the flow sheet. Choose the 'Hand Sel' option under 'Selector Mode', select 'Sine wave' under 'Output Bias Fn' and specify the magnitude as well as the period. Connect to the inlet feed at 'output' region and select appropriate unit for the signal as required. Then, you need to enter the value for 'x' as 30, the input scale 'm' value for 1 and leave the input bias 'b' value for 0. The selector block automatically calculates the scaled input 'y' and the scaled output 'z'. Finally, start the integrator to see whether the selector block can properly generate the sine wave for the inlet temperature as expected. If your output value is still wrong or biased, make sure the 'Output Scale' is specified as 1, and 'Output Bias', 'Noise Magnitude' and 'Noise Decay Time' are specified as 0.

- Complete Tables 3.2–3.4 for each separator level and then plot the Attenuation in percent versus Disturbance Period in minutes, with a curve for each level.

[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.]

Table 3.3 *Attenuation for level at 50%.*

Disturbance period (minutes)	Frequency (1/minute)	ΔT – product temperature (°C)	Attenuation
10			
20			
30			
40			
100			

Table 3.4 *Attenuation for level at 95%.*

Disturbance period (minutes)	Frequency (1/minute)	ΔT – product temperature (°C)	Attenuation
10			
20			
30			
40			
100			

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 feed heater and a separator acting as a storage tank, shown in Figure 3.1, except that additional equipment will be added between the heater and the separator. The additional equipment will be a pipe segment with a volume of 0.3 m^3 and length of 2 m, which gives a dead time of 6 minutes for the specified inlet flow rate. Figure 3.3 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 separator.

Again this simple example is illustrative of many real plant situations. Even the time it takes for the fluid to move through pipes between connecting items of equipment is an example of dead time. A sensor located at 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 and 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 level controller with a set point of 50%, increase and decrease the steam rate to the heater and record the separator temperature response. What is the dead time for the

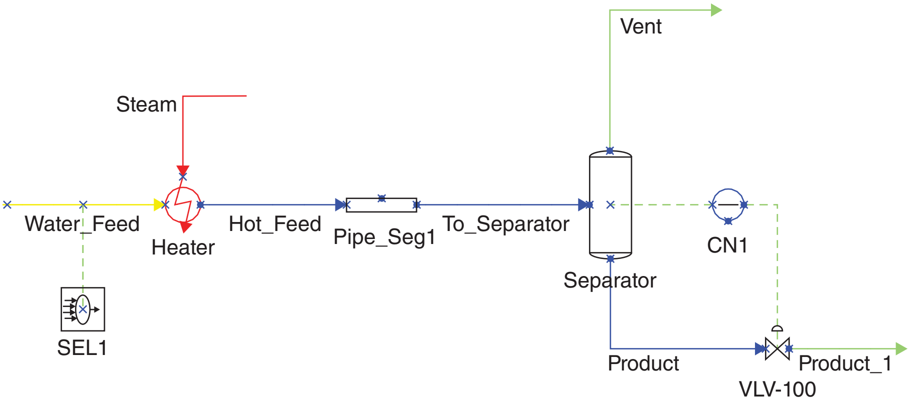


Figure 3.3 *A process containing dead time.*

Table 3.5 Dead time/time constant ratio.

Separator level (%)	Dead time (minutes)	Time constant (minutes)	Dead time/time constant
5			
50			
95			

process shown in Figure 3.3? 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 separator level between 5% and 95%. 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, the dead time has little or no effect on the process response. However, if the ratio is greater than 0.3, 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 3.5.
- To test the hypothesis that dead time has no effect on the open loop process attenuation for a capacity-dominated process, again perform a response test using the sine wave input. To test the hypothesis, run the 40 minutes, 25°C amplitude disturbance through the process of Figure 3.3 with the separator level set to 50%. Has the attenuation changed from the value you calculated earlier?

Present your findings on CD, DVD or thumb drive in a short report using MS-Word. Also include on the submitted media a copy of the VMGSim 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.

A Real-Time Approach to Process Control, Third Edition. William Y. Svrcek, Donald P. Mahoney and Brent R. Young.

5. Derivative action can only be used where there is no significant process noise and relatively little dead time.
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% level and fully open at 75% level. $K_c > 1.0$ won't hold the level between 0% and 100%.
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, reduce K_c and add integral action to ensure that the level stays in the tank ($K_c \times T_i = 4.0$).
   - If the hold-up time is long, you do not need any integral action.
   - If the level is cycling, 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 4.1. Check that you still have the Wilson property package specified. The only components required are water and methanol, and O_2 and N_2 for the vapour phase hydraulics. The inlet temperature should be 30°C, while 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 pipe segment 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, separator outlet temperature, heater duty, feed molar flow. Select suitable ranges for each variable.

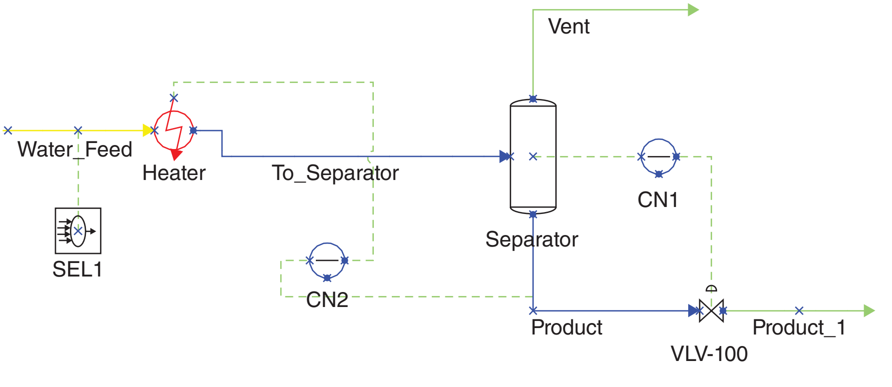


Figure 4.1 Low-capacity, no dead time process.

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 separator level accordingly. Your simulation should still contain a transfer block operation, which adds noise to the system. Set the feed water temperature disturbance period to 30 minutes and its amplitude to 25°C.

Now add a controller for the separator outlet temperature. The controller should directly manipulate the heater duty between 0 and 1×10^6 kJ/h. Please note that since VMGSim does not support direct control on energy streams, you will need to select the manipulated variable from the heater InQ. The PV range should be 0–100°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 disturbances 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 pipe segment to the system shown in Figure 4.1 in order to simulate dead time. The pipe segment should have a length of 2.0 m, a total volume of 0.5 m^3 (dead time = 10 minutes) and a pressure drop of 0 kPa. The process layout is presented in Figure 4.2. Remember that you want to work with a capacity-dominated process, so ensure that the separator level is set accordingly. Also remember to save your case after setting up the controller, since an unstable system may stop the integrator and the system will be difficult to restart.

- 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 minutes. Vary the size of the PFR to change the amount of dead time in the system. Record the results in Table 4.1.

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

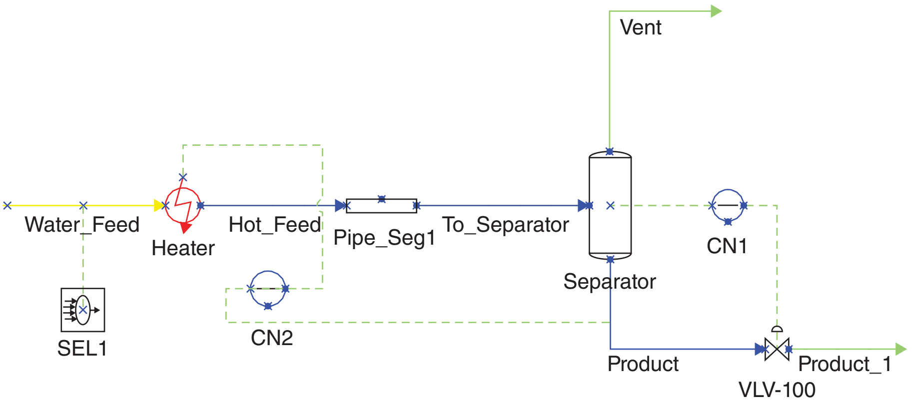


Figure 4.2 *High-capacity process dead time.*

the requirements of an effective controller. The other main requirement is that we should be able to change the set point whenever we want 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, disturbance rejection and controller performance, are used to assess the effectiveness of a controller.

Eliminate dead time from the system by deleting the pipe segment 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 separator outlet temperature set point to 80°C.

- Where does the separator outlet temperature stabilize? The difference between this value and the set point is called offset.
- Plot the relationship between offset and the controller gain.
- Can you explain the relationship shown on the plot between offset and the controller gain in terms of proportional-only controller equation given in Equation 4.1?

$$\text{Output} = \text{Gain} \times (\text{SP} - \text{PV}) + \text{Bias}. \tag{4.1}$$

4. PI and PID Control

The feedback controller we have employed to this point has only contained one term: gain. As suggested above, this type of controller is called proportional-only controller, which

Table 4.1 *Process attenuation with dead time.*

Dead time (minutes)	ΔT – product temperature (°C)	Attenuation
2		
5		
10		
20		

Table 4.2 PI controller optimization.

Test	Gain	Integral time (minutes)	Offset	Time to steady state (minutes)

suffers from the problem of offset. Offset can be reduced with high 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: integral time. The control equation for a proportional-integral (PI) controller is

$$\text{Output} = \text{Gain} \times \left((\text{SP} - \text{PV}) + \frac{1}{T_\text{i}} \int_0^T (\text{SP} - \text{PV})\, \text{d}t \right). \tag{4.2}$$

- How does PI control eliminate offset? Use Equation 4.2 to help explain.

Add integral time to your controller, starting with $T_\text{i} = 1.0$ minutes, 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, that is, step response testing.

- Summarize your results for PI control in Table 4.2. Record under the 'Test' column the details of each type of step test performed, that is, 30–40 kmol/h.

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 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

$$\text{Output} = \text{Gain} \times \left((\text{SP} - \text{PV}) + \frac{1}{T_\text{i}} \int_0^T (\text{SP} - \text{PV})\, \text{d}t + T_\text{d} \frac{\text{d}(\text{SP} - \text{PV})}{\text{d}t} \right). \tag{4.3}$$

Derivative time increases the controller response when the controlled variable is moving away from its set point most quickly, that is, straight after a disturbance has affected 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), the derivative action will

Table 4.3 PI controller optimization.

Test	Gain	Integral time (minutes)	Derivative time (minutes)	Offset	Time to steady state (minutes)

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, starting with $T_d = 1.0$ minutes, to your system 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 the process disturbances and the ability to track a set point. Record your results in Table 4.3. Under 'Test' column, record the type of step change test performed.

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 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 the 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 and reboiler duty will have the desired effect on product compositions and yields without introducing additional lag to the system.

In order to better understand how averaging level control works, build a simple system consisting only of a 2 m^3 separator. The feed to the separator should be 250 kmol/h of water at 25°C and 100 kPa. Add a level controller and enter a set point of 50%. Finally add a feed disturbance using a selector block unit operation set up to vary the feed rate

Table 4.4 *Averaging level control.*

Disturbance period (minutes)	$K_c = 1.0$ $T_i = 50$ minutes	$K_c = 2.0$ $T_i = 25$ minutes	$K_c = 0.5$ $T_i = 100$ minutes
5			
10			
20			
30			

sinusoidally with an amplitude of 25 kmol/h. You need to be aware of the control valve CV value.

- Test the following combinations of PI control for the separator level controller for the range of disturbance periods given in Table 4.4:
  1. $K_c = 1.0$, $T_i = 50$ minutes;
  2. $K_c = 2.0$, $T_i = 25$ minutes;
  3. $K_c = 0.5$, $T_i = 100$ minutes.
- 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 CD, DVD or thumb drive in a short report using MS-Word. Also include on the submitted media a copy of the VMGSim 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 VMGSim 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 two 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 no control action occurs and the process response can be isolated.

To generate a process reaction curve, the process is allowed to reach the steady state or as close to steady state as possible. Then, in open loop so there is no control action, a small step disturbance is introduced and the reaction of the process variable (PV) is recorded. Figure 5.1 shows a typical process reaction curve for the PV generated using the above method for a generic self-regulating process. The term self-regulating refers to a process

A Real-Time Approach to Process Control, Third Edition. William Y. Svrcek, Donald P. Mahoney and Brent R. Young.

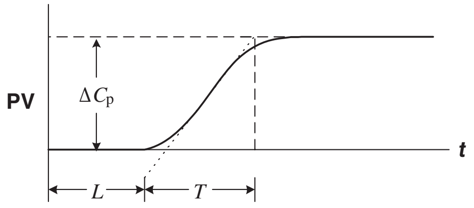


Figure 5.1 Process reaction curve.

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 are as follows:

L = lag time (minutes);
T = time constant estimate (minutes);
P = initial step disturbance (%);
ΔC_p = change in PV in response to step disturbance, (change in PV)/(PV span) × 100, (%);
$N = \dfrac{\Delta C_p}{T}$ = reaction rate (%/min);
$R = \dfrac{L}{T} = \dfrac{NL}{\Delta C_p}$ = lag ratio (dimensionless).

Z-N Process Reaction Curve Tuning Method for a PI Controller

1. Determine a *reasonable* value for the step valve change, P. This value is arbitrarily chosen, but typically 5% is reasonable.
2. With the controller in *manual* mode, manually move the valve 'P'%.
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:

$$\text{Controller gain} = K_c = 0.9\left(\frac{P}{NL}\right),$$

$$\text{Controller integral time} = T_i = 3.33(L).$$

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 quarter decay ratio (QDR).

Auto Tune Variation (ATV) 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

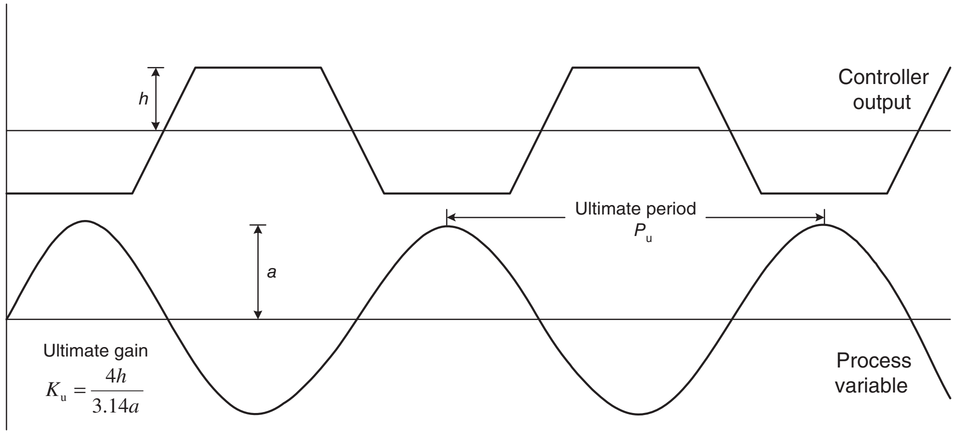


Figure 5.2 *ATV critical parameters.*

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 5.2 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.

General ATV Tuning Method for a PI Controller

1. Determine a *reasonable* value for the valve change, h. This value is arbitrarily chosen, but typically 0.05 is reasonable, that is, 5%.
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 5.2.
6. Record the value of the amplitude, a, by picking it off the response graph.
7. Perform the following calculations:
   ultimate period $= P_u =$ period taken from the limit cycle;
   ultimate gain $= K_u =$ 4h/3.14a;
   controller gain $= K_c = K_u/3.2$;
   controller integral time $= T_i = 2.2P_u$.

Closed Loop Tuning Technique

The closed loop technique originally developed by Ziegler–Nichols (Z-N) is another technique that is commonly used to determine the two important system constants, ultimate period and ultimate gain. Historically speaking it was one of the first tuning techniques to be widely adopted, although it is aggressive for process control and other tuning recommendations are recommended (e.g. Cohen–Coon or IMC tuning).

In 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, closed loop tuning is done by disturbing the closed loop system and using the disturbance response to extract the values of these constants.

Closed Loop Tuning Method for a PI Controller

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:
   ultimate period, P_u = period taken from limit cycle;
   ultimate gain, K_u = controller gain that produces the limit cycle;
   controller gain = $K_c = K_u/2.2$.
   Z-N Recommendation (Note: Aggressive for process control)
   controller integral time = $T_i = P_u/1.2$.
   Z-N Recommendation (Note: Aggressive for process control)

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 auto tune variation (ATV) tuning technique is a powerful method for many loops.
7. The closed loop technique is also useful, but more aggressive than ATV, especially with Z-N recommendations.
8. The process reaction curve technique is also useful, as it provides estimates for the key process parameters.
9. VMGSim 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 5.3. A 50/50 feed mixture of water and methanol ($T = 30°C$, $P = 200$ kPa, $F = 100$ kmol/h) is heated in a steam heater to approximately 70°C. The hot stream passes through a dead time lag before being stored in a separator for future use. Use a pipe segment unit operation to simulate the dead time with volume of 0.3 m^3 and length of 2 m. This was the process you worked on in Workshop 4.

Set the separator level to 50% with no incoming disturbances. With the temperature controller in manual, adjust the steam valve to get a separator temperature of approximately 70°C. Bring up the temperature controller faceplate.

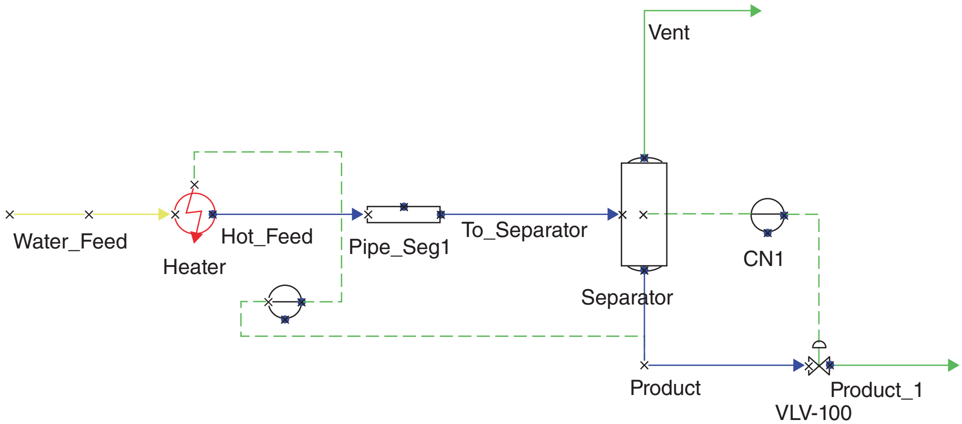


Figure 5.3 *Illustrative capacity plus dead time process.*

First use the process reaction curve technique to determine the controller settings at 50% tank level. Determine the controller settings at two more tank levels (5% and 95%).

Second, use the ATV technique to determine the controller settings.

Bring the process to a limit cycle by alternating the OP of the controller by 10% in each direction, with a constant interval between changes long enough for the PV to change by 5°C (see Figure 5.4).

- 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 separator 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 separator levels (5% and 95%).
- Now use the closed loop tuning technique to determine the controller settings.
- Compare the results using both the ATV and Z-N tuning.

2. Controller Contributions to Attenuation

We have seen in Workshop 3 that the process itself is able to attenuate with no control, that is, 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 (closed loop) attenuation we can determine what the controller itself contributes to the overall process attenuation. Delete the pipe segment and configure the temperature controller set point to 60°C to avoid liquid feed boiling.

- Determine the total closed loop attenuation of the separator operating at the 50% level for sinusoidal disturbances of periods 10, 20, 30, 40 and 100 minutes and an amplitude of 25°C.

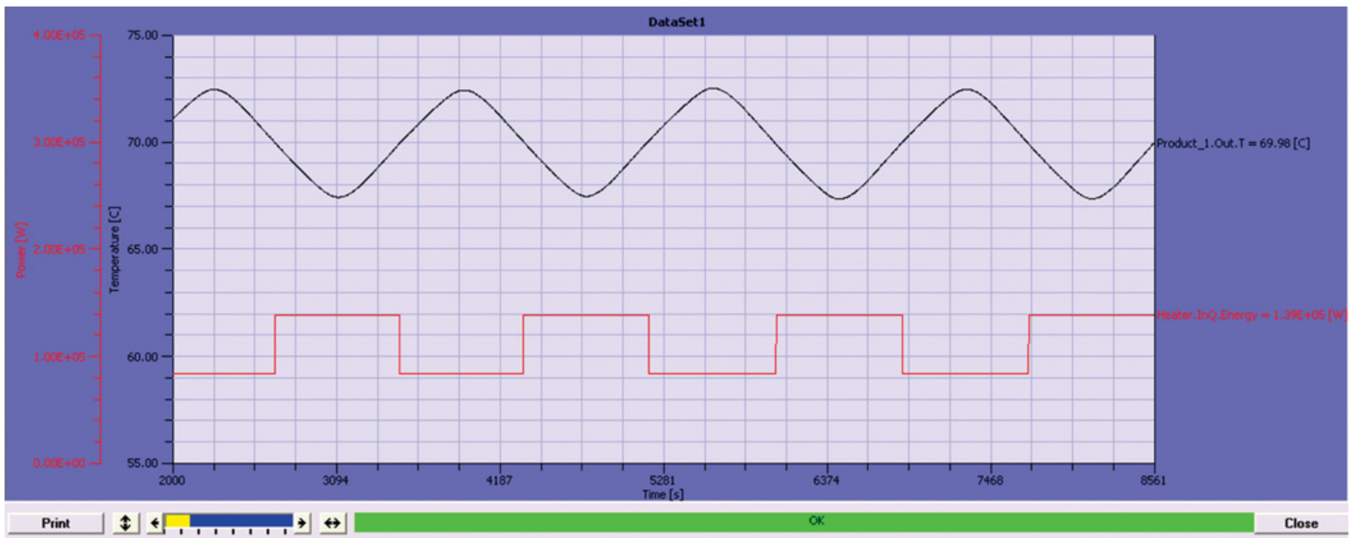


Figure 5.4 Process limit cycle.

- Compute the controller contribution to attenuation for these disturbances.
- At 5% level determine the controller attenuation for sinusoidal disturbances of periods 5, 10, 20, 30 and 50 minutes and amplitude 25°C.
- At 95% level determine the controller attenuation for sinusoidal disturbances of periods 10, 20, 30, 40 and 100 minutes and amplitude 25°C.
- Plot attenuation versus the logarithm of the disturbance period. Briefly comment on your results.
- Dead time in the system might have a considerable negative impact on the system. Add the pipe segment back to simulate the dead time and redo all the tests above. Briefly comments on your results and findings.

Present your findings on CD, DVD or thumb drive in a short report using MS-Word. Also include on the submitted media a copy of the VMGSim 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 (FBC) 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 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 non-linear processes, where non-linearities exist between the disturbance measurement and controlled variable.

A Real-Time Approach to Process Control, Third Edition. William Y. Svrcek, Donald P. Mahoney and Brent R. Young.

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.
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 loops, 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 6.1 using the Wilson package. The feed is pure water with a temperature of 20°C, atmospheric pressure and a flow rate of 1.5 m^3/h. 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, that is, set equal to zero. Incorporate dead time into the process by adding a pipe segment 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 separator with a volume of 1.2 m^3 and set the liquid level set point at 50%. 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, product molar flow. Select suitable ranges for each variable and iconize the strip chart view for later use.

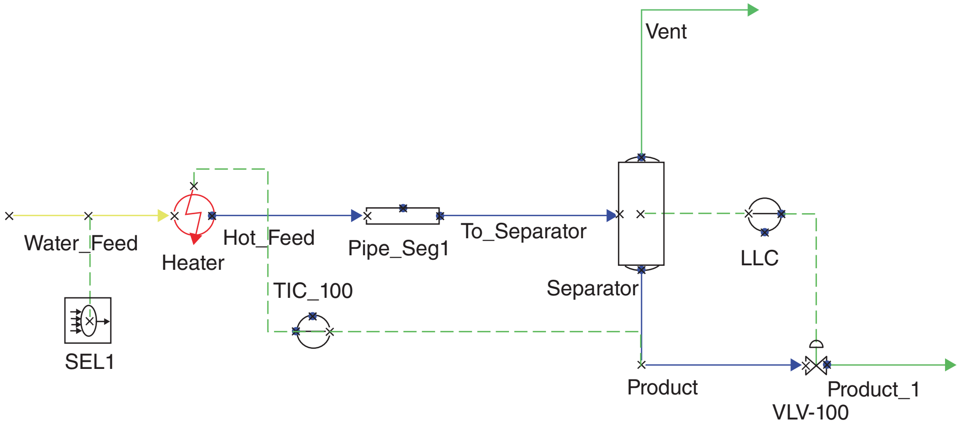


Figure 6.1 Simple heating system.

Add two controllers to the system to set up FBC of the process:

1. The first controller should manipulate the steam rate to the heater between 0 and 5 $\times$ 10^5 kJ/h (direct Q) to control the product temperature between 0°C and 100°C. The set point should be 55°C initially, with tuning constants of $K_c = 0.5$ and $T_i; = 10$.
2. The second controller should manipulate the product flow from the separator between 0 and 200 kmol/h to control the separator level between 0% and 100%. The set point should be 50% 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.

Finally add feed noise to the system using the Selector Block unit operation. The selector block is connected to the feed temperature. Create sine wave noise with an amplitude of 10°C and a period of 10 minutes. The feed temperature should still oscillate around a mean of 20°C.

2. Determine Base-Line Control Performance

The heater–tank system, shown in Figure 6.1, 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, that is, 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 6.1 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 FBC system simulated above. If necessary, vary the temperature controller tuning constants to try to improve the performance of the control loop.
  [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, that is, step response testing?]

Table 6.1 *Base-line control performance.*

Disturbance period (minutes)	Frequency (1/min)	ΔT – product temperature (°C)	Attenuation
5			
10			
20			
40			
60			

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 completely cancel out the effect of measured disturbances 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 FBC 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 process calculator function in VMGSim, 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 FBC to see if process response can be improved.

In order to obtain the actual feedback duty directly from the TIC-100 controller, create a fake heater with two fake streams (you can input the stream specifications as you like) and connect to the output of TIC-100 controller. If you are interested, you can find other ways to obtain the feedback duty without creating the fake heater.

Set up the following titles in cells Al–A6:

Al	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
A8	Feedback Duty
A9	Total Duty

Complete the spreadsheet as follows:

B1	Import the feed temperature from the process feed stream.
B2	Import 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	Connect to 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%?]*

B7 The feedforward duty can be calculated from Equation 6.1. Incorporate this equation into the spreadsheet.

$$\Delta Q = \frac{\Delta T}{K_{\mathrm{p}1}} \times \frac{\mathrm{SVSpan}}{K_{\mathrm{p}2}}, \tag{6.1}$$

where ΔQ represents the changes in heater duty required to produce 1°C change in the product temperature; ΔT is Nominal Feed Temperature – Actual Feed Temperature; $K_{\mathrm{p}1}$ is Process Gain 1 between the product temperature and the feed temperature; SVSpan is Steam Valve Span; $K_{\mathrm{p}2}$ is Process Gain 2 between the product temperature and the heater duty.

B8 Import the heat duty from the fake heater.
B9 Add the FF duty from cell B7 to the FB duty from cell B8.

Equation 6.1 is not necessarily exact at all values of the feed temperature due to the process non-linearity. An exact expression is not necessary for successful feedforward control; even if the calculated duty is incorrect by 50%, the controller will still perform better than with no feedforward action.

Export the result in cell B9 to the heater duty port. Your process should be similar to the one shown in Figure 6.2.

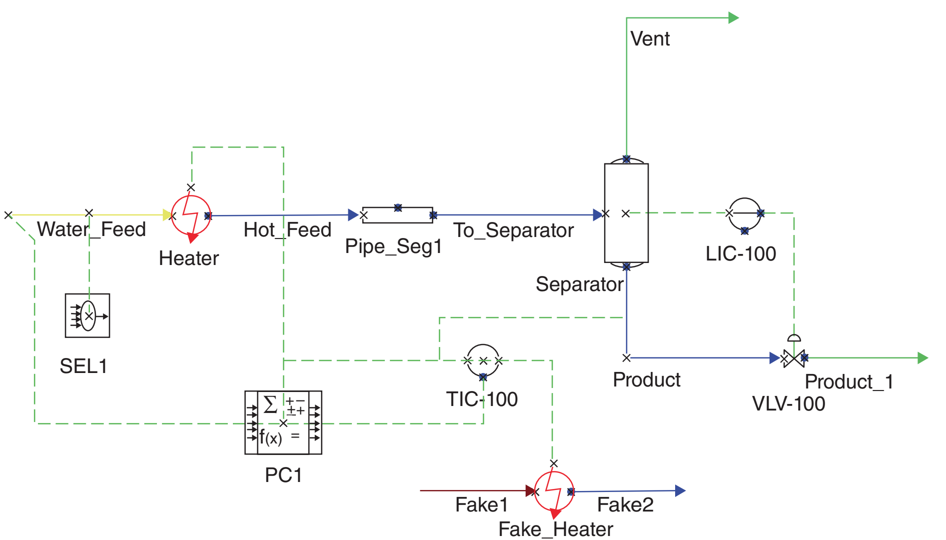


Figure 6.2 *Feedforward control system.*

Table 6.2 *Feedforward control performance.*

Disturbance period (minutes)	Frequency (1/min)	ΔT – product temperature (°C)	Attenuation
5			
10			
20			
40			
60			

- Test the feedforward controller for the same range of feed temperature disturbances that you analysed for FBC in the previous task. (Due to the complexity of the feedforward controller designing, you might observe minor offset between the set point and PV in this case. If you are interested, verify the reason and find a solution to solve the offset problem.) Record the results in Table 6.2.
- 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.]*
- 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 responds faster 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% of the time constant of the master loop for cascade control to be effective. 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 (separator temperature) should provide the set point for the slave loop and the slave controller should manipulate the steam rate directly. The process is shown in Figure 6.3.

To implement cascade control into your existing simulation, delete the feedforward controller but retain the separator level controller and the original separator temperature controller. The separator temperature controller will now be the master controller for the cascade loop.

Add another controller unit operation to the flow sheet. 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. Connect the SP point (cascaded set point source) to

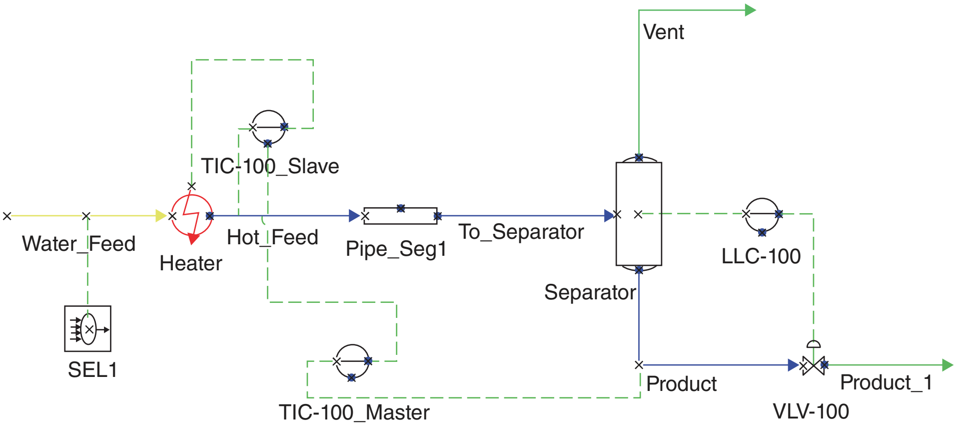


Figure 6.3 *Cascade control system.*

the PV point of the master controller. Configure the controller mode to 'cascade' for the slave controller otherwise it will not work properly. The slave loop should be tuned tightly; specify a gain of 5 and an integral time of 10 minutes. The master loop can be tuned more loosely; specify a gain of 1.0 and an integral time of 10 minutes. 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 6.3.

- 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 in Table 6.3.
- Try varying the tuning constants for both the slave loop and the master loop. Which combination(s) of tuning constants work best?
- 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, that is, is there too much control action?]

Table 6.3 *Cascade control performance.*

Disturbance period (minutes)	Frequency (1/min)	ΔT – product temperature (°C)	Attenuation
5			
10			
20			
40			
60			

- 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 the 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 separators and a mixer using the Wilson thermodynamic package. Pick any two components that are at 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 and the second stream to 100 kg/h. These flows are consistent with the desired ratio of 4:1 between components A and B. The separators are used to simulate dead time in the system so choose relatively small volumes for the separators and locate them in series with the first stream. Both separators should be on level control rather than liquid flow control. Beware of the CV value of both control valves. Simulate process noise with a sine wave input to the first stream using an amplitude of 50 kg/h and a period of 10 minutes. The system should resemble the one shown in Figure 6.4.

Incorporate ratio control via a process calculator. Import the flow on the first stream into cell Al. 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 results of cell A3 to the mass flow of the second stream.

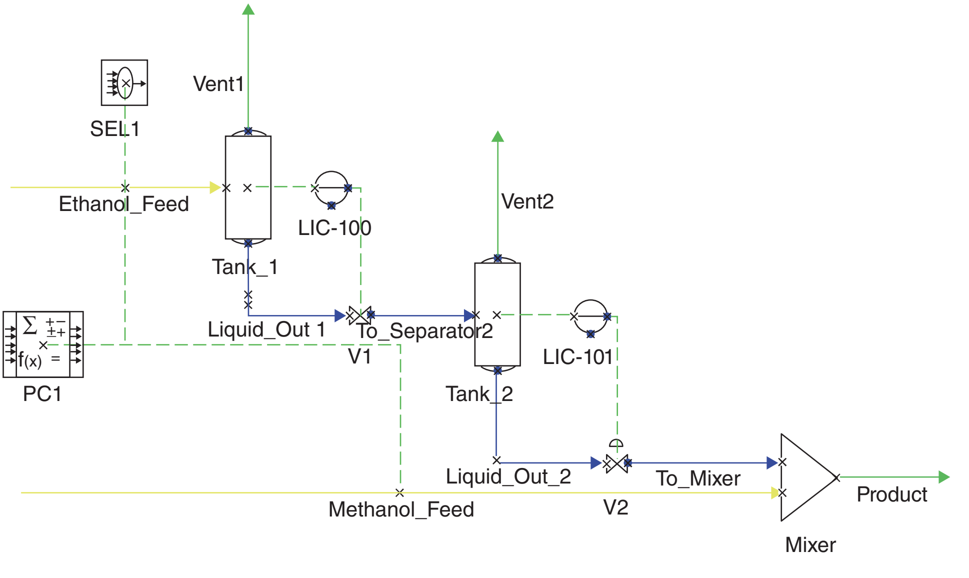


Figure 6.4 *Ratio control system.*

Run the simulation in dynamic mode with several values of disturbance period. Watch how the combined flow rate and concentration in the product stream changes with different conditions. You may need to reduce the integrator step size to see the effects of very high frequency disturbances (period > 5 minutes).

- How effective is the ratio controller at filtering out low-frequency noise?
- How effective is the ratio controller at filtering out high-frequency noise?
- 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 CD, DVD or thumb drive in a short report using MS-Word. Also include on the submitted media a copy of the VMGSim 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 compared 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 (MIMO) control problem. It is critical that variable pairing is done appropriately between controlled variables 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 two degrees of freedom for steady state and five degrees of freedom present for dynamic 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: $\text{DF} = N_{\text{cv}} = N_{\text{mv}} = N_{\text{valves}}$.

A Real-Time Approach to Process Control, Third Edition. William Y. Svrcek, Donald P. Mahoney and Brent R. Young.

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.
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, which is usually referred as inferential control.
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. Basic Process Configuration

The distillation column shown in Figure 7.1 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% propane, 40% isobutane, 40% *n*-butane and 15% isopentane. The total flow rate is 40 000 bbl/day at 720 kPa and 30°C.

Use the Advanced Peng–Robinson property package and the above information to build a steady-state simulation model of the distillation column. The column contains 20 trays and a total condenser (a condenser vapour rate spec value of 0 can be used to simulate 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. Under steady-state design, size the distillation column using the 'tower sizing' function, enter the weir height as 10 cm and tray spacing as 60 cm, select the column

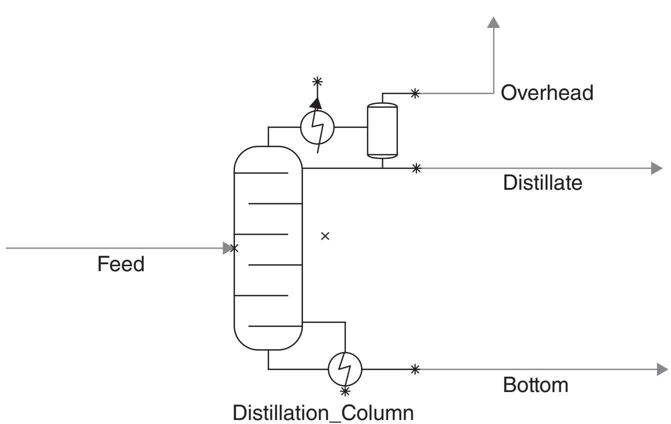


Figure 7.1 Stabilizer.

type as 'GlitschValve', then run the 'auto section design'. The automatic design result will be shown at the bottom and you will need these values in the dynamic model design. Your sizing result should look like this:

Configuration | Results

Section	Section_1	Section_2
Start	2	10
End	9	19
internalType	GlitschValve	GlitschValve
⊟ **Design Variables**		
DesignFloodFactor [Fraction]	0.8000	0.8000
SystemFactor [Fraction]	1.00	1.00
DowncomerClearance [cm]	3.810	3.810
WeirHeight [cm]	10.000	10.000
TraySpacing [cm]	60.000	60.000
GlitschValveType	V1	V1
ValveMaterial	SS	SS
ValveGage	14	14
ValvesPerActiveArea [Valves/m2]	129.17	129.17
DeckGage	14	14
LiquidFlowPerWfpDesign [(m3/s)/m]	0.0199	0.0199
LiquidFlowPerWfpMaximum [(m3/s)/m]	0.0497	0.0497
SectionAreaFactor [Fraction]	0.8000	0.8000
⊟ **Calculated Variables**		
ActiveArea [m2]	5.40	6.31
DowncomerArea [m2]	0.5941	1.26
TrayDiameter [m]	2.8956	3.3528
NumberofTrayPasses	1	3
TotalTrayArea [m2]	6.59	8.83
HoleArea [m2]	0.8250	0.9651
WeirLength [m]	2.0436	8.3477
Fpl [m]	2.0514	0.7247
Wfp [m]	2.6309	8.711

Make three specifications for all three degrees of freedom at the 'Spec/Estimate' tab, which are the flow of condenser vapour stream is 0, the C_3 fraction of bottoms product is 0.01 wt% and the column is operating with a reflux ratio of 1.0.

View the results to get a feel of the column in steady-state mode.

[Hint: As VMGSim only have distillation columns with a partial condenser, the third degree of freedom in steady state is to set the overhead flow of the condenser as zero.]

Investigate the steady-state model and try to find out the answer for the below questions:

- How to obtain the reflux ratio, reboil ratio, reflux flow rate and reboil flow rate in your steady-state model?

- How much is the value for the condenser duty and reboiler duty?
- What is the temperature and pressure on trays 1, 5, 10, 15 and 20?

Note down all these values as they are the baseline and will be useful in later tasks.

2. Distillation Control Design via Steady-State Analysis

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 7.1 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 C4+ components and
- Overhead temperature to meet utility requirement.

How many process objectives can be met simultaneously by the control system? A simple distillation column with no side draws and a total condenser has five degrees of freedom. A partial condenser (where there are two overhead products) adds one extra degree of freedom, as does each side draw. The degrees of freedom correspond to the number of control valves in the system:

- Distillate flow,
- Bottoms flow,
- Reflux flow,
- Cooling medium to the condenser (condenser duty),
- Heating medium to the reboiler (reboiler duty),

Three inventory/capacity variables must always be controlled in distillation:

1. Pressure,
2. Reflux accumulator level,
3. Reboiler sump level.

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.

Table 7.1 *Summary of process parameters.*

Feed rate (bbl/day)	Base case 40 000	Minimum flow 20 000	Maximum flow 50 000
C_3 wt%	5	1	12
i-C_4 wt%	40	44	33
n-C_4 wt%	40	45	37
i-C_5 wt%	15	10	18

After the inventory and capacity variables are controlled, two further controlled variables can be fixed by the control system. In the stabilizer discussed above, we have already noted that we would like to control the propane content of the bottoms product. This still leaves one degree of freedom unused. For this example, we will assume that the last degree of freedom will be used to control the energy use. We could also have picked another composition to control or a product yield and so on.

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 7.1 should be used to test any control system design.

Simulations allow essentially all types of variables 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 given below.

Control Strategy Selection Using Steady-State Analysis

1. Select two pairs of controlled variables. Note the values of these variables from the base case solution (see Task 1).
2. Modify the specifications used in the base case solution to incorporate the candidate controlled variables and values from step 1. Solve the column with the new specifications.
3. Repeat step 2 for the expected extremes in flow rate and composition. Record the values of the control objectives for all cases.
4. Determine which combination of controlled variables produces the best overall control performance considering the full range of feed variance.

5. After control variable pairings have been established to control the feed split and column fractionation, pair variables to provide inventory and capacity control.

Often the variable pairings required for inventory and capacity control will 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 7.1 with the following control objectives:

Case 1:

- wt% propane in the bottoms. (primary obj)
- fixed energy input. (secondary obj)

Case 2:

- wt% propane in the bottoms. (primary obj)
- fixed reflux flow rate. (secondary obj)

- Table 7.2 supports you with a good starting point of several pairs of controlled variables. Fill in all values in the table for the candidate control system.
- Are there any particular advantages or disadvantages between these pairing?

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 7.1) 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 both cases in Table 7.1.
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.

3. Dynamic Column Control Configurations and Responsiveness

Dynamic Control Configurations for a Distillation Column

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, that is, levels and pressures. The remaining two degrees of freedom

Table 7.2 *Control strategy selection using steady-state analysis.*

Primary controlled variable	Set point	Secondary controlled variable	Set point	Control objectives					
				Base case		Min. flow		Max. flow	
				Prim. obj	Sec. obj	Prim. obj	Sec. obj	Prim. obj	Sec. obj
Tray 5 temp.		Reflux flow rate							
Tray 10 temp.		Reflux flow rate							
Tray 15 temp.		Reflux flow rate							
Tray 5 temp.		Reboiler duty							
Tray 10 temp.		Reboiler duty							
Tray 15 temp.		Reboiler duty							

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, liquid flow down the column, reflux rate;
V, vapour flow up the column, boilup or reboiler duty;
D, distillate rate; and
B, bottoms rate.

The relationship between boilup 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.

The LV control configuration is often described as an energy balance configuration while 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 by changing one of the product rates. However, the LV configuration only affects the feed split indirectly through the level controllers.

- Complete Table 7.3 by listing the manipulated variables (mv) for each of the controlled variables.

The basic distillation control configurations have been listed above. However, there are many other configurations which use linear or even non-linear 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 boilup 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 or a fixed reflux flow rate. Describe which of the three control configurations listed in Table 7.3 could be set up to satisfy these control objectives.

Table 7.3 *Distillation column control configurations.*

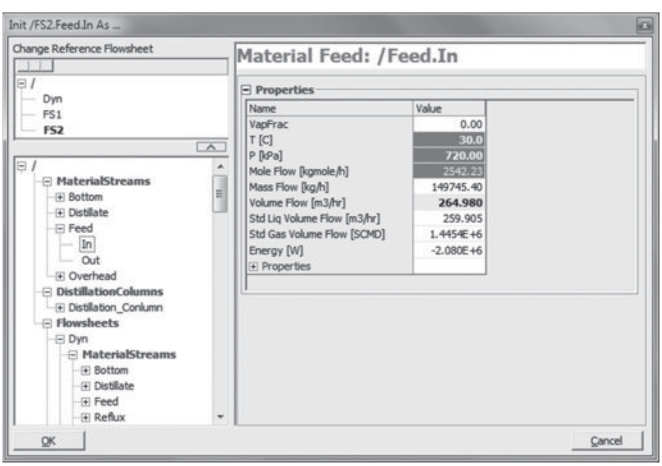
Control configuration	MV for reflux accumulator level control	MV for reboiler sump level control	MV for primary composition control	MV for secondary composition control
LV				
DV				
LB				

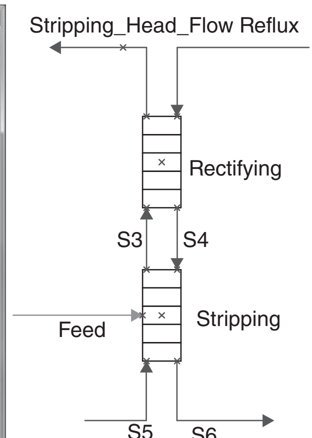
Building a Dynamic Model for a Distillation Column

After you get familiar with common control configurations for a distillation column, a dynamic model should be built in VMGSim to help you further comprehending the theory. In this workshop, LV and DV control configurations are selected and we will use LV configuration for the illustration.

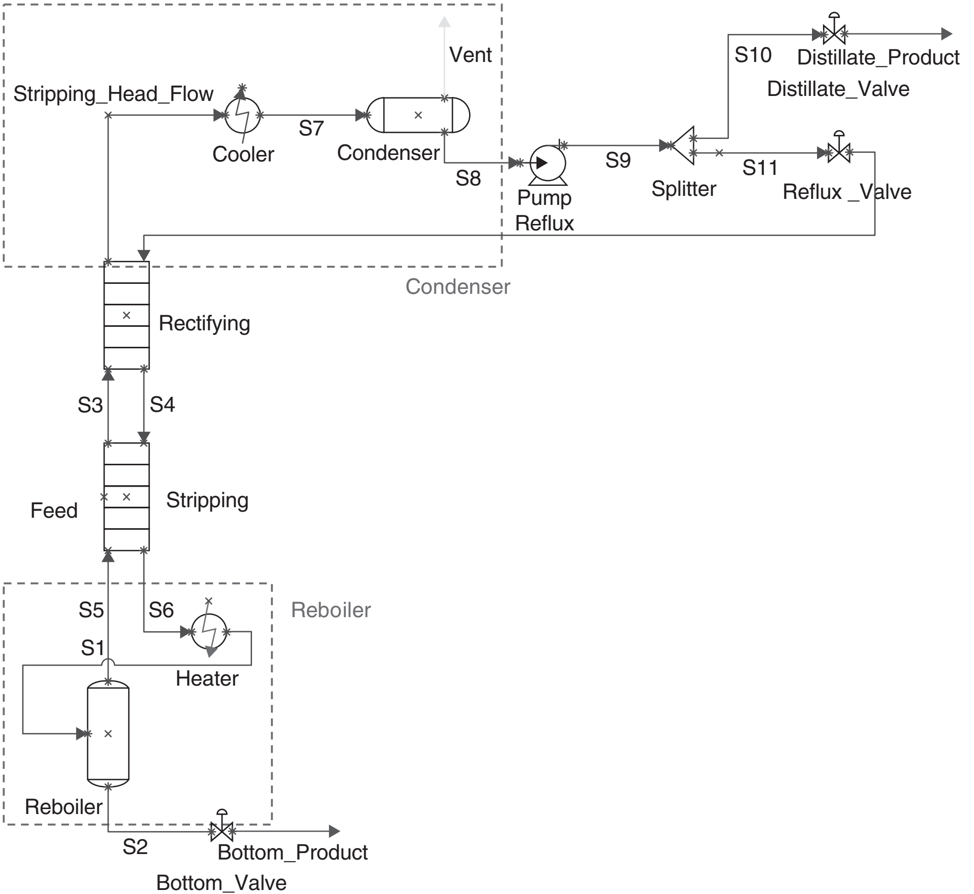
Building a dynamic distillation column in VMGSim is a totally different idea from your pervious workshops. Instead of directly converting from steady-state distillation column model to dynamic model (which is not supported by VMGSim for a distillation column), you will need to construct your own distillation column sections, condenser, reboiler and control loops.

1. On your steady-state flow sheet, add a new flow sheet unit operation. Change the active engine of the flow sheet to 'dynamics' and set the integrator execution mode to 'StdAlone'.
2. Add two distillation sections on your new flow sheet, simulating the stripping and rectifying section of the distillation column. The tray numbers and configurations should follow the column design in the steady state. Note that you will only need a total of 18 trays in dynamic model since tray 1 is considered as condenser and tray 20 is considered as reboiler in steady-state model.
3. Add a feed input on the stripping section and connect to the feed stream. Initialize the feed from the steady-state flow sheet (you will need to click the 'change reference flow sheet' and choose '/', then find the feed under the material stream, see below figure for more illustration). Make specifications to the temperature, volume flow and mass fraction for the feed stream. Your model should look like this at this step:





4. Now we are ready to add the condenser and reboiler section to the distillation column. Build the condenser and reboiler according to the figure below:



Make specifications to the inventories, valves and boundary streams:

(a) Condenser volume: 50 m^3. The condenser volume should be low at the moment in order to achieve the desired level as soon as possible to avoid huge surges at the beginning of the dynamic simulation. You will need to increase the volume after the dynamic model is activated.

(b) Pump: 10 m design head and 65% efficiency. The design volume flow should be calculated by you.

(c) CV value for the distillate valve and reflux valve: The pressure drop across both valves should be estimated as 50 kPa. Calculate the CV value for these valves.

(d) Mass flow for the condenser vent stream: 0 kg/h. (This simulates a total condenser.)

(e) Pressure of the distillate product: 700 kPa.

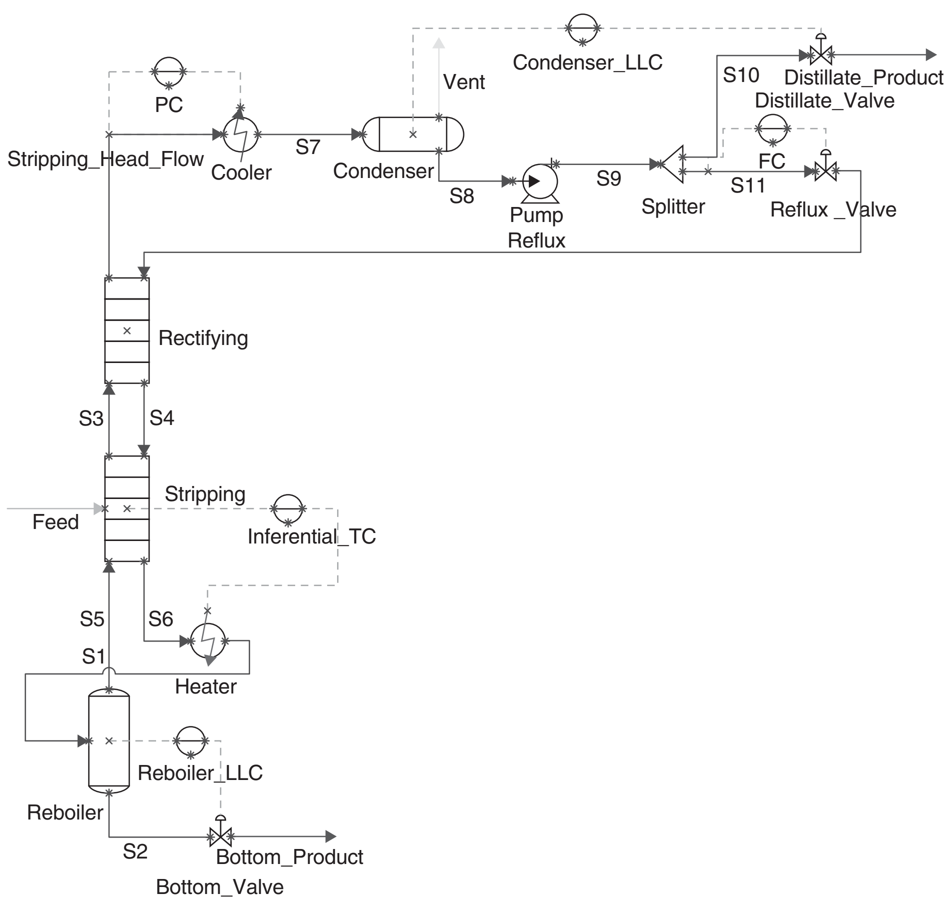
(f) Calculate and specify the CV value of the bottom valve.

(g) Specify the reboiler volume as 50 m^3 and the bottom pressure of 710 kPa. For safety consideration, you also need to set the bottom valve as check valve to avoid possible backflow at the beginning of the simulation.

Save and make a copy of your case since LV and DV control configurations start to be different from here.

5. Apply the LV control configuration:
   (a) The head pressure for the rectifying section is inferentially controlled by the condenser duty. The set point can be obtained from the steady-state model.
   (b) The condenser liquid level is controlled by the distillate flow rate. Add a liquid level controller to achieve average level control of the condenser.
   (c) One of the degrees of freedom in the LV control configuration is in the condenser part, which is the reflux flow rate. Since LV is an energy balance control scheme, inferential control can only be used to control the tray temperature by energy input (V), which leaves L to be set as a constant by the controller. The set point should be obtained from the steady-state model. Tune the flow controller and specify an appropriate CV value for the reflux valve.
   (d) A liquid level controller should control the reboiler inventory by adjusting the bottom flow rate. Average level control scheme should also be applied here.
   (e) An inferential temperature controller is used to set the tray 10 temperature to the desired value, by controlling the reboiler duty. The temperature control should be tight, thus a high gain of 5–10 and a integral time of 5–10 are normally applied.

   Remember to set the controller to automatic and check if the controller should operate in direct or reverse. Your case should look like this at this step:



6. Before running the dynamic integrator, you need to initialize both distillation sections according to the steady-state model. Click 'Init. From Tower' and select the distillation column in the steady-state flow sheet. The rectifying section should start at tray 2 and the stripping section should start at tray 10.
7. Save your case and make a backup, since once the integrator is started, the process cannot be reversed.
8. Start the integrator, wait for a while until a positive flow can be obtained in the bottom product.
9. After a positive flow can be obtained, stop the integrator. Set the volume of condenser and reboiler so that the fluid should have about 10 minutes' residence time.
10. Restart the integrator, wait until the composition in the bottom product stream is steady.

Note:

1. You might need to wait for a long period until the composition is steady, be patient.
2. If dynamic model does not work properly, check
   (a) If the PV and OP range for every controller is suitable.
   (b) If the controller is operating in a right direction.
   (c) If the CV value of the valve is too small so there is a choke flow.

Answer the below questions based on the dynamic model you built:

- Compare the composition of the bottom flow to the steady-state model. Has your primary control objective been achieved in the dynamic model?
- Compare the condenser duty and reboiler duty value with the steady-state model. Give a brief discussion on the dynamic result.
- Add a disturbance to the feed using the selector block. Generate a C_3 fraction disturbance of 0.05 and a total flow rate of 1000 bbl/day. Observe the fraction of C_3 in the bottom flow. Is your dynamic model capable of rectifying disturbances? Why or why not?
- Modify the feed to the minimum flow rate and maximum flow rate as mentioned in task 2. Compare and analyse your dynamic model result to the steady-state model result.
- For DV control configuration, use your saved case from step 5 and add controllers for the DV configuration. Redo all the tests and give your answers to the above questions again for the DV configuration.

Present your findings on CD, DVD or thumb drive in a short report using MS-Word. Also include on the submitted media a copy of the VMGSim files which you used to generate your findings.

Reference

(1) Ryskamp, C.J. (1980) New strategy improves dual composition column control (also effective on thermally coupled columns). *Hydrocarbon Processing*, **June**, 51–59.

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 not only be highly advantageous in the steady state but 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, Third Edition. William Y. Svrcek, Donald P. Mahoney and Brent R. Young.

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 relative gain array (RGA) is a control system design technique that can be used to minimize control loop interaction.
4. Process understanding and a clear understanding of the key process objectives are essential to the development of a good control scheme.
5. 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.
6. There is often a trade-off between steady-state cost savings and dynamic operability.
7. 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.
8. 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 similar 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 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 (DMC) [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.

Table 8.1 *Ratios for steady-state analysis technique.*

Manipulated variables (ratios)	Base case	Minimum flow	Maximum flow
B			
L			
D			
QR			
B/F			
L/D			
QR/*B*			

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% propane in the bottoms and 0.1% isopentane in the distillate. Also assume that you have two perfect analysers (i.e. no dead time, no error) avai1able 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.

- 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 using Table 8.1:
  - For each of the three feed conditions given in Table 7.1 of the previous workshop, record the distillate rate, bottoms rate, reflux rate and reboiler duty when the two composition specifications are satisfied simultaneously.
  - Calculate ratios of the manipulated variables and/or ratios to the feed rate (e.g. *B/F*, *L/D*, QR/*B*)
  - Find the combination of two manipulated variables (ratios) that shows the smallest variability for the whole range of feed conditions. This is done since it will maximize the natural disturbance attenuation of the control system.
  - Outline a candidate 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 may need to use the spreadsheet operation and/or cascade controllers to implement your control scheme.]

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 7.1?

2. Relative Gain Array

The RGA is a tool that can be used to select an appropriate control structure from several candidate structures in a MIMO 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 RGA, Λ, has a property which makes the calculation of all the elements of the array unnecessary. This is shown in Equation 8.1 and applied to a 2×2 system in Equation 8.2. λ_{ij} refers to the relative gain between the ith controlled variable and jth manipulated variable:

$$\sum \lambda_{ij} = \sum \lambda_{ji} = 1, \tag{8.1}$$

$$\Lambda = \begin{bmatrix} \lambda_{11} & \lambda_{12} \\ \lambda_{21} & \lambda_{22} \end{bmatrix} = \begin{bmatrix} \lambda_{11} & 1-\lambda_{11} \\ 1-\lambda_{11} & \lambda_{11} \end{bmatrix}. \tag{8.2}$$

- Determine the λ_{11} element of the RGA 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 steady-state 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 reactor system shown in Figure 8.1. The reactor should be modelled as an equilibrium reactor 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 K) is also given below. The system is very non-ideal, so an activity model should be used, that is, the UNIQUAC model, and the reaction equilibrium should be measured in terms of activities rather than molar concentrations:

$$\begin{aligned} &CH_3OH + CH_2 = C(CH_3)_2 \Leftrightarrow CH_3OC(CH_3)_3, \\ &\text{Methanol} + \text{isobutene} \qquad \Leftrightarrow \text{MTBE}, \\ &\ln\ K_{eq} = -10.54 + 4870/T\ \text{(K)}. \end{aligned}$$

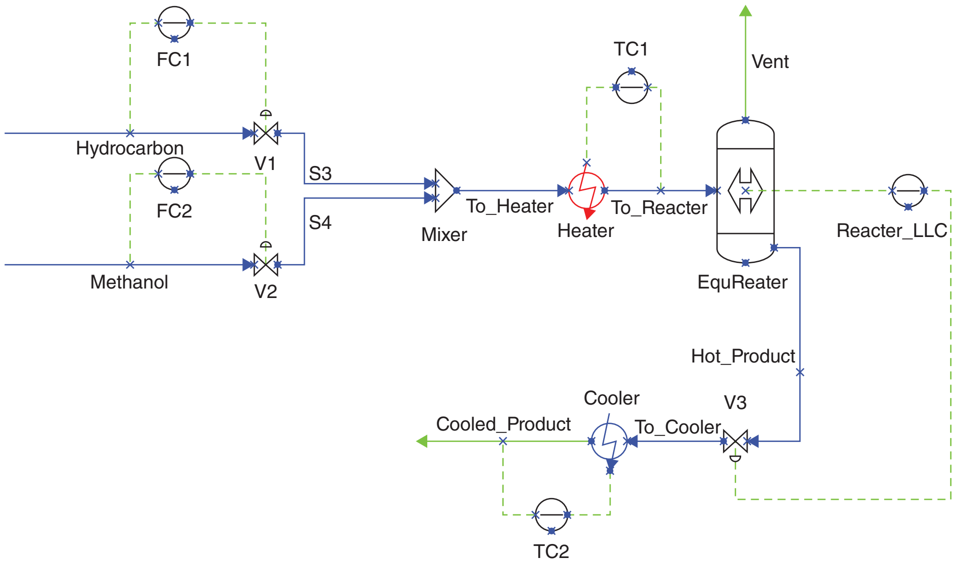


Figure 8.1 *MTBE reaction system without heat integration.*

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% molar excess of the reaction stoichiometry. The hydrocarbon feed rate is 1000 kg/h. Both streams are at 30°C and 3500 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%. The reactor pressure drop is 140 kPa while the pressure drops through the exchangers are 70 kPa. The exchanger volumes can be estimated at 0.1 m^3 each.

Set up the temperature control loops on both heaters and coolers. Then, tune PI controllers for both of these temperature control loops. Carefully specify all the valves and pressure flow characteristics of the system. Also, you will need to add the reaction equation to the reactor and specify the equilibrium constant.

- Test your control system for disturbances in the feed rate, feed temperature and feed composition. The following disturbances are suggested 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 8.2. Again, assume pressure drops of 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.

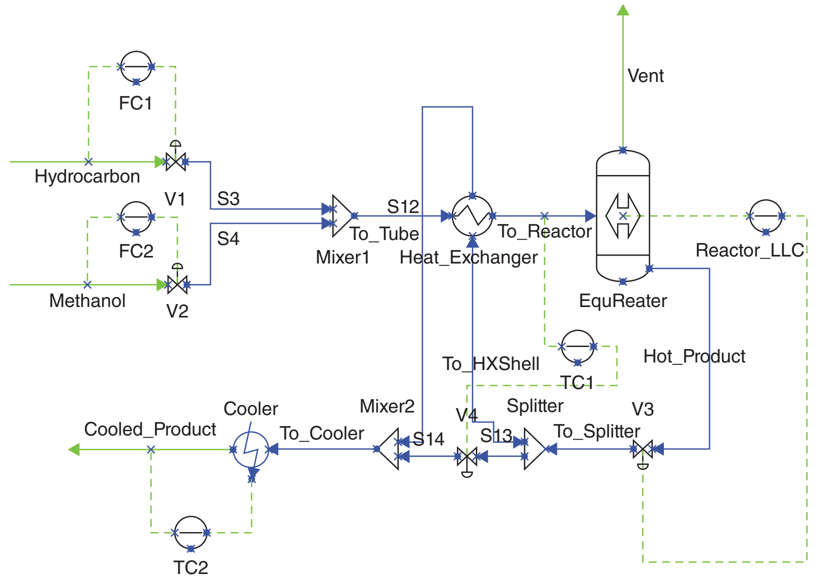


Figure 8.2 MTBE reaction system with heat integration.

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? 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 process design to minimize utility consumption without compromising operability, controllability and safety?

Present your findings on CD, DVD or thumb drive in a short report using MS-Word. Also include on the submitted media a copy of the VMGSim files which you used to generate your findings.

References

(1) Ryskamp, C.J. (1980) New strategy improves dual composition control (also effective on thermally coupled columns). *Hydrocarbon Processing*, **June**, 51–59.
(2) Cutler, C.R. and Ramaker, B.L. (1979) Dynamic Matrix Control – A Computer Control Algorithm. AIChE National Meeting, Houston, 1979; Joint Auto Control Conference, San Francisco.
(3) Prett, D.M. and Gillette, R.D. (1979) Optimization and Constrained Multivariable Control of a Catalytic Cracking Unit. AIChE National Meeting, Houston, 1979; Joint Auto Control Conference, San Francisco.

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