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PART

1

Basics

CHAPTER

1

Introduction

1.1 Overview

Plantwide process control involves the systems and strategies required to control an entire chemical plant consisting of many interconnected unit operations.

One of the most common, important, and challenging control tasks confronting chemical engineers is: How do we design the control loops and systems needed to run our process? We typically are presented with a complicated process flowsheet containing several recycle streams, energy integration, and many different unit operations: distillation columns, reactors of all types, heat exchangers, centrifuges, dryers, crystallizers, liquid-liquid extractors, pumps, compressors, tanks, absorbers, decanters, etc. Given a complex, integrated process and a diverse assortment of equipment, we must devise the necessary logic, instrumentation, and strategies to operate the plant safely and achieve its design objectives.

This is, in essence, the realm of control system synthesis for an entire plant. What issues do we need to consider? What is of essential importance within this immense amount of detail? How does the dynamic behavior of the interconnected plant differ from that of the individual unit operations? What, if anything, do we need to model or test? How do we even begin?

This book addresses each of these questions and explains the fundamental ideas of control system synthesis. As its core, the book presents a general heuristic design procedure that generates an effective plantwide base-level regulatory control structure for an *entire*, *complex* process flowsheet and not simply individual units.

The nine steps of the design procedure center around the fundamental principles of plantwide control: energy management; production rate; product quality; operational, environmental, and safety constraints; liquid level and gas pressure inventories; makeup of reactants; component balances; and economic or process optimization.

We first review in Part 1 the basics of plantwide control. We illustrate its importance by highlighting the unique characteristics that arise when operating and controlling complex integrated processes. The steps of our design procedure are described. In Part 2, we examine how the control of individual unit operations fits within the context of a plantwide perspective. Reactors, heat exchangers, distillation columns, and other unit operations are discussed. Then, the application of the procedure is illustrated in Part 3 with four industrial process examples: the Eastman plantwide control process, the butane isomerization process, the HDA process, and the vinyl acetate monomer process.

1.2 HDA Process

Let's begin with an example of a real industrial process to highlight what we mean by *plantwide process control*. The hydrodealkylation of toluene (HDA) process is used extensively in the book by Douglas (1988) on conceptual design, which presents a hierarchical procedure for generating steady-state flowsheet structures. Hence the HDA process should be familiar to many chemical engineering students who have had a course in process design. It also represents a flowsheet topology that is similar to many chemical plants, so practicing engineers should recognize its essential features.

The HDA process (Fig. 1.1) contains nine basic unit operations: reactor, furnace, vapor-liquid separator, recycle compressor, two heat exchangers, and three distillation columns. Two vapor-phase reactions are considered to generate benzene, methane, and diphenyl from reactants toluene and hydrogen.

$$\text{Toluene} + H_2 \rightarrow \text{benzene} + CH_4 \tag{1.1}$$

$$2\text{Benzene} \rightleftharpoons \text{diphenyl} + H_2 \tag{1.2}$$

The kinetic rate expressions are functions of the partial pressures of toluene p_T, hydrogen p_H, benzene p_B, and diphenyl p_D, with an Arrhenius temperature dependence. By-product diphenyl is produced in an equilibrium reaction.

$$r_1 = k_{1(T)} p_T p_H^{1/2} \tag{1.3}$$

$$r_2 = k_{2f(T)} p_B^2 - k_{2r(T)} p_D p_H \tag{1.4}$$

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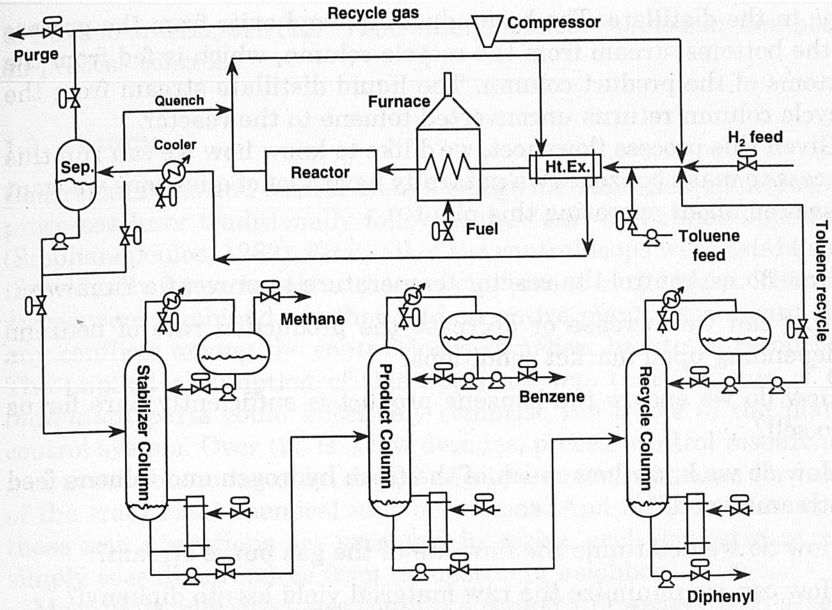

Figure 1.1 HDA process flowsheet.

and one liquid for toluene) are combined with the gas and liquid recycle streams. This combined stream is the cold inlet feed to the process-to-process heat exchanger, where the hot stream is the reactor effluent after the quench. The cold outlet stream is heated further, via combustion of fuel in the furnace, up to the required reactor inlet temperature. The reactor is adiabatic and must be run with an excess of hydrogen to prevent coking. The reactor effluent is quenched with liquid from the separator to prevent fouling in the process-to-process heat exchanger.

The hot outlet stream from the process-to-process heat exchanger goes to a partial condenser and then to a vapor-liquid separator. The gas stream from the overhead of the separator recycles unconverted hydrogen plus methane back to the reactor via a compressor. Since methane enters as an impurity in the hydrogen feed stream and is further produced in the reactor, it will accumulate in the gas recycle loop. Hence a purge stream is required to remove methane from the process. Part of the liquid from the separator serves as the reactor quench stream.

The remainder of the liquid from the separator is fed to the stabilizer column to remove any of the remaining hydrogen and methane gas from the aromatic liquids. The bottoms stream from the stabilizer column feeds the product column, which yields the desired product benzene in the distillate. The by-product diphenyl exits from the process in the bottoms stream from the recycle column, which is fed from the bottoms of the product column. The liquid distillate stream from the recycle column returns unconverted toluene to the reactor.

Given this process flowsheet, we'd like to know how we can run this process to make benzene. We naturally have a lot of questions we want answered about operating this plant:

- How do we control the reactor temperature to prevent a runaway?
- How can we increase or decrease the production rate of benzene depending upon market conditions?
- How do we ensure the benzene product is sufficiently pure for us to sell?
- How do we know how much of the fresh hydrogen and toluene feed streams to add?
- How do we determine the flowrate of the gas purge stream?
- How can we minimize the raw material yield loss to diphenyl?
- How do we prevent overfilling any liquid vessels and overpressuring any units?
- How do we deal with units tied together with heat integration?
- How can we even test any control strategy that we might develop?

Answering these questions is not at all a trivial matter. But these issues lie at the foundation of control system synthesis for an entire plant. The plantwide control problem is extremely complex and very much open-ended. There are a combinatorial number of possible choices and alternative strategies. And there is no unique "correct" solution.

Reaching a solution to the complex plantwide control problem is a creative challenge. It demands insight into and understanding of the chemistry, physics, and economics of real processes. However, it is possible to employ a systematic strategy (or engineering method) to get a feasible solution. Our framework in tackling a problem of this complexity is based upon heuristics that account for the unique features and concerns of integrated plants. This book presents such a general plantwide control design procedure.

The scope embraces continuous processes with reaction and separation sections. Because our approach in this book is based upon a plantwide perspective, we cover what is relevant to this particular area. We omit much basic process control material that constitutes the framework and provides the tools for dynamic analysis, stability, system identification, and controller tuning. But we refer the interested reader

Dynamic controllability

Inventory control

ben and Luyben (1997) and other chemical engineering textbooks
cess control.

History

ol analysis and control system design for chemical and petroleum sses have traditionally followed the "unit operations approach" hanopoulos, 1983). First, all of the control loops were established idually for each unit or piece of equipment in the plant. Then the s were combined together into an entire plant. This meant that conflicts among the control loops somehow had to be reconciled. implicit assumption of this approach was that the sum of the *vidual* parts could effectively comprise the *whole* of the plant's rol system. Over the last few decades, process control researchers practitioners have developed effective control schemes for many ne traditional chemical unit operations. And for processes where e unit operations are arranged in series, each downstream unit ply sees disturbances from its upstream neighbor.

Iost industrial processes contain a complex flowsheet with several ycle streams, energy integration, and many different unit operans. Essentially, the plantwide control problem is how to develop the trol loops needed to operate an *entire* process and achieve its design ectives. Recycle streams and energy integration introduce a feedback material and energy among units upstream and downstream. They o interconnect separate unit operations and create a path for disturnce propagation. The presence of recycle streams profoundly alters e dynamic behavior of the plant by introducing an integrating effect at is not localized to an isolated part of the process.

Despite this process complexity, the unit operations approach to conol system design has worked reasonably well. In the past, plants with cycle streams contained many surge tanks to buffer disturbances, to inimize interaction, and to isolate units in the sequence of material ow. This allowed each unit to be controlled individually. Prior to the 970s, low energy costs meant little economic incentive for energy itegration. However, there is growing pressure to reduce capital inestment, working capital, and operating cost and to respond to safety nd environmental concerns. This has prompted design engineers to tart eliminating many surge tanks, increasing recycle streams, and ntroducing heat integration for both existing and new plants. Often his is done without a complete understanding of their effects on plant operability.

So economic forces within the chemical industry are compelling improved capital productivity. Requirements for on-aim product quality control grow increasingly tighter. More energy integration occurs. Improved product yields, which reduce raw material costs, are achieved via lower reactant per-pass conversion and higher material recycle rates through the process. Better product quality, energy integration, and higher yields are all economically attractive in the steady-state flowsheet, but they present significant challenges to smooth dynamic plant operation. Hence an effective control system regulating the entire plant operation and a process designed with good dynamic performance play critical parts in achieving the business objectives of reducing operating and capital costs.

Buckley (1964) proposed a control design procedure for the plantwide control problem that consisted of two stages. The first stage determined the material balance control structure to handle vessel inventories for low-frequency disturbances. The second established the product quality control structure to regulate high-frequency disturbances. This procedure has been widely and effectively utilized. It has served as the conceptual framework in many subsequent ideas for developing control systems for complete plants. However, the two-stage Buckley procedure provides little guidance concerning three important aspects of a plantwide control strategy. First, it does not explicitly discuss energy management. Second, it does not address the specific issues of recycle systems. Third, it does not deal with component balances in the context of inventory control. By placing the priority on material balance over product quality controls, the procedure can significantly limit the flexibility in choosing the latter.

We believe that chemical process control must move beyond the sphere of unit operations into the realm of viewing the plant as a whole system. The time is ripe in the chemical and petroleum industry for the development of a plantwide control design procedure. The technology, insight, and understanding have reached a state where general guidelines can be presented. The computer software needed for plantwide dynamic simulations is becoming commercially available. While linear methods are very useful to analyze control concepts, we strongly believe that the final evaluation of any plantwide control structure requires rigorous nonlinear dynamic simulations, not linear transfer function analysis.

1.4 Model-Based and Conventional Control

Some people claim that the plantwide control problem has already been solved by the application of several commercial forms of model predictive control (MPC). MPC rests on the idea that we have a fair amount of knowledge about the dynamic behavior of the process and that this knowledge can be incorporated into the *controller* itself. The controller uses past information and current measurements to predict the future response and to adjust its control valves so that this anticipated response is optimal in some sense.

Model predictive control is particularly useful when several control valves (or manipulators) affect an output of interest (what is called *interaction*) and also when some sort of constraint comes into play either on the inputs or on some measured variable. Since the controller itself *knows* about these interactions and constraints, it can in theory avoid those perils. It is important to remember that MPC merely suggests that the controller can predict the process response into the future, only to be checked (and corrected) by the next round of measurements.

On the other hand, conventional control approaches also rely on models, but they are usually not built into the controller itself. Instead the models form the basis of simulations and other analysis methods that guide in the selection of control loops and suggest tuning constants for the relatively simple controllers normally employed [PI, PID, I-only, P-only, lead-lag compensation, etc. (P = proportional, PI = proportional-integral, PID = proportional-integral-derivative)]. Conventional control approaches attempt to build the *smarts* into the *system* (the process and the controllers) rather than only use complex control algorithms.

Our understanding is that MPC has found widespread use in the petroleum industry. The chemical industry, however, is still dominated by the use of distributed control systems implementing simple PID controllers. We are addressing the plantwide control problem within this context. We are not addressing the application of multivariable model-based controllers in this book.

Very few unbiased publications have appeared in the literature comparing control effectiveness using MPC versus a well-designed conventional control system. Most of the MPC applications reported have considered fairly simple processes with a small number of manipulated variables. There are no published reports that discuss the application of MPC to an entire complex chemical plant, with one notable exception. That is the work of Ricker (1996), who compared MPC with conventional PI control for the Eastman process (TE problem). His conclusion was "there appears to be little, if any, advantage to the use of nonlinear model predictive control (NMPC) in this application. In particular, the decentralized strategy does a better job of handling constraints—an area in which NMPC is reputed to excel."

One of the basic reasons for his conclusion ties into the plantwide context that our procedure explicitly addresses, namely the need to regulate all chemical inventories. MPC gives no guidance on how to make the critical decisions of what variables need to be controlled. As Ricker states, "the naive MPC designer might be tempted to control only variables having defined setpoints, relying on optimization to make appropriate use of the remaining degrees of freedom. This fails in the

control structure that does not work

TE problem. As discussed previously, all chemical inventories must be regulated; it cannot be left to chance. Unless setpoints for key internal concentrations are provided, MPC allows reactant partial pressures to drift to unfavorable values." Our design procedure considers the concept of component balances as an explicit step in the design.

Another reason is related to the issue of constraints and priorities, which we address in the sequence of steps for our design procedure. Ricker says that "the TE problem has too many competing goals and special cases to be dealt with in a conventional MPC formulation." Normally this is addressed within MPC by the choice of weights, but for the Eastman process the importance of a variable changes depending upon the situation. "Ricker and Lee found that no single set of weights and constraints could provide the desired performance in all cases."

While we use conventional control systems here, our plantwide control design procedure does not preclude the use of MPC at a certain level. Our focus is on the issues arising from the operation of an integrated process. We find that a good control structure provides effective control, independent of any particular controller algorithm, while a poor one cannot be greatly improved with any algorithm (MPC or PID controllers).

1.5 Process Design

The traditional approach to developing a new process has been to perform the design and control analyses sequentially. First, the design engineer constructs a steady-state process flowsheet, with particular structure, equipment, design parameters, and operating conditions. The objective is to optimize the economics of the project in evaluating the enormous number of alternatives. The hierarchical design procedure proposed by Douglas (1988) is a way to approach this task. Little attention is given to dynamic controllability during the early stages of the design.

After completion of the detailed design, the control engineer then must devise the control strategies to ensure stable dynamic performance and to satisfy the operational requirements. The objective is to operate the plant in the face of potentially known and unknown disturbances, production rate changes, and transitions from one product to another.

While this staged approach has long been recognized as deficient, it is defensible from a certain perspective. For example, it would be difficult for the control engineers to specify the instrumentation and the distributed control system (DCS) without knowing exactly what process it was intended for. Similarly, it would make no sense for the process engineers to request a control system design for all those flowsheets that were considered but rejected on the basis of steady-state economics alone. However, this staged approach can result in missed opportunities because of the close connection between process design and controllability. How a process is designed fundamentally determines its inherent controllability, which means qualitatively how well the process rejects disturbances and how easily it moves from one operating condition to another. In an ideal project system, dynamics and control strategies would be considered during the process synthesis and design activities.

This issue grows increasingly important as plants become more highly integrated with complex configurations, recycle streams, and energy integration. Competitive economic pressures, safety issues, and environmental concerns have all contributed to this. However, if a control engineer becomes involved early enough in the process design, he or she may be able to show that it would be better in the long run to build a process with higher capital and utility costs if that plant provides more stable operation and less variability in the product quality.

We believe that process design impacts controllability far more than control algorithms do. We base our opinion on many years of experience. We have participated as control engineers in many design projects. Some involved building new plants with new process technology, some involved new plants with existing technology, and some projects were modernizations of the control system on an existing plant. We have found that a consideration of dynamics and control strategies for new process designs has a much larger positive economic impact (when the design can potentially be modified) compared with control strategy upgrades on an existing process (with a fixed design). However, we stress that for those new plants and technologies we became involved before the process design was fixed. We performed dynamic simulations and undertook control system design as soon as the process engineers had an economically viable flowsheet. Most importantly, by working together with the process engineers and plant engineers, we changed the flowsheet until we were all satisfied that we had developed the most profitable process when viewed over the entire life time of the project. This inevitably involved making trade-offs between steady-state investment economics and dynamic performance measured in uptime, throughput, product quality, and yield.

One of the important themes weaving through this book is the central role we place on the process design. Good control engineers need also to be good process engineers!

1.6 Spectrum of Process Control

We can view the field of process control as five parts of a continuous spectrum (Fig. 1.2). Each part is important, can be economically signifi-

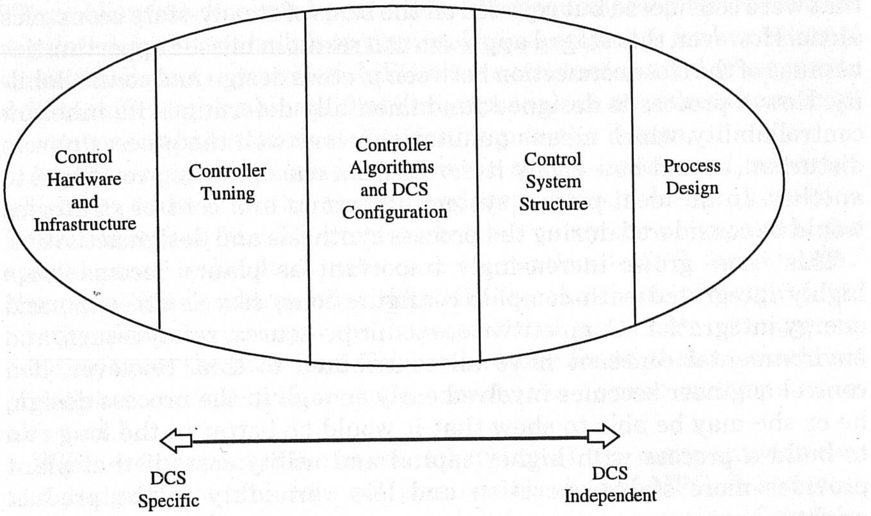

Figure 1.2 Spectrum of process control.

cant, and interacts in some manner with the others. Moving toward the left on the spectrum means dealing with more detailed issues on the level of the distributed control system (DCS). Moving toward the right means operating on a more general level with issues that are independent of the DCS.

The far left part of the spectrum deals with the control hardware and infrastructure required to operate a plant. We need to assemble the proper types of control valves and process measurements (for temperature, flow, pressure, composition, etc.). These are the sensory devices of the plant and are essential for any control system to function. Any control strategy, no matter how clever, will have severe difficulties without the right measurements and valves in the process. An Instrument Society of America (ISA) publication catalog (67 Alexander Drive, P.O. Box 12277, Research Triangle Park, NC 27709) contains many references that deal with control hardware.

The next part involves controller tuning. We must determine the tuning constants for the controllers in the plant. While this task is often performed by using heuristics and experience, it can sometimes be a nontrivial exercise for certain loops. We recommend using a relay-feedback test that determines the ultimate gain and period for the control loop, from which controller settings can be calculated (Luyben and Luyben, 1997).

The middle of the spectrum deals with the controller algorithms and DCS configuration. We must decide the type of controller to use (proportional, integral, derivative, multivariable, nonlinear, model predictive, etc.). We must also determine whether we need dynamic elements (lead/lags, feedforward, etc.) and how to handle overrides and interlocks. In addition, input and output variables must be assigned loop numbers, displays must be created, alarms must be specified, instrument groupings must be determined, etc.

The next part is the determination of the control system structure. We must decide what variables to control and manipulate and how these should be paired. The control structure is vitally important because a poor strategy will result in poor performance no matter what type of control algorithm we use or how much we tune it. There is little information or guidance in the literature or in process control textbooks (both introductory and advanced) on how to develop an effective control structure for an entire complex chemical plant. This is the main subject of this book.

The far right part of the spectrum is the design of the process itself. We sometimes can change the flowsheet structure, use different design parameters, and employ different types of process equipment to produce a plant that can be controlled more easily than other alternatives. At this level, a good process control engineer can potentially have an enormous economic impact. Most companies in the chemical and petroleum industries have had the unfortunate and unwelcome experience of building a plant that could not easily be started up because of operational difficulties arising from the plant design. Fixing these kinds of problems after the plant is built can often require large amounts of additional capital expense in addition to the lost sales opportunities.

In this book, we focus primarily on control structure selection. Interactions between design and control are illustrated by examples, and the effects of design parameters on control are discussed. However, we do not present a synthesis procedure for process design that is capable of generating the most controllable flowsheet for a given chemistry. This is still very much an open area for further research.

1.7 Conclusion

In this first chapter we have defined the plantwide process control problem. This was illustrated by using the HDA process, which will figure prominently in later parts of the book. We have provided a historical perspective and context. Finally we explained where the material in this book fits into the spectrum of process control activities.

1.8 References

Buckley, P. S. *Techniques of Process Control*, New York: Wiley (1964).
Douglas, J. M. *Conceptual Design of Chemical Processes*, New York: McGraw-Hill (1988).

Luyben, W. L., and Luyben M. L. *Essentials of Process Control*, New York: McGraw-Hill (1997).
Ricker, N. L. "Decentralized Control of the Tennessee Eastman Challenge Process," *J. Proc. Cont.*, **6,** 205–221 (1996).
Stephanopoulos, G. "Synthesis of Control Systems for Chemical Plants—A Challenge for Creativity," *Comput. Chem. Eng.*, **7,** 331–365 (1983).