CHAPTER

3

Plantwide Control Design Procedure

3.1 Introduction

In an industrial environment, a plant's control strategy should be simple enough, at least conceptually, so that everyone from the operator to the plant manager can understand how it works. Our governing philosophy is *it is always best to utilize the simplest control system that will achieve the desired objectives.* The more complex the process, the more desirable it is to have a simple control strategy. This view differs radically from much of the current academic thinking about process control, which suggests that a complex process demands complex control. Our viewpoint is a result of many years of working on practical plant control problems, where it is important to be able to identify whether an operating problem has its source in the process or in the control system.

The goals for an effective plantwide process control system include (1) safe and smooth process operation; (2) tight control of product quality in the face of disturbances; (3) avoidance of unsafe process conditions; (4) a control system run in automatic, not manual, requiring minimal operator attention; (5) rapid rate and product quality transitions; and (6) zero unexpected environmental releases.

As illustrated in the previous chapter, the need for a plantwide control perspective arises from three important features of integrated processes: the effects of material recycle, of chemical component inventories, and of energy integration. We have shown several control strategies that highlight important general issues. However, we did not describe how we arrived at these strategies, and many of our choices may seem mysterious at this point. Why, for instance, did we choose to use fresh liquid reactant feed streams in the control of liquid inventories? What prompted us to have a reactor composition analyzer? Why were we concerned with a single direct handle to set production rate?

In this chapter we outline the nine basic steps of a general heuristic plantwide control design procedure (Luyben et al., 1997). After some preliminary discussion of the fundamentals on which this procedure is based, we outline each step in general terms. We also summarize our justification for the sequence of steps. The method is illustrated in applications to four industrial process examples in Part 3.

The procedure essentially decomposes the plantwide control problem into various levels. It forces us to focus on the unique features and issues associated with a control strategy for an entire plant. We highlighted some of these questions in Chap. 1 in discussing the HDA process. How do we manage energy? How is production rate controlled? How do we control product quality? How do we determine the amounts of fresh reactants to add?

Our plantwide control design procedure (Fig. 3.1) satisfies the two fundamental chemical engineering principles, namely the overall conservation of energy and mass. Additionally, the procedure accounts for nonconserved entities within a plant such as chemical components (produced and consumed) and entropy (produced). In fact, five of the nine steps deal with plantwide control issues that would not be addressed by simply combining the control systems from all of the individual unit operations.

Steps 1 and 2 establish the objectives of the control system and the available degrees of freedom. Step 3 ensures that any production of heat (entropy) within the process is properly dissipated and that the propagation of thermal disturbances is prevented. In Steps 4 and 5 we

1. Establish Control Objectives
2. Determine Control Degrees of Freedom
3. Establish Energy Management System
4. Set Production Rate OK
5. Control Product Quality and Handle Safety, Environmental, and Operational Constraints
6. Fix a Flow in Every Recycle Loop and Control Inventories (Pressures and Liquid Levels)
7. Check Component Balances
8. Control Individual Unit Operations
9. Optimize Economics and Improve Dynamic Controllability

Figure 3.1 Nine steps of plantwide control design procedure.

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satisfy the key business objectives concerning production rate, product quality, and safety. Step 6 involves total mass balance control, whereas in Step 7 we ensure that nonconserved chemical components are accounted for. That concludes the plantwide control aspects. In Step 8 we complete the control systems for individual unit operations. Finally, Step 9 uses the remaining degrees of freedom for optimization and improved dynamic controllability. This heuristic procedure will generate a workable plantwide control strategy, which is not necessarily the *best* solution. Because the design problem is open-ended, the procedure will not produce a unique solution.

The plantwide control design procedure presented here was developed after many years of work and research in the fields of process control and process design. Research efforts by a number of people in industry and at universities have contributed essential ideas and concepts. We have assembled, analyzed, and processed this prior work to reach a logical, coherent, step-by-step procedure. We want to acknowledge these previous contributions and state that we are indeed fortunate to stand upon the shoulders of many giants. Listed below are some of the fundamental concepts and techniques that form the basis of the procedure.

3.2 Basic Concepts of Plantwide Control

3.2.1 Buckley basics

Page Buckley (1964), a true pioneer with DuPont in the field of process control, was the first to suggest the idea of separating the plantwide control problem into two parts: material balance control and product quality control. He suggested looking first at the flow of material through the system. A logical arrangement of level and pressure control loops is established, using the flowrates of the liquid and gas process streams. No controller tuning or inventory sizing is done at this step. The idea is to establish the inventory control system by setting up this "hydraulic" control structure as the first step.

He then proposed establishing the product-quality control loops by choosing appropriate manipulated variables. The time constants of the closed-loop product-quality loops are estimated. We try to make these as small as possible so that good, tight control is achieved, but stability constraints impose limitations on the achieveable performance.

Then the inventory loops are revisited. The liquid holdups in surge volumes are calculated so that the time constants of the liquid level loops (using proportional-only controllers) are a factor of 10 larger than the product-quality time constants. This separation in time constants permits independent tuning of the material-balance loops and the product-quality loops. Note that most level controllers should be proportional-only (P) to achieve flow smoothing.

3.2.2 Douglas doctrines

Jim Douglas (1988) of the University of Massachusetts has devised a hierarchical approach to the conceptual design of process flowsheets. Although he primarily considers the steady-state aspects of process design, he has developed several useful concepts that have control structure implications.

Douglas points out that in the typical chemical plant the costs of raw materials and the value of the products are usually much greater than the costs of capital and energy. This leads to the two *Douglas doctrines:*

1. Minimize losses of reactants and products.
2. Maximize flowrates through gas recycle systems.

The first idea implies that we need tight control of stream compositions exiting the process to avoid losses of reactants and products. The second rests on the principle that yield is worth more than energy. Recycles are used to improve yields in many processes, as was discussed in Chap. 2. The economics of improving yields (obtaining more desired products from the same raw materials) usually outweigh the additional energy cost of driving the recycle gas compressor.

The control structure implication is that we do not attempt to regulate the gas recycle flow and we do not worry about what we control with its manipulation. We simply maximize its flow. This removes one control degree of freedom and simplifies the control problem.

3.2.3 Downs drill

Jim Downs (1992) of Eastman Chemical Company has insightfully pointed out the importance of looking at the chemical component balances around the entire plant and checking to see that the control structure handles these component balances effectively. The concepts of overall component balances go back to our first course in chemical engineering, where we learned how to apply mass and energy balances to any system, microscopic or macroscopic. We did these balances for individual unit operations, for sections of a plant, and for entire processes.

But somehow these basics are often forgotten or overlooked in the complex and intricate project required to develop a steady-state design for a large chemical plant and specify its control structure. Often the design job is broken up into pieces. One person will design the reactor and its control system and someone else will design the separation

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section and its control system. The task sometimes falls through the cracks to ensure that these two sections operate effectively when coupled together. Thus it is important that we perform the *Downs drill.*

We must ensure that all components (reactants, products, and inerts) have a way to leave or be consumed within the process. The consideration of inerts is seldom overlooked. Heavy inerts can leave the system in the bottoms product from a distillation column. Light inerts can be purged from a gas recycle stream or from a partial condenser on a column. Intermediate inerts must also be removed in some way, for example in sidestream purges or separate distillation columns.

Most of the problems occur in the consideration of reactants, particularly when several chemical species are involved. All of the reactants fed into the system must either be consumed via reaction or leave the plant as impurities in the exiting streams. Since we usually want to minimize raw material costs and maintain high-purity products, most of the reactants fed into the process must be chewed up in the reactions. And the stoichiometry must be satisfied *down to the last molecule.*

Chemical plants often act as pure integrators in terms of reactants. This is due to the fact that we prevent reactants from leaving the process through composition controls in the separation section. Any imbalance in the number of moles of reactants involved in the reactions, no matter how slight, will result in the process gradually filling up with the reactant component that is in excess. The ternary system considered in Chap. 2 illustrated this effect. There must be a way to adjust the fresh feed flowrates so that exactly the right amounts of the two reactants are fed in.

"LAV" FOR LOW THROUGHPUT (UNCLASS.) CASE

3.2.4 Luyben laws

Three laws have been developed as a result of a number of case s of many types of systems:

1. A stream somewhere in all recycle loops should be flow controlled. This is to prevent the snowball effect and was discussed in Chap. 2. 22

2. A fresh reactant feed stream cannot be flow-controlled unless there is essentially complete one-pass conversion of one of the reactants. This law applies to systems with reaction types such as $A + B \rightarrow$ products and was discussed in Chap. 2. In systems with consecutive reactions such as $A + B \rightarrow M + C$ and $M + B \rightarrow D + C$, the fresh feeds can be flow-controlled into the system because any imbalance in the ratios of reactants is accommodated by a shift in the amounts of the two products (M and D) that are generated. An excess of A will result in the production of more M and less D. An excess of B results in the production of more D and less M.

? You mean all fresh reactant (to avoid imbalance)

3. If the final product from a process comes out the top of a distillation column, the column feed should be liquid. If the final product comes out the bottom of a column, the feed to the column should be vapor (Cantrell et al., 1995). Changes in feed flowrate or feed composition have less of a dynamic effect on distillate composition than they do on bottoms composition if the feed is saturated liquid. The reverse is true if the feed is saturated vapor: bottoms is less affected than distillate. If our primary goal is to achieve tight product quality control, the basic column design should consider the dynamic implications of feed thermal conditions. Even if steady-state economics favor a liquid feed stream, the profitability of an operating plant with a product leaving the bottom of a column may be much better if the feed to the column is vaporized. This is another example of the potential conflict between steady-state economic design and dynamic controllability.

Interesting, but can also do well with right control

3.2.5 Richardson rule

Bob Richardson of Union Carbide suggested the heuristic that the largest stream should be selected to control the liquid level in a vessel. This makes good sense because it provides more *muscle* to achieve the desired control objective. An analogy is that it is much easier to maneuver a large barge with a tugboat than with a life raft. We often use the expression that you can't make a garbage truck drive like a Ferrari. But this is not necessarily true. If you put a 2000-hp engine in the garbage truck (and redesigned the center of gravity), you could make it handle just like a sports car. The point is that the bigger the handle you have to affect a process, the better you can control it. This is why there are often fundamental conflicts between steady-state design and dynamic controllability.

3.2.6 Shinskey schemes

Greg Shinskey (1988), over the course of a long and productive career at Foxboro, has proposed a number of "advanced control" structures that permit improvements in dynamic performance. These schemes are not only effective, but they are simple to implement in basic control instrumentation. Liberal use should be made of ratio control, cascade control, override control, and valve-position (optimizing) control. These strategies are covered in most basic process control textbooks.

3.2.7 Tyreus tuning

One of the vital steps in developing a plantwide control system, once both the process and the control structure have been specified, is to

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determine the algorithm to be used for each controller (P, PI, or PID) and to tune each controller. We strongly recommend the use of P-only controllers for liquid levels (even in some liquid reactor applications). Tuning of a P controller is usually trivial: set the controller gain equal to 1.67. This will have the valve wide open when the level is at 80 percent and the valve shut when the level is at 20 percent (assuming the stream flowing out of the vessel is manipulated to control liquid level; if the level is controlled by the inflowing stream the action of the controller is reverse instead of direct).

For other control loops, we suggest the use of PI controllers. The relay-feedback test is a simple and fast way to obtain the ultimate gain (K_u) and ultimate period (P_u). Then either the Ziegler-Nichols settings (for very tight control with a closed-loop damping coefficient of about 0.1) or the Tyreus-Luyben (1992) settings (for more conservative loops where a closed-loop damping coefficient of 0.4 is more appropriate) can be used:

$$K_{ZN} = K_u/2.2 \qquad \tau_{ZN} = P_u/1.2$$

$$K_{TL} = K_u/3.2 \qquad \tau_{TL} = 2.2P_u$$

The use of PID controllers should be restricted to those loops where two criteria are both satisfied: the controlled variable should have a very large signal-to-noise ratio and tight dynamic control is really essential from a feedback control stability perspective. The classical example of the latter is temperature control in an irreversible exothermic chemical reactor (see Chap. 4).

3.3 Steps of Plantwide Process Control Design Procedure

In this section we discuss each step of the design procedure in detail.

Step 1: Establish control objectives

Assess the steady-state design and dynamic control objectives for the process.

This is probably the most important aspect of the problem because different control objectives lead to different control structures. There is an old Persian saying "If you don't know where you are going, any road will get you there!" This is certainly true in plantwide control. The "best" control structure for a plant depends upon the design and control criteria established.

These objectives include reactor and separation yields, product qual-

But where do these come from?
Economics!

ity specifications, product grades and demand determination, environmental restrictions, and the range of safe operating conditions.

Step 2: Determine control degrees of freedom

Count the number of control valves available.

This is the number of degrees of freedom for control, i.e., the number of variables that can be controlled to setpoint. The valves must be legitimate (flow through a liquid-filled line can be regulated by only one control valve). The placement of these control valves can sometimes be made to improve dynamic performance, but often there is no choice in their location.

Most of these valves will be used to achieve basic regulatory control of the process: (1) set production rate, (2) maintain gas and liquid inventories, (3) control product qualities, and (4) avoid safety and environmental constraints. Any valves that remain after these vital tasks have been accomplished can be utilized to enhance steady-state economic objectives or dynamic controllability (e.g., minimize energy consumption, maximize yield, or reject disturbances).

During the course of the subsequent steps, we may find that we lack suitable manipulators to achieve the desired economic control objectives. Then we must change the process design to obtain additional handles. For example, we may need to add bypass lines around heat exchangers and include auxiliary heat exchangers.

Step 3: Establish energy management system

← What is this ?!
Removing heat from exothermic reaction

Make sure that energy disturbances do not propagate throughout the process by transferring the variability to the plant utility system.

We use the term *energy management* to describe two functions: (1) We must provide a control system that removes exothermic heats of reaction from the process. If heat is not removed to utilities directly at the reactor, then it can be used elsewhere in the process by other unit operations. This heat, however, must ultimately be dissipated to utilities. (2) If heat integration does occur between process streams, then the second function of energy management is to provide a control system that prevents the propagation of thermal disturbances and ensures the exothermic reactor heat is dissipated and not recycled. Process-to-process heat exchangers and heat-integrated unit operations must be analyzed to determine that there are sufficient degrees of freedom for control.

Heat removal in exothermic reactors is crucial because of the potential for thermal runaways. In endothermic reactions, failure to add

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enough heat simply results in the reaction slowing up. If the exothermic reactor is running adiabatically, the control system must prevent excessive temperature rise through the reactor (e.g., by setting the ratio of the flowrate of the limiting fresh reactant to the flowrate of a recycle stream acting as a thermal sink). More details of reactor control are discussed in Chap. 4.

Heat transfer between process streams can create significant interaction. In the case of reactor feed/effluent heat exchangers it can lead to positive feedback and even instability. Where there is partial condensation or partial vaporization in a process-to-process heat exchanger, disturbances can be amplified because of heat of vaporization and temperature effects.

For example, suppose the temperature of a stream being fed to a distillation column is controlled by manipulating steam flowrate to a feed preheater. And suppose the stream leaving the preheater is partially vaporized. Small changes in composition can result in very large changes in the fraction of the stream that is vaporized (for the same pressure and temperature). The resulting variations in the liquid and vapor rates in the distillation column can produce severe upsets.

Heat integration of a distillation column with other columns or with reactors is widely used in chemical plants to reduce energy consumption. While these designs look great in terms of steady-state economics, they can lead to complex dynamic behavior and poor performance due to recycling of disturbances. If not already included in the design, trim heaters/coolers or heat exchanger bypass lines must be added to prevent this. Energy disturbances should be transferred to the plant utility system whenever possible to remove this source of variability from the process units. Chapter 5 deals with heat exchanger systems.

What if the process has no reactor?

Step 4: Set production rate

Establish the variables that dominate the productivity of the reactor and determine the most appropriate manipulator to control production rate.

Throughput changes can be achieved only by altering, either directly or indirectly, conditions in the reactor. To obtain higher production rates, we must increase overall reaction rates. This can be accomplished by raising temperature (higher specific reaction rate), increasing reactant concentrations, increasing reactor holdup (in liquid-phase reactors), or increasing reactor pressure (in gas-phase reactors).

Our first choice for setting production rate should be to alter one of these variables in the reactor. The variable we select must be dominant ? for the reactor. Dominant reactor variables always have significant effects on reactor performance. For example, temperature is often a

what if we have a separation process without a reactor?

dominant reactor variable. In irreversible reactions, specific rates increase exponentially with temperature. As long as reaction rates are not limited by low reactant concentrations, we can *increase* temperature to increase production rate in the plant. In reversible exothermic reactions, where the equilibrium constant decreases with increasing temperature, reactor temperature may still be a dominant variable. If the reactor is large enough to reach chemical equilibrium at the exit, we can *decrease* reactor temperature to increase production.

There are situations where reactor temperature is not a dominant variable or cannot be changed for safety or yield reasons. In these cases, we must find another dominant variable, such as the concentration of the limiting reactant, flowrate of initiator or catalyst to the reactor, reactor residence time, reactor pressure, or agitation rate.

Once we identify the dominant variables, we must also identify the manipulators (control valves) that are most suitable to control them. The manipulators are used in feedback control loops to hold the dominant variables at setpoint. The setpoints are then adjusted to achieve the desired production rate, in addition to satisfying other economic control objectives.

Whatever variable we choose, we would like it to provide smooth and stable production rate transitions and to reject disturbances. We often want to select a variable that has the least effect on the separation section but also has a rapid and direct effect on reaction rate in the reactor without hitting an operational constraint.

When the setpoint of a dominant variable is used to establish plant production rate, the control strategy must ensure that the right amounts of fresh reactants are brought into the process. This is often accomplished through fresh reactant makeup control based upon liquid levels or gas pressures that reflect component inventories. We must keep these ideas in mind when we reach Steps 6 and 7.

However, design constraints may limit our ability to exercise this strategy concerning fresh reactant makeup. An upstream process may establish the reactant feed flow sent to the plant. A downstream process may require on-demand production, which fixes the product flowrate from the plant. In these cases, the development of the control strategy becomes more complex because we must somehow adjust the setpoint of the dominant variable on the basis of the production rate that has been specified externally. We must balance production rate with what has been specified externally. This cannot be done in an open-loop sense. Feedback of information about actual internal plant conditions is required to determine the accumulation or depletion of the reactant components. This concept was nicely illustrated by the control strategy in Fig. 2.16. In that scheme we fixed externally the flow of fresh reactant *A* feed. Also, we used reactor residence time (via the effluent flowrate)

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as the controlled dominant variable. Feedback information (internal reactant composition information) is provided to this controller by the ratio of the two recycle stream flows.

Step 5: Control product quality and handle safety, operational, and environmental constraints

Select the "best" valves to control each of the product-quality, safety, and environmental variables.

We want tight control of these important quantities for economic and operational reasons. Hence we should select manipulated variables such that the dynamic relationships between the controlled and manipulated variables feature small time constants and deadtimes and large steady-state gains. The former gives small closed-loop time constants and the latter prevents problems with the rangeability of the manipulated variable (control valve saturation). Good point!

It should be noted that establishing the product-quality loops first, before the material balance control structure, is a fundamental difference between our plantwide control design procedure and Buckley's procedure. Since product quality considerations have become more important in recent years, this shift in emphasis follows naturally.

The magnitudes of various flowrates also come into consideration. For example, temperature (or bottoms product purity) in a distillation column is typically controlled by manipulating steam flow to the reboiler (column boilup) and base level is controlled with bottoms product flowrate. However, in columns with a large boilup ratio and small bottoms flowrate, these loops should be reversed because boilup has a larger effect on base level than bottoms flow (Richardson rule). However, inverse response problems in some columns may occur when base level is controlled by heat input. High reflux ratios at the top of a column require similar analysis in selecting reflux or distillate to control overhead product purity.

Step 6: Fix a flow in every recycle loop and control inventories (pressures and levels)

Fix a flow in every recycle loop and then select the best manipulated variables to control inventories.

In most processes a flow controller should be present in all liquid recycle loops. This is a simple and effective way to prevent potentially large changes in recycle flows that can occur if all flows in the recycle loop are controlled by levels, as illustrated by the simple process examples in Chap. 2. Steady-state and dynamic benefits result from this flow control strategy. From a steady-state viewpoint, the plant's separation ok section is not forced to operate at significantly different load conditions, which could lead to turndown or flooding problems.

From a dynamic viewpoint, whenever all flows in a recycle loop are set by level controllers, wide dynamic excursions can occur in these flows because the total system inventory is not regulated. The control system is attempting to control the inventory in each individual vessel by changing the flowrate to its downstream neighbor. In a recycle loop, all level controllers see load disturbances coming from the upstream unit. This causes the flowrate disturbances to propagate around the recycle loop. Thus any disturbance that tends to increase the total inventory in the process (such as an increase in the fresh feed flowrate) will produce large increases in all flowrates around the recycle loop.

Fixing a flowrate in a recycle stream does not conflict with our discussion of picking a dominant reactor variable for production rate control in Step 4. Flow controlling a stream somewhere in all recycle loops is an important simple part of any plantwide control strategy.

Gas recycle loops are normally set at maximum circulation rate, as limited by compressor capacity, to achieve maximum yields (Douglas doctrine).

Once we have fixed a flow in each recycle loop, we then determine what valve should be used to control each inventory variable. This is the material balance step in the Buckley procedure. Inventories include all liquid levels (except for surge volume in certain liquid recycle streams) and gas pressures. An inventory variable should typically be controlled with the manipulated variable that has the largest effect on it within that unit (Richardson rule). Because we have fixed a flow in each recycle loop, our choice of available valves has been reduced for inventory control in some units. Sometimes this actually eliminates the obvious choice for inventory control for that unit. This constraint forces us to look outside the immediate vicinity of the holdup we are considering.

For example, suppose that the distillate flowrate from a distillation column is large compared to the reflux. We normally would use distillate to control level in the reflux drum. But suppose the distillate recycles back to the reactor and so we want to control its flow. What manipulator should we use to control reflux drum level? We could potentially use condenser cooling rate or reboiler heat input. Either choice would have implications on the control strategy for the column, which would ripple through the control strategy for the rest of the plant. This would lead to control schemes that would never be considered if one looked only at the unit operations in isolation.

Inventory may also be controlled with fresh reactant makeup streams as discussed in Step 4. Liquid fresh feed streams may be added to a location where level reflects the amount of that component in the pro-

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cess. Gas fresh feed streams may be added to a location where pressure reflects the amount of that material in the process.

Proportional-only control should be used in nonreactive level loops for cascaded units in series. Even in reactor level control, proportional control should be considered to help filter flowrate disturbances to the downstream separation system. There is nothing necessarily sacred about holding reactor level constant.

Step 7: Check component balances

Identify how chemical components enter, leave, and are generated or consumed in the process.

Component balances can often be quite subtle, but they are particularly important in processes with recycle streams because of their integrating effect. They depend upon the specific kinetics and reaction paths in the system. They often affect what variable can be used to set production rate or reaction rate in the reactor. The buildup of chemical components in recycle streams must be prevented by keeping track of chemical component inventories (reactants, products, and inerts) inside the system.

We must identify the specific mechanism or control loop to guarantee that there will be no uncontrollable buildup of any chemical component within the process (Downs drill).

What are the methods or loops to ensure that the overall component balances for all chemical species are satisfied at steady state? We can limit their intake, control their reaction, or adjust their outflow from the process.

As we noted in Chap. 2, we can characterize a plant's chemical components into reactants, products, and inerts. We don't want reactant components to leave in the product streams because of the yield loss and the desired product purity specifications. Hence we are limited to the use of two methods: consuming the reactants by reaction or adjusting their fresh feed flow. Product and inert components all must have an exit path from the system. In many systems inerts are removed by purging off a small fraction of the recycle stream. The purge rate is adjusted to control the inert composition in the recycle stream so that an economic balance is maintained between capital and operating costs.

We recommend making a Downs drill table that lists each chemical component, its input, its generation or consumption, and its output. This table should specify how the control system will detect an imbalance in chemical components and what specific action it will take if an imbalance is detected.

Should do for gas and liquid phase also

Step 8: Control individual unit operations

Establish the control loops necessary to operate each of the individual unit operations.

Many effective control schemes have been established over the years for individual chemical units (Shinskey, 1988). For example, a tubular reactor usually requires control of inlet temperature. High-temperature endothermic reactions typically have a control system to adjust the fuel flowrate to a furnace supplying energy to the reactor. Crystallizers require manipulation of refrigeration load to control temperature. Oxygen concentration in the stack gas from a furnace is controlled to prevent excess fuel usage. Liquid solvent feed flow to an absorber is controlled as some ratio to the gas feed. We deal with the control of various unit operations in Chaps. 4 through 7.

Step 9: Optimize economics or improve dynamic controllability

Establish the best way to use the remaining control degrees of freedom.

No, you should have thought of this at step 1.

After satisfying all of the basic regulatory requirements, we usually have additional degrees of freedom involving control valves that have not been used and setpoints in some controllers that can be adjusted. These can be utilized either to optimize steady-state economic process performance (e.g., minimize energy, maximize selectivity) or to improve dynamic response.

For example, suppose an exothermic chemical reactor may be cooled with both jacket cooling water and brine (refrigeration) to a reflux condenser. For fast reactor temperature control, manipulation of brine is significantly better than cooling water. However, the utility cost of brine is much higher than cooling water. Hence we would like the control system to provide tight reactor temperature control while minimizing brine usage. This can be achieved with a valve position control strategy. Reactor temperature is controlled by manipulating brine. A valve position controller looks at the position of the brine control valve and slowly adjusts jacket cooling water flow to keep the brine valve approximately 10 to 20 percent open under steady-state operation (Fig. 3.2).

Additional considerations

Certain quantitative measures from linear control theory may help at various steps to assess relationships between the controlled and manipulated variables. These include steady-state process gains, open-loop time constants, singular value decomposition, condition numbers, eigenvalue analysis for stability, etc. These techniques are described in

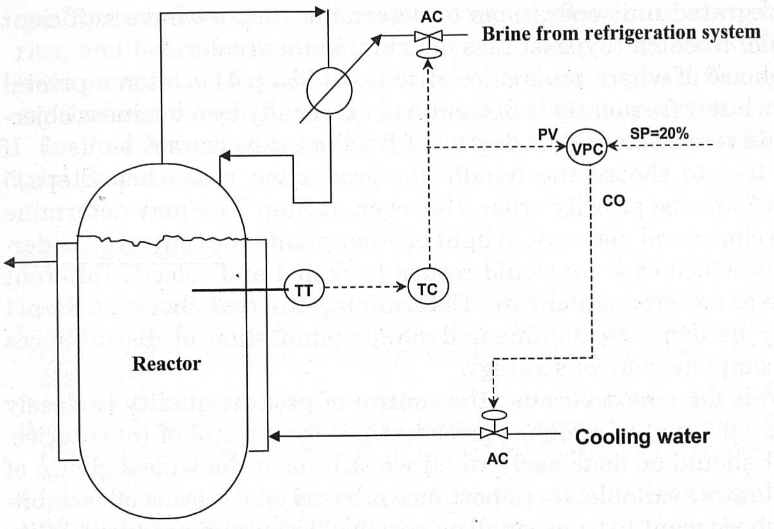

Figure 3.2 Illustration of valve position control strategy.

detail in most process control textbooks. The plantwide control strategy should ultimately be tested on a nonlinear dynamic model that captures the essential process behavior.

Since the design of a chemical process profoundly affects its dynamic controllability, another part of the problem's open-ended nature is the opportunity to change the *process* design. The design-and-control interaction problem remains as yet an open research area in terms of the plantwide control problem.

3.4 Justification of Sequence

Although the order of the steps in the design procedure may initially seem arbitrary, the sequence comes from a consideration first of choices that have already been assigned due to equipment or business constraints and then the importance in a hierarchy of priorities. Steps 1 and 2 are straightforward in determining the objectives and available degrees of freedom.

Step 3 is next because the reactor is typically the heart of an industrial process and the methods for heat removal are intrinsically part of the reactor design. So it is usually not optional what degrees of freedom can be used for exothermic reactor control. When the heat generated in an exothermic reactor is used within the process via energy integration, we must ensure that the energy is dissipated and not recycled. Hence we examine process-to-process heat exchangers and

But if it is for stabilizatn then setpoint remains DOF, so we can't violate this!

heat-integrated unit operations to determine that we have sufficient degrees of freedom (bypass lines or trim heaters/coolers).

The choice of where production rate is set (Step 4) is often a pivotal decision, but it frequently is determined externally by a business objective. This removes another degree of freedom that cannot be used. If we are free to choose the handle for production rate, then Steps 5 through 7 are the priority order. However, at Step 7 we may determine that the choice will not work in light of other plantwide control considerations, in which case we would return to Step 4 and select a different variable to set production rate. Determining the *best* choice at Step 4 can only be done via nonlinear dynamic simulation of disturbances with a complete control strategy.

Step 5 is done next because the control of product quality is closely tied to Step 1 and is a higher priority than the control of inventories. Hence it should be done early when we still have the widest choice of manipulators available. Its importance is based on the issue of variability, which we want to be as small as possible for on-aim product quality control. Variability in inventory control tends to be not as critical, which is the reason it is done in Step 6.

Only after the total process mass balance has been satisfied can we check on the individual component balances in Step 7. That then settles the plantwide issues. We now apply our knowledge of unit operation control in Step 8 to improve performance and remain consistent with the plantwide requirements. Finally, Step 9 addresses higher level concerns above the base regulatory control strategy.

This, then, is a general and straightforward method for tackling the control system design problem for an entire process. Using the procedure as a framework, we should be able to transform an initially complex and seemingly intractable problem into one that can be solved. Before we illustrate the application of the procedure to four industrial processes, we analyze and summarize the control systems for individual unit operations. We also discuss how they fit into the plantwide perspective.

3.5 Conclusion

We have discussed in detail each of the nine steps in our plantwide control design procedure. The first two steps establish the control objectives and control degrees of freedom for the plant. In the third step we discuss how the plantwide energy management problem can be converted to a local unit operation energy management problem by using the plant utility system.

The heart of the plantwide control problem lies in Steps 4 through 7, where we establish how to set production rate, maintain product quality, prevent excessive changes in recycle flowrates, control inventories, and balance chemical components. These steps demand a plantwide perspective that often leads to control strategies differing significantly from those devised by looking at isolated unit operations.

In Part 3 we illustrate the application of these steps in four industrial processes.

3.6 References

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