Process Control Design: Managing the Design Procedure

CHAPTER 25

25.1 ▫ INTRODUCTION

To this point, the control design problem has been defined, and the range of decisions has been presented. It becomes clear that tens to hundreds of decisions are made during the control design of an industrial process. One would expect, as is shown later in this chapter, that the sequence in which these decisions are made can influence the time required to complete the design and, perhaps, the quality of the control performance provided by the final design. Thus, the engineer is faced with the challenge of managing a large quantity of information and a large set of possible design decisions during the design procedure.

There is no single, correct way to manage this procedure. Different skilled engineers perform tasks in different sequences to reach equally good solutions, and different problems can be solved more easily by different sequences. However, the procedure presented here provides a structured problem-solving approach that is tailored to the control design task. The procedure represents, to the ability of the author to document such a fuzzy entity, the approach used by many practitioners.

There are several advantages to the novice engineer for using this procedure. Since the most difficult aspect of the design is often starting this ill-defined task, the first advantage is that a prescribed procedure provides a way to begin the design task. Second, the procedure provides a step-by-step approach that ensures that many important issues are addressed. Third, the procedure decomposes the problem in a manner that determines whether control is possible before continuing to detailed decisions on control strategies. Finally, the procedure provides some guidance on managing the interactions among the numerous design decisions.

Interactions occur because some decisions made to satisfy specific control objectives affect the possible control performance with respect to other control objectives. Therefore, the engineer must try to make each decision with full recognition of its impact on the entire design and all control objectives. This thought process is demanding and not always possible, so the engineer often has to iterate by returning to initial decisions, changing some, and proceeding from these modified decisions to the completion of the design. The successful design engineer has the foresight to make (generally) good initial decisions, identify improper initial decisions early in the procedure, and minimize the iterations to the final design.

25.2 ◻ DEFINING THE DESIGN PROBLEM

We begin again with the definition of the problem provided in the control design form (CDF) because of the crucial importance of this step to the quality of the design. In this section, some guidance is given on how an engineer goes about filling in a blank CDF. The CDF provides a useful checklist of the information needed in designing control systems and gives an organized manner for documenting the information.

Typically, people need some stimulation when defining problems; that is, they need some questions and issues to consider when beginning the design procedure. To stimulate the thought process, abbreviated tables of sample questions are presented here for the various control objectives. The first three objectives—safety, environmental protection, and equipment protection—are combined in Table 25.1 because they all address major deviations from normal operation, many of which could have common causes that influence all three objectives. Smooth operation, product quality, efficiency and optimization, and monitoring and diagnosis are addressed in Tables 25.2 through 25.5, respectively. The issues raised in the tables should be considered for each design, and issues relevant to the plant should be

TABLE 25.1

Checklist for safety, equipment, and environment

Limitations on operating conditions due to equipment, material, e.g.,
- Composition
- Flow
- pH
- Pressure
- Temperature

Explosion
- Fuel source
- Oxidizing source
- Energy source

Release of hazardous material
Failure of process equipment
Failure of control equipment
Human mistakes and their consequences

TABLE 25.2

Checklist for smooth operation

Unstable processes (do not reach steady state without control)
- Levels
- Chemical reactors

Single controller that influences the production rate
Processes that are very sensitive to disturbances
- Gas pressures
- Liquid pressure

Process integration that either propagates or attenuates disturbance (especially recycle systems)
Manipulated variables that are easily interpreted by operating personnel
Disturbance sources

TABLE 25.3

Checklist for product quality

Target average value and variability
- One or multiple specifications
- Average value
- Variability
- $\pm$ deviation from target at which product is unacceptable

Variability in a property that affects future use by customer
- Standard deviation or other measure
- Nonlinearity between measurement and quality in future use

Disturbances that affect quality
- Magnitude
- Frequency

Factors affecting control performance
- Availability of on-stream measurement
- Degrees of freedom
- Controllability
- Feedback dynamics
- Modelling errors

noted in the control design form, thereby developing a comprehensive statement of control objectives.

An additional way to identify control issues is to pose the following question for every stream or important location (e.g., the volume of a reactor or flash drum) in the process.

TABLE 25.4

Checklist for efficiency and optimization

External manipulated variables not used for control, potentially for optimization
Changes in targets or inputs (disturbances)
- Frequency
- Need for optimization
- Complexity of optimizing strategy

Parallel units
- Different product quality
- Different yields
- Energy consumption

Recycle flows
- Composition of recycle

Separation units
- Energy-yield tradeoff

Chemical reactors
- Conversion
- Yield

Operating condition
- Internal optimum
- Operation at a constraint

TABLE 25.5

Checklist for monitoring and diagnosis

Performance that changes rapidly
- Alarms
- Emergency shutdowns
- Constraint violations
- Product quality
- Inventories

Performance that changes slowly
- Heat transfer coefficient
- Catalyst activity
- Corrosion
- Coking or fouling

Performance requiring complex calculations
- Fired heater efficiency
- Turbine and compressor efficiency

Utilization of control
- Percent of time in automatic

Temporal correlation of good or poor operation with external disturbances (feed type, equipment operation, and so forth)

> What is the effect on A if the B in this stream or location C?
>
> where A = each **control objective** (safety, environmental protection, equipment protection, smooth operation, product quality, efficiency, yield and profit, and monitoring and diagnosis)
>
> B = **property word** indicating key operating variables (e.g., flow, temperature, pressure, composition, inventory, and so forth)
>
> C = **guide word** indicating direction of changes in operation (e.g., increases, decreases) and rate of change (e.g., rapidly, slowly, periodically).

The application of this question to the process will help the engineer identify the significant effects on the objectives. When a significant effect is identified, the engineer should determine the cause of the effect and how it can be retained (if the effect is beneficial) or prevented or compensated (if the effect degrades performance). This is a simplification of an approach that has been developed in much greater detail for hazards and operability (HAZOP) studies, which consider a broader range of issues influencing the safety of a process. Detailed descriptions of the procedures followed in HAZOP studies are available (AIChE, 1992).

The methods described in this section are intended to generate information on all major headings in the CDF, not just objectives, although the tables of questions are organized by objectives. When considering the objectives in such detail, information on the constraints and disturbances should also be identified and recorded in their proper locations. It is important to recognize that the CDF cannot be completed with only a cursory understanding of the process and quick review of a process sketch; a thorough understanding of the physics, chemistry, product quality, and economics is required.

At this preliminary design stage, the engineer should concentrate on determining the needs of the plant and not attempt to define the solutions. The control objectives and other critical issues should be clearly and quantitatively stated even when no solution is initially apparent, and the definition procedure should not be delayed by lengthy analysis of a particular issue, since too much attention to detail during the initial "brainstorming" activity tends to slow the flow of ideas. Also, it is important that the engineer not be overly concerned about the initial location for an element of information in a CDF. It is expected that the CDF will be reviewed and rationalized before the design procedure continues to the decision-making step.

25.3 SEQUENCE OF DESIGN STEPS

There is almost an infinite number of ways in which the numerous design decisions can be reached. There is no one best sequence for all control designs; in fact, various skilled practitioners use different sequences to arrive at equally good designs. However, there are certainly some sequences that are better than others, and some simple sequences can be used by novice engineers until they gain enough experience to modify the sequence to take advantage of their special insights. The sequence given in the flowchart in Figure 25.1 is recommended for control design and discussed further in this and the next sections.

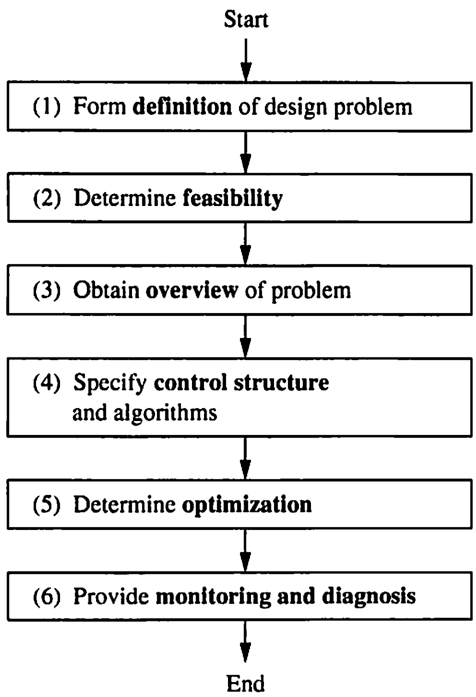


FIGURE 25.1

Overview of control design sequence.

Step 1: Definition

The first step involves the collection of information appearing in the control design form and, for especially complex problems, the formal preparation of the form. At this step, the objectives are translated to specific variables, either directly measured or calculated using measurements, which are to be controlled.

Step 2: Feasibility

The second step determines the feasibility of the control objectives for the equipment design, operating conditions, and disturbances given in the problem definition. An analysis of degrees of freedom and controllability determines whether it is possible to control the proposed controlled variables with the proposed manipulated variables. Since controllability rigorously addresses only the base-case operating point, the operating window is determined to ensure that the process can be maintained within specified limits for the defined disturbance magnitudes. Thus, this step ensures that the system has sufficient capacity as well as degrees of freedom and controllability. As noted in Chapter 24, a dynamic analysis may have to be performed to evaluate the operating window fully. Also, the ability to measure or infer important variables is evaluated. If any of the results of these steps indicate that control is not possible, the design procedure must include an iteration in which an engineer alters the process so that the control objectives can be achieved.

Step 3: Overview

The third step establishes an integrated view of the plant operation, concentrating on the most important variables. The goal of this step is to obtain an *overview* of the feedback process dynamics, the disturbance dynamics, the interaction in the process, and the types of measurements and manipulated variables available for control. This overview is essential because the design engineer makes one decision at a time and needs this overview to be able to "look ahead" so that all decisions form a compatible design. Objectives that are easily achieved or likely to be difficult to achieve are noted. Also, potential changes to the instrumentation and process are identified for future use, if needed. However, no control designs are decided at this step.

Step 4: Control Structure

The fourth step involves specific decisions on control structure, algorithms, and tuning. Here, if single-loop control technology is used, the single-loop controlled and manipulated variables are paired, and the modes of the PID controllers are specified. In addition, special requirements for the tuning are made in conjunction with the pairing. For example, level controllers are specified as tight or averaging. Also, tight and loose tuning of interacting loops is specified, to reduce the effects of unfavorable interaction while retaining the beneficial effects of favorable interaction, as required. The next sections of this chapter provide additional guidance on this step, discussing a hierarchy and decomposition for managing the design decisions.

Step 5: Optimization

The fifth step determines whether optimization opportunities are available after consistently high product quality has been achieved and, if so, whether additional manipulated variables, not used for control at previous steps, exist. It may be necessary to add sensors to provide information for optimization and to automate additional manipulated variables for optimization. If opportunities exist, an analysis is performed to determine the economic benefits which can be realized through optimization, as explained in Chapter 26. If significant benefits are available and can be realized through real-time control, the strategy is designed at this step.

Step 6: Monitoring and Diagnosis

The sixth and final step evaluates monitoring and diagnostics. At this step, the major analysis is the sensors required for this function. In addition, any calculations required for the monitoring are defined.

The sequence of steps is selected to maximize information gathering and understanding at the early steps and to reduce the need for iterations. The first two steps identify the capabilities of the process and instrumentation and the control objectives. Inconsistencies between process capability and objectives are identified so that they can be resolved soon in the design procedure, because inconsistencies should be resolved before further design steps are performed. Next, the overview of the process in the third step enables the engineer to understand the process responses before attempting to design controllers. The design of the controllers, up to and including product quality, is performed in the fourth step to give the best performance for the more important variables. Special controls for safety should be designed at this stage in an integrated manner. In the fifth step, the remaining degrees of freedom, which are not used at the previous stages (perhaps because they have the poorest dynamic responses for control of key variables), are used for profit maximization. Finally, the monitoring and diagnosis is designed.

25.4 □ TEMPORAL HIERARCHY OF CONTROL STRUCTURE

In this section the activities in the fourth step in the sequence, addressing control structure, are presented in greater detail. Proper design relies on an integrated analysis of the entire process or plant under consideration; however, the integrated design may involve too many variables and processes to be analyzed by currently available methods. Therefore, the engineer temporarily separates the design problem into smaller segments, and if the interactions among the segments are small, each can be analyzed individually to develop provisional control designs. Two approaches for selecting segments are discussed: *temporal hierarchy* in this section and *process decomposition* in the next section. It is important to recognize that these methods are used only when required by the large scope of the problem and that the methods employ approximations to simplify the analysis. It is essential that each decision contribute to the good performance when considering all factors in the integrated process.

A common approach for decomposing the design decisions is based on a temporal hierarchy, as originally suggested by Buckley (1964) and expanded here:

In hierarchical decomposition, the control decisions are usually made in the following order.

1. Flow and inventory
2. Process environment
3. Product quality (and safety)
4. Efficiency and profit
5. Monitoring and diagnosis

This hierarchy has the advantage of designing control loops in the order of the fastest to the slowest; the possible exception are the liquid and solid inventories, which may employ averaging controllers (see Chapter 18). In addition, the hierarchy is commonly used because it is difficult to design controllers for product quality without first defining how feed and product flows and process environments are controlled. Thus, the sequence makes sense from the viewpoint of control structure.

Flow and Inventory

Here, the flows and inventories considered are for the "process" materials, which are used to make the product. The flows of utility streams, such as fuel, cooling water, and steam, are not specified here, because they are manipulated to achieve other control objectives. The structures that control the process flows determine how the feed and production rates are specified and whether flow rates are nearly constant or are likely to vary significantly. Note that the inventories—liquid and solid levels and gas pressures—must be designed in conjunction with the flow controllers, to ensure that requirements for inventories and product deliveries are satisfied concurrently.

The goal is to provide a design in which the overall material and component compositions are stable without further control. Naturally, this does not imply that satisfactory performance is achieved with only these controls, only that all material entering the process leaves the process at steady state, which is a reasonable basis for further analysis. One controller should influence the production rate; this is usually a flow controller at the beginning (feed) or end (product) of the plant, although other designs are possible. Then, the liquid levels and gas pressures are controlled in a manner to achieve a self-regulatory process.

Particular attention should be paid to the compositions in *recycle processes*. Because of the economic value of materials, material that is not reacted or not of sufficient purity is typically recycled to an upstream position in the process. If no method is provided for impurities (e.g., inerts) to exit the system, they will accumulate in the process and ultimately lead to major upsets. One common technique to improve dynamic behavior is to provide a small purge to allow inerts to behave in a self-regulatory manner; this design is common in spite of the economic losses due to valuable materials also leaving in the purge. Control designs should ensure that feed components are self-regulating, so that they do not accumulate in the process. The reactor in Section 25.7 demonstrates the unique dynamic responses associated with compositions in a process with recycle.

Process Environment

The second level addresses the process environment variables: pressure, temperature, feed ratios, catalyst addition, and so forth. These variables have a great influence on the product quality and are often manipulated, in a cascade structure, by the product quality controllers. Thus, this level provides tight control of the environment by compensating for many disturbances, and it can be adjusted by cascade feedback from higher levels.

Product Quality (and Safety)

The third level provides the essential product quality regulation. This is typically achieved by adjusting set points of controllers at the lower levels in a cascade structure, but it may adjust final elements directly. Control for safety should be addressed at this level of the decision hierarchy, because control strategies up to this level can influence the safe operation. As discussed in the previous chapter, the safety controllers will normally be implemented in a lower level of the implementation hierarchy.

Efficiency and Profit

The fourth level capitalizes on additional flexibility to improve profitability of the plant. These controllers perform their function slowly so that smooth operation and excellent product quality are not sacrificed. It is good practice for the optimizing controllers to influence the process through the lower levels in the implementation hierarchy; this ensures that higher-priority objectives such as safety and product quality are not compromised.

Partial Control

Partial control is not a separate level in the control hierarchy, but concepts related to partial control influence decisions in levels 2 to 4 of the hierarchy. Recall that partial control involves selecting of a subset of variables that can be measured and controlled, so that all key variables remain within an acceptable range as disturbances occur. To achieve partial control, the engineer seeks *dominant variables* that strongly influence the process behavior, and when regulated, yield good process performance. Some examples of typical dominant variables are given in the following summary.

Unit operation	Typical dominant variables for partial control: Process environment sensors (T,P,F,L)	Typical dominant variables for partial control: Analyzers	Typical product qualities
Chemical reactor	Temperature Pressure (gas phase) Liquid level Flow rate	Reactant concentration Product concentration	All concentrations in the product stream Product properties, e.g., octane or average molecular weight
Distillation	Tray temperature(s) Pressure Reflux ratio Boilup ratio	Heavy or light key component concentration	All concentrations in the product streams Product properties, e.g., vapor pressure
Heat exchange	Coolant flow rate Coolant temperature Level of boiling refrigerant		Effluent temperature

> It is important to recognize that detailed knowledge of the specific process behavior is required to select proper dominant variables.

For example, the coolant flow rate usually has an effect on the hot-side effluent temperature from a shell and tube heat exchanger. However, if the heat exchanger is "pinched," i.e., the hot effluent temperature is essentially the same as the entering coolant temperature, an increase in the coolant flow will not have an effect on hot stream exit temperature. In this situation, the inlet coolant temperature could serve as a dominant variable. There is no alternative to good process knowledge! For further discussion of partial control and dominant variables, see Luyben et al. (1998) and Arbel et al. (1996, 1997).

Monitoring and Diagnosis

The fifth level involves monitoring and diagnosis of process and control performance. This includes rapid monitoring and reporting to plant operating personnel, as well as longer-term monitoring for periodic analysis. Plant operations are influenced by decisions made at this level through actions of plant personnel, usually after detailed analysis of likely causes of unusual process performance. These decisions may not be implemented through the control strategies, because they may involve variables, such as feed purchases and reactor regeneration scheduling, that are outside of the purview of the continuous control system.

This analysis hierarchy conforms to the way many control systems are implemented. A typical implementation hierarchy is shown in Figure 25.2. The lowest

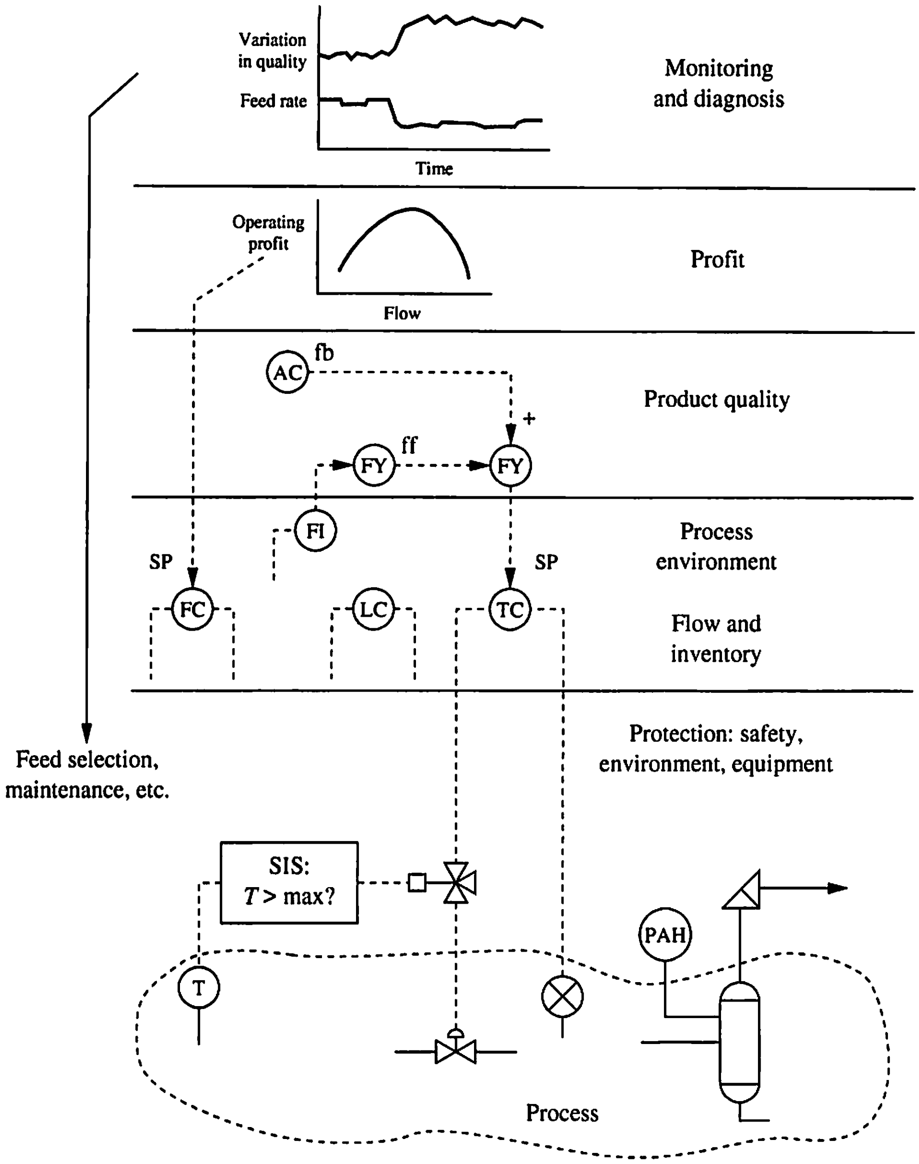


FIGURE 25.2

Schematic of the typical process control hierarchy.

level of the continuous control involves the flow and inventory loops and provides the basis for higher levels in the hierarchy. Note that the interaction between levels in the hierarchy is primarily through cascade control principles; this approach has several advantages:

1. It uses conventional technology.
2. It satisfies the requirements for relative dynamics so that good disturbance response is achieved.
3. It does not create conflicts in degrees of freedom (see Section 14.2).
4. The system is easily commissioned or decommissioned by changing controller cascade status between closed and open.

Although this hierarchical approach has many advantages and has been found easy to apply by many engineers, it does not remove one of the most challenging features of the design procedure: the need for iteration. When making decisions at each level, the engineer attempts to look ahead to the completed design and determine the effects of the current decisions on the control performance. However, looking ahead is not always simple, or even possible, in complex plants; thus, the engineer may find that the final design is not satisfactory. When such a situation is encountered, the engineer should investigate whether the performance could be improved by another design that starts with different decisions at the previously designed, lower levels in the hierarchy.

EXAMPLE 25.1.

Consider the flash process in Figure 25.3, which is similar to the process previously analyzed in Chapter 24. The case considered here involves two different initial flow and level control decisions, shown in Figure 25.3*a* and *b*. The first level of the hierarchy in Figure 25.3*b* has resulted in the control design in which the feed is on flow control, and the level is controlled by adjusting the heat transferred to the feed by adjusting the steam, which affects the amount of liquid vaporized. These initial decisions satisfy the relevant control objectives. However, given these flow and level decisions, the product quality controller has only one degree of freedom to adjust: the product flow rate. Therefore, the lower-level design decisions have dictated the higher-level control strategy.

To understand how the quality controller in Figure 25.3*b* would function, consider the case in which the light key in the liquid product component is too high. In response to the disturbance, the product quality controller would decrease the product flow rate, which would cause the level to increase; the level controller would increase the steam flow rate, which would increase the percentage vaporized; and the light key in the liquid product would decrease. Therefore, this quality control design is feasible, but it has slow dynamics, because the level control process and controller appear in the product quality feedback path. In fact, this

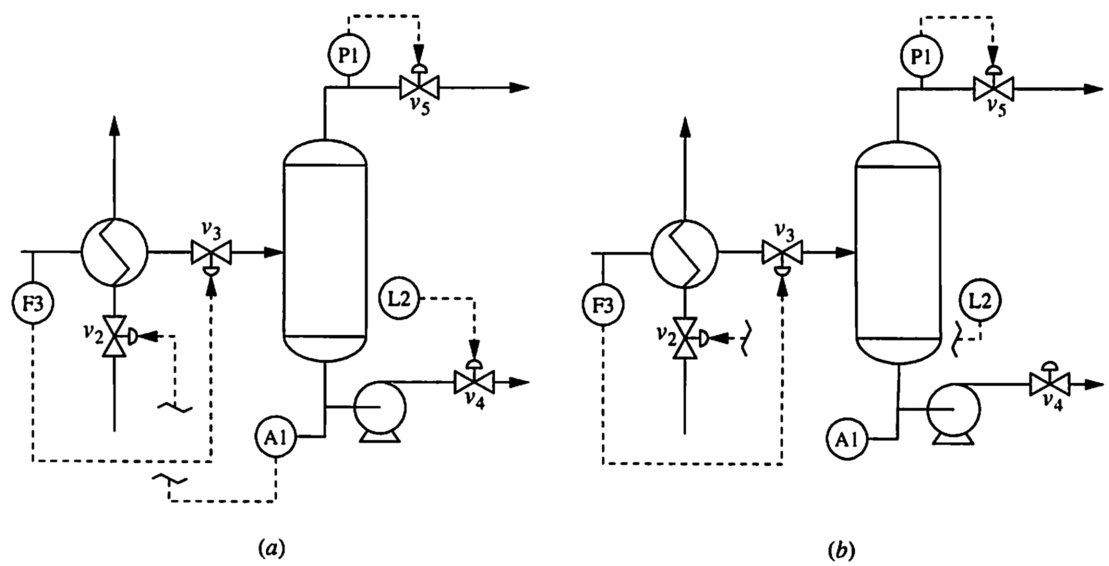


FIGURE 25.3

Two different flash control designs discussed in Example 25.1.

design is another example of pairing single-loop controllers with a relative gain of zero. This can be verified using the following steady-state model, which has been extracted from equation (24.5) for the flash process in Chapter 24:

$$\begin{bmatrix} A_1 \\ \dfrac{dL}{dt} \end{bmatrix} = \begin{bmatrix} -0.11 & 0.0 \\ -0.136 & -0.179 \end{bmatrix} \begin{bmatrix} v_2 \\ v_4 \end{bmatrix} \tag{25.1}$$

The relative gain for the $A_1 \rightarrow v_4$ pairing is zero, because the steady-state process gain between the product flow and the composition is zero, when all other loops are open. The relative gain for this system has ones on the diagonal and zeros on the off-diagonal elements. However, since the system is controllable for either pairing, a pairing on a zero relative gain would function, albeit with poor performance in this case.

Thus, the initial flow/inventory control design decisions have resulted in relatively poor product quality control. During the iteration, the engineer would be looking for a faster-responding manipulated variable for product control, because the cause of the poor performance is slow feedback dynamics. Another goal would be to find a pairing with a nonzero relative gain. After the iteration, the control design should be as shown in Figure 25.3*a*.

How does the engineer properly perform the "look-ahead" to satisfy the control objective under consideration while preventing, as much as possible, an undesirable effect on other control objectives? The effect of "keeping in mind" is to ensure that the initial control design, in addition to meeting its control objectives,

1. Leaves unallocated some manipulated variables that can give good control performance for important controlled variables appearing at higher levels in the hierarchy.
2. Attenuates disturbances and does not introduce unfavorable process control interactions.
3. Provides good integrity, if possible, so that critical controllers can perform their tasks properly even if some other controllers are not functioning (e.g., are in manual) without retuning.

This look-ahead requires an overview of all control objectives, which again reinforces the importance of a good problem definition and process overview in steps 1 to 3 of the sequence. Then the engineer must keep all of the key controlled variables in mind when designing the lower levels of the hierarchy.

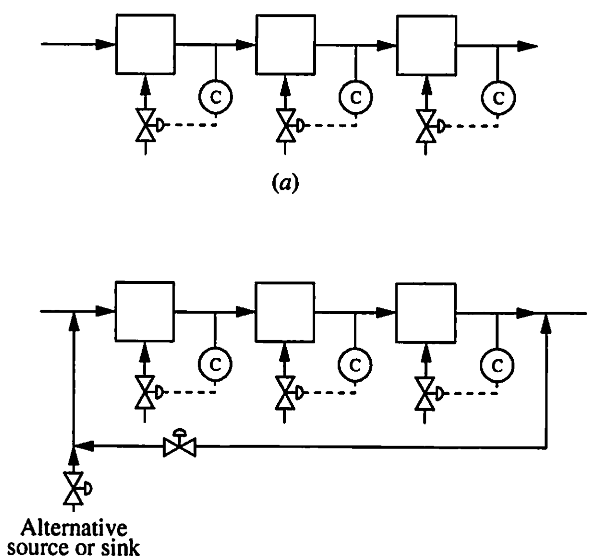
> When performing the control design procedure, the engineer continually *looks ahead* to predict the effects of current decisions on later control objectives at higher levels in the hierarchy.

25.5 □ PROCESS DECOMPOSITION

Large plants may have hundreds or thousands of manipulated and controlled variables. Although the entire plant must be considered in designing controls, it is essentially impossible to analyze all aspects of the plant simultaneously while making each control decision. Therefore, the plant is often decomposed into several process units that have only weak interactions, if possible. The proper decomposition is particularly easy for the series process design structure of chemical process plants, shown in Figure 25.4*a*. For this process structure the upstream units affect the downstream units, but the downstream units do not affect the upstream units. Since the interaction among the units is in only one direction, upstream units are simply sources of disturbances to the downstream units. Thus, the general goal is to reduce the disturbances that leave one unit and propagate to downstream units, with special care to isolate units that are highly sensitive to disturbances. The controls within each process unit can then be designed using the standard procedures.

Process plants often have recycle streams, as shown in Figure 25.4*b*. These plants do not strictly allow such a simple decomposition, because two-way interaction occurs between processes. As demonstrated in Chapters 20 and 21, two-way interaction can significantly affect dynamic behavior and control performance. Usually, the control system is designed to reduce the effects of recycle on the overall plant dynamics. This is often achieved by providing alternative sources of the material or energy provided by the recycle, so that short-term variation in the recycle can be compensated by the alternative source. (This is the same concept used in Figure 24.11*a* and *b* for energy recycle.)

Two examples of material recycle are shown schematically in Figures 25.4*b* and 25.5. In the first, an alternative source of material is provided to ensure a steady recycle flow; in this design, the alternative must be available immediately to provide the total process flow required. In the second example in Figure 25.5, the recycle system includes an inventory so that the level in the inventory can vary while the material supplied to the beginning of the process remains undisturbed.



(*b*)

FIGURE 25.4

Typical structure of process plants: (*a*) series; (*b*) recycle without storage.

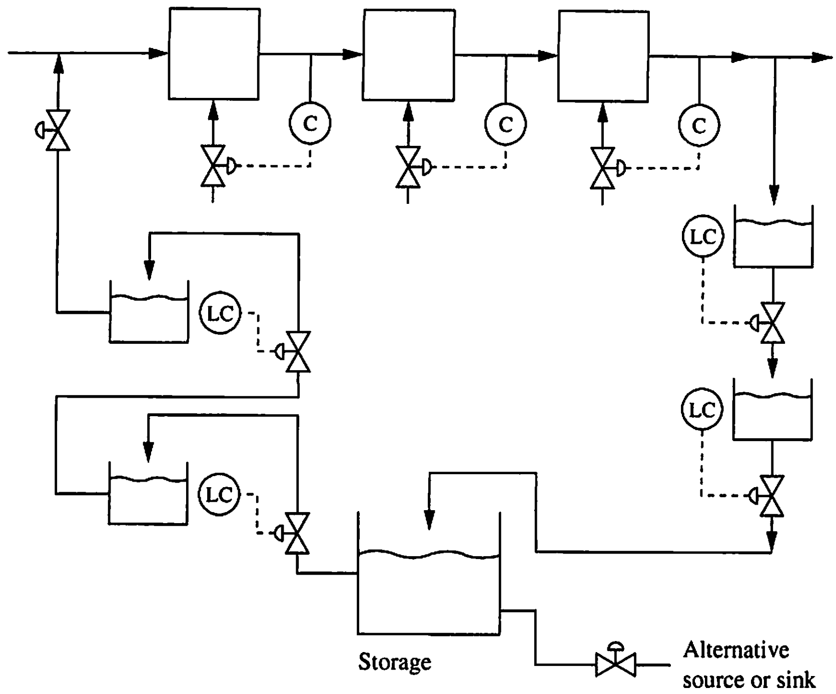


FIGURE 25.5

Typical recycle process with storage.

Note that the level-flow pairing directs all recycled material to the storage tank, regardless of the current recycle flow returned to the process. Naturally, the storage tank must be large enough for the level to remain within acceptable limits during expected short-term transients, whereas the net flow in or out must compensate for longer-term accumulation. This concept is discussed by Buckley (1974), where he describes the principle that recycle level-flow systems should generally be paired in the manner shown in Figure 25.5.

25.6 ▫ INTEGRATING THE CONTROL DESIGN METHODS

Several methods for organizing information and making design decisions have been presented in this and the previous chapter. In this section, the methods are combined into an integrated design thought process that demonstrates how the previously discussed methods can be combined to reach an adequate design. Novice engineers will most likely follow this integrated approach closely for their initial designs. As they gain more experience and learn to use their process understanding, they will adapt the approach to suit the problem at hand.

The integrated procedure is shown in Table 25.6, which combines the concepts of sequence, hierarchy, and design decisions. The procedure begins with a statement of the process design and plant requirements and ends with a complete control structure and algorithm specification. The major steps in the design sequence—(1) definition, (2) feasibility, (3) overview, (4) control structure, (5) optimization, and (6) monitoring and diagnosis—provide milestones for the procedure. Several quantitative design analyses are performed at each step in the design sequence,

TABLE 25.6

Integrated control design procedure

START: Acquire information about the process

(a) Process equipment and flow structure [Modify process and instrumentation]

(b) Operating conditions

(c) Product quality and economics

(d) Preliminary location of sensors and final elements

1. DEFINITION: Complete the Control Design Form

(a) Use checklists

(b) Sample questions [Modify objectives]

(c) Prepare a preliminary set of controlled variables

2. FEASIBILITY: Determine whether objectives are possible

(a) Degrees of freedom [Iterate]

(b) Select controlled variables and evaluate controllability

(c) Operating window for key operating conditions

3. OVERVIEW: Develop understanding of entire process to enable "look-ahead" in decisions

(a) Key production rate variables

(b) Inventories for potential control

(c) Open-loop unstable processes

(d) Complex dynamics (long delays, inverse response, recycle, strong interactions)

(e) Key product qualities

(f) Key constraints

(g) Key disturbances

(h) Useful manner for decomposing the analysis (and control design), if necessary and appropriate

4. CONTROL STRUCTURE: Selection of controlled and manipulated variables, interconnections (pairings in decentralized control), and relevant tuning guidelines

(a) Preliminary decisions on overall process flows and inventories

(b) Process segment (Unit) 1

(c) Process segment (Unit) 2

Control hierarchy (temporal decomposition) for every unit [Iterate]

1. Flow and inventory
2. Process environment
3. Product quality
4. Safety

[Modify control]

(d) Integrate control designs as needed for good overall performance

5. OPTIMIZATION: Strategy for excess manipulated variables

(a) Clear strategy for improved operation, or

(b) Measure of profit using real-time data

(c) Sensors and final elements [Iterate]

(d) Minimize unfavorable interaction with product quality

6. MONITORING AND DIAGNOSIS

(a) Real-time operations monitoring

1. Alarms
2. Graphic displays and trends

[Iterate]

(b) Process performance monitoring

1. Variability of key variables (histogram and frequency range)
2. Calculated process performances (efficiencies, recoveries, etc.)

FINISH: Completed specification, meeting objectives in step 1

(a) Process equipment and operating conditions

(b) Control equipment, sensors, and final elements

(c) Control structure and algorithms

(d) Tuning guidelines as needed, e.g., level control and interacting loops

(e) Safety controls and alarms

(f) Optimization

(g) Monitoring calculations

and the first three levels of the temporal hierarchy are performed for each process segment at the fourth step. The engineer will encounter all major design decisions presented in Chapter 24 in a logical order by using the procedure in Table 25.6.

Iterations are possible at several steps in the procedure. If the process does not have sufficient degrees of freedom, lacks independent input-output relationships to provide a controllable system, or lacks sufficient range, an iteration is required in step 2 to change the process. Also, if an analysis of the dynamics identifies poor control performance, an iteration in step 4 is appropriate. Further iterations may be needed to provide all sensors necessary for optimization and monitoring in steps 5 and 6. During each iteration, the control objectives should also be reevaluated to be sure that the quantitative performance targets are proper and that the cost associated with achieving the demanding goals is justified.

As previously discussed, the engineer makes every effort to reduce or eliminate iterations by making the sequential design decisions with due consideration for future decisions. Information in steps 1 through 3 enables the engineer to identify the likely key elements of the design (i.e., the controlled variables requiring tight control). This enables the engineer to "set aside" manipulated variables that may be used for the control of the key variables. The integrated control design procedure is demonstrated in the following example.

25.7 ◻ EXAMPLE DESIGN: CHEMICAL REACTOR WITH RECYCLE

The integrated control design procedure will be applied to a simple chemical process in this section. The process, shown in Figure 25.6, involves feed of a raw material from storage to a chemical reactor. The reaction is A $\rightarrow$ B with first-order

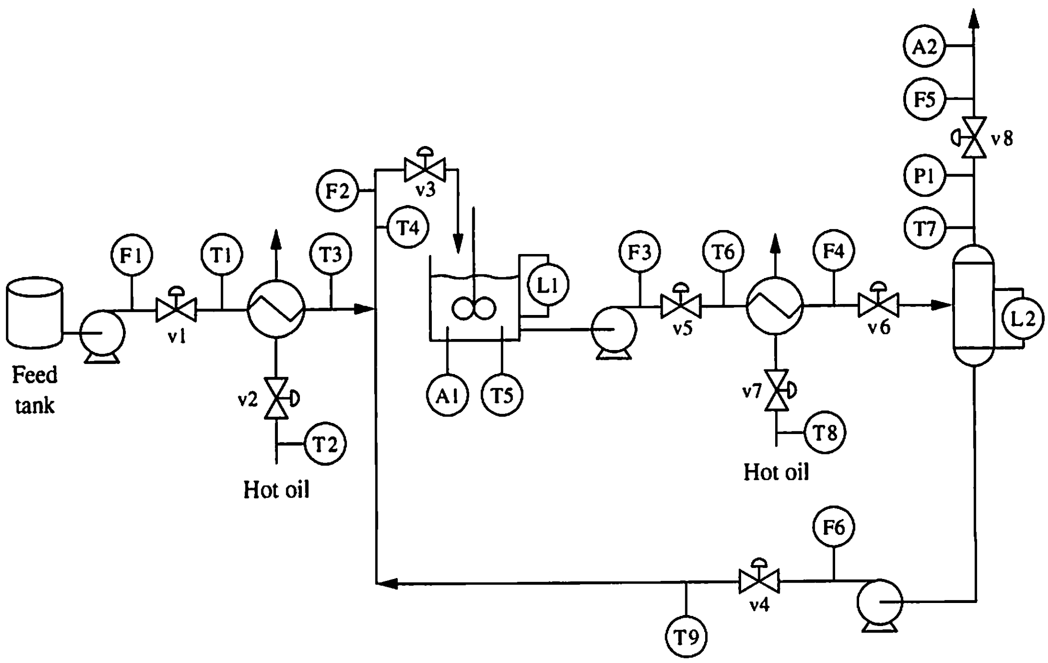


FIGURE 25.6

Chemical reactor and separator with recycle.

rate expression $-r_A = k_0 e^{-E/RT} C_A$ and negligible heat of reaction. The products of the reactor are heated and sent to a flash drum, from which the product is taken as a vapor flow which is predominantly component B, but contains some A. A liquid stream consisting of unreacted feed, along with some product B, is recycled to mix with the fresh feed and flows to the reactor. The base-case (initial) operating variables are given in Table 25.7.

Definition Step

The control design will be developed through the procedure shown in Table 25.6. The first step in the sequence involves a complete definition of the problem, which is summarized in the control design form in Table 25.8. (The reader should review the table before proceeding.) This serves as the basis for all further design decisions.

Feasibility Step

The second step determines whether the control objectives are possible with the equipment available. This step involves the analysis of degrees of freedom and controllability. We assume that an analytical model of the process is not available; thus, the design is based on qualitative analysis from the process structure and on linear models identified empirically. There are eight manipulated external variables, so at most eight dependent variables can be controlled. A preliminary selection of controlled variables is made based on the CDF: (*a*) feed or production rate (1); (*b*) liquid and vapor inventories (3); and (*c*) product quality (1). Thus, at least five controlled variables exist. The number of external manipulated variables is greater than this minimum value. Therefore, it is concluded that the degrees of freedom do not preclude a possible design, and the design procedure can continue.

TABLE 25.7

Operating conditions for reactor with recycle

Variable	Symbol	Initial value	*Final value for Design I in Figure 24.9	*Final value for Design II in Fiqure 24.10
Fresh feed	F1	5	5	5.0
Reactor inlet flow	F2	20	34	20
Reactor outlet flow	F3	20	34	20
Vapor product	F5	5	5	5
Recycle flow	F6	15	29	15
Reactor level	L1, %	50	50	50
Flash level	L2, %	50	50	50
Fresh feed temperature	T3, °C	99	105	106.8
Reactor feed temperature	T4, °C	92	92	93.9
Reactor temperature	T5, °C	92	92	93.9
Flash temperature	T7, °C	90	90	90
Reactor concentration of A	A1, mole %	69.4	77.1	69.4
Vapor product concentration of A	A2, mole %	10	10	10

*After response to disturbance (1) in Table 25.8.

To extend the analysis further, the controllability of the system is evaluated. Controllability requires that linearly independent relationships must exist between the selected manipulated and controlled variables, or, in other words, the gain matrix must have a nonzero determinant. To perform this analysis, the model equations have to be linearized, and the matrix of gains evaluated at the base-case operation. Since inventory control is quite important, the level and pressure control loop pairings are decided first. The reactor level can be controlled with either v_5 or v_6, the flash drum level with v_4, and the flash drum pressure with v_8. The steady-state gains with these inventories under closed-loop control were determined by making small changes to the manipulated variable and determining the steady-state change in the potential controlled variables. The gain matrix for this example is

$$\begin{bmatrix} F_1 \\ F_2 \\ T_3 \\ T_5 \\ A_1 \\ A_2 \end{bmatrix} \begin{bmatrix} 0.020 & 0.0622 & 0.0106 \\ 0.45 & -0.38 & -0.127 \\ -0.13 & 2.14 & -0.13 \\ -0.02 & 0.44 & 1.105 \\ 0.0035 & -0.0063 & -0.0018 \\ 0.00025 & -0.0008 & 0.0006 \end{bmatrix} \begin{bmatrix} v_1 \\ v_2 \\ v_7 \end{bmatrix} \tag{25.2}$$

Note that the matrix is not square, so that control of all the potential controlled variables in equation (25.2) is not possible.

By the completion of the design procedure, there will be a strategy for every valve, and the system will be square, but at this point the goal is to determine whether the selected variables can be controlled. One way to answer this question is to select subsets of the controlled and manipulated variables until either (1) a subset results in a nonsingular gain matrix, in which case the system is controllable, or (2) all possibilities have been exhausted without finding a nonsingular system, in which case the system is not controllable. A more direct approach is to find the rank of the matrix, which gives the smallest square subset that is nonsingular. As subsets of the variables are selected, the controllability will be verified.

Overview Step

The third step of the control design sequence, which yields an overview of the process and control objectives, is now performed. The purpose of this step is to gather observations about the entire system that can be used when making sequential design decisions. The observations at this step are presented below by hierarchy level.

LEVEL 1: FLOW AND INVENTORY.

1. The feed tank has periodic deliveries of material and continuous outflow to the process. Therefore, it is not possible or necessary to control the level. The tank must be large enough so that it neither overflows nor goes empty for expected delivery and outflow policies.
2. The feed to the reactor is a combination of fresh feed and recycle. The flow and inventory design must consider this factor, to prevent oscillations caused by interactions. Also, there seem to be several possible ways to control the flow to the reactor, because there are valves in the fresh feed, recycle flow, and combined flow.

TABLE 25.8

Preliminary Control Design Form for the chemical reactor and separator process in Figure 25.6

TITLE: Chemical reactor	ORGANIZATION: McMaster Chemical Engineering
PROCESS UNIT: Hamilton chemical plant	DESIGNER: I. M. Learning
DRAWING: Figure 25.6	ORIGINAL DATE: January 1, 1994
	REVISION No. 1

Control objectives

1. Safety of personnel
   (*a*) The maximum pressure in the flash drum must not be exceeded under any circumstances.
   (*b*) No material should overflow the reactor vessel.
2. Environmental protection
   (*a*) None
3. Equipment protection
   (*a*) None
4. Smooth, easy operation
   (*a*) The production rate, F5, need not be controlled exactly constant; its instantaneous value may deviate by 1 unit from its desired value for periods of up to 20 minutes. Its hourly average should be close to its desired value, and the daily feed rate should be set to satisfy a daily total production target.
   (*b*) The interaction of fresh and recycle feed should be minimized.
5. Product quality
   (*a*) The vapor product should be controlled at 10 mole% A, with deviations of ±0.7% allowed for periods of up to 10 minutes.
6. Efficiency and optimization
   (*a*) The required equipment capacities should not be excessive.
7. Monitoring and diagnosis
   (*a*) Sensors and displays needed to monitor the normal and upset conditions of the unit must be provided to the plant operator.
   (*b*) Sensors and calculated variables required to monitor the product quality and thermal efficiency of the unit should be provided for longer-term monitoring.

Measurements

Variable	**Sensor principle**	**Range**	**Special information**
F1	Orifice	0–10	
F2	Orifice	0–40	
F3	Orifice	0–40	
F4	Orifice	0–40	
F5	Orifice	0–10	
F6	Orifice	0–40	
L1	Δ pressure		Reactor residence time is 5 minutes
L2	Δ pressure		Drum liquid hold-up time is 5 minutes
P1	Piezoelectric		
T1	Thermocouple	0–100°C	
T2	Thermocouple	100–200°C	
T3	Thermocouple	50–150°C	
T4	Thermocouple	50–150°C	
T5	Thermocouple	50–150°C	

TABLE 25.8

Continued

Measurements

Variable	Sensor principle	Range	Special information
T6	Thermocouple	50–200°C	
T7	Thermocouple	50–150°C	
T8	Thermocouple	250–350°C	
A1	Continuous	0–100 mole%	mole% A in reactor
A2	Continuous	0–15 mole%	mole% A in product

Manipulated variables

I.D.	Capacity (at design pressures)	I.D.	Capacity (at design pressures)
	(% open, maximum flow)		(% open, maximum flow)
v1	50.6%, 10	v5	70.0%, 29
v2	9.6%, 100	v6	18.1%, 110
v3	50.0%, 40	v7	60.3%, 67
v4	26.9%, 58	v8	50.0%, 10

Constraints

Variable	Limit values	Measured/ inferred	Hard/ soft	Penalty for violation
Drum pressure	High	Measured	Hard	Personnel injury
Reactor level	Low	Measured	Hard	Pump damage
	High	Measured	Hard	Overflow hazard
Light key A in product	High	Measured	Soft	Reduced selectivity in downstream reactor

Disturbances

Source	Magnitude	Period	Measured?
1. Impurity in feed (influences the reaction rate, basically affecting the frequency factor k_0)	10% rate reduction	Day	No
2. Hot oil temperature	±20°C	200+ min	Yes (T2)
3. Hot oil temperature	±20°C	200+ min	Yes (T8)
4. Feed rate	±1, step	Shift-day	Yes (F1)

Dynamic responses
(Input = all manipulated variables and disturbances)
(Output = all controlled and constraint variables)

Input	Output	Gain	Dynamic model
[see equation (25.2) for some steady-state gains]			

Additional considerations

None

3. There is no option for the disposition of the reactor effluent; it must proceed directly to the flash drum.
4. The vapor product comes from a small drum inventory, and flow rate fluctuations can be expected. Since the control objectives allow for variability in the product rate, this is not likely to be a concern.
5. Two liquid levels are non-self-regulatory and should be controlled via feedback to prevent them from exceeding their limits. Also, one vapor space pressure, while theoretically self-regulating, can quickly exceed the acceptable pressure of the equipment; therefore, the pressure should also be controlled.

LEVEL 2: PROCESS ENVIRONMENT.

6. The liquid phase chemical reactor operation is influenced by several *dominant variables*, temperature, volume, flow rate, and compositions. Based on the concept of *partial control*, we will likely select one (or more) of these to control the reactor. Recall that the best dominant variable will maintain all other key variables close to the best possible operation, as measured by product quality and profit.

   Since the plant has a recycle, we should be sure that the total material and all component compositions are self-regulating. Three categories of components are considered.

   - *Volatile inerts* will exit the plant in the product stream. If present, *heavy inerts* have no stream by which to exit the process and would accumulate without bound in the liquid phase. Since no inerts are considered in this problem, we will not design a liquid purge; however, at least a periodic liquid purge controlled manually by plant personnel should be provided for a plant of this design.
   - *Products* will leave the plant in the vapor stream from the flash separator and will not accumulate.
   - *Reactant* A is not completely converted in the reactor, and the unconverted A will remain in the liquid phase of the flash separator and return to the reactor via the recycle. (Note that only a fixed percentage of reactant is allowed to exit with the product.) Therefore, the reactant will have a tendency to accumulate in the plant. Clearly, one important control objective is to provide self-regulation for the composition of reactant A.

   Further discussion of the potential for component accumulation and designs to provide self-regulation are available in Downs (1992), and Luyben et al. (1998).

LEVEL 3: PRODUCT QUALITY.

7. There appear to be several manipulated variables that affect the flash product quality, A_2.

LEVEL 4: PROFIT.

8. There are no objectives specified to increase profit beyond controlling product flow rate and quality. However, there appear to be extra manipulated variables, or at least extra valves in the process. This inconsistency must be resolved.

We should note that the plant could have been designed without recycle, but the high conversion of A would have required a very large (and expensive) reactor. Another typical reason for recycle in reaction systems is the suppression of side reactions; for the reaction A $\rightarrow$ B $\rightarrow$ C, a low concentration of B in the reactor ensures a small production of unwanted byproduct C. A low "single-pass" conversion leads to a large recycle to achieve a high "overall or total" conversion. However, recycles involve costs as well. Additional equipment is required to process the material and extra heating and cooling is typically required for the recycled material. Therefore, a balance is required in the design and operation of recycle systems. The control design should maintain "moderate" changes in recycle flow rates in response to disturbances, because very large changes would require expensive equipment with very large maximum capacities.

Control Structure Step

Since no severe difficulties were identified in the third step, we proceed to the fourth step, where we begin to design the control structure. Since we anticipate strong interaction among variables because of the process recycle, process decomposition is not applied. However, the control is designed according to the five-level temporal hierarchy. The overall structure is first selected; then, enhancements are added; finally, algorithms and modes are chosen.

LEVEL 1: FLOW AND INVENTORY. The first decision is usually the flow controller, which determines the throughput in the process. Usually, this controls either the feed rate or the production rate. The control objectives state that the production rate does not have to be maintained invariant, which is fortunate, because controlling the vapor flow from a flash drum would be difficult without allowing the pressure to vary excessively. For this process and objectives, the feed rate F_1 will be controlled. Any of three valves, v_1, v_3, or v_4, could be adjusted to control F_1. From the overview, it is realized that v_4 may be adjusted to control the liquid level control in the flash drum, so this is eliminated from consideration as a manipulated variable for controlling F_1. Either of the remaining valves may be adjusted to control F_2. Somewhat arbitrarily, we select v_1 as the manipulated variable; this selection has the minor advantage that the fresh feed can be reduced to zero and the system operated on total recycle for a short time. The remaining valve, v_3, is not needed and could be removed; in the example, we will simply maintain the valve position constant at its base-case value.

The reactor level must be controlled, because it is non–self-regulating, and the residence time affects the chemical reaction. The outlet flow is manipulated to control the level, because the inlet flow has already been selected as the feed flow controller. The outlet flow is affected by both valves v_5 and v_6; thus, there are one controlled and two manipulated variables. We select valve v_6, to maintain the highest pressure in the heat exchanger, which tends to prevent vaporization. The redundant valve, v_5, will not be adjusted.

The liquid level in the flash must also be controlled within limits, and no objective compels tight or averaging control. Tight level control is selected, because the level control is part of the recycle process, and the entire process would not attain steady-state operation until the level attains steady state. The valve v_4 was allocated to control the level when the feed flow was designed.

The final issue at this level of the hierarchy is the pressure control of the flash drum. The vapor valve v_8 is selected to give fast control of pressure.

In summary, the following allocation of controlled and manipulated variables has been made at this point.

Controlled	Manipulated
F_1	v_1
L_1	v_6
L_2	v_4
P_1	v_8

LEVEL 2: PROCESS ENVIRONMENT. Here, we will select a dominant variable for control of the chemical reactor. In general, the temperature, flow, level, and composition(s) are dominant for a liquid-phase reactor. In this process, the feed flow, F_2, is the sum of the fresh feed flow and recycle flow, and these flows have been determined by level 1 controllers; therefore, they are not available for reactor dominant controlled variables. Also, we somewhat arbitrarily decide to maintain the chemical reactor volume constant. Therefore, the dominant reactor variable will be either the temperature or the concentration, and either of these variables can be controlled by adjusting the preheating valve, v_2. We will evaluate two control designs using different reactor dominant variables and select the best design based on closed-loop dynamic performance.

LEVEL 3: PRODUCT QUALITY. The flash composition is to be controlled, because it is the key measure of product quality; it is controlled directly, without a temperature cascade, because the composition sensor is continuous with fast dynamics. The proper choice for the manipulated variable would be the heating oil valve v_7, because it gives fast feedback dynamics over a large range of operation.

In summary, the following allocation of controlled and manipulated variables has been made at levels 2 and 3.

Controlled	Manipulated
Reactor	v_2
A_2	v_7

A reactor variable to be controlled has not yet been selected and could be temperature or concentration. Two alternative designs will be evaluated: temperature control and reactor concentration control.

Optimization Step

There are no optimization objectives in the control design form. The control design to this point has allocated all manipulated variables, except for v_3 and v_6, which were found to be redundant for the previous control objectives. These valves provide no additional process flexibility, except that of controlling some intermediate pressures in liquid flow lines. There seems to be no reason to control these pressures, and ordinarily, these valves would be eliminated to save equipment and pumping costs. In this case, the valves will simply be retained at their base-case percent opening.

To complete this step, enhancements to the basic structure of controller pairings are considered. For this simple process, the enhancements will be restricted to cascade and feedforward, and each controlled variable is discussed individually.

F_1: The flow process is very fast, and the control design needs no enhancement. A PI controller is appropriate for this process, with nearly no dead time and significant high-frequency noise.

L_1: The process has little or no dead time, and the pump pressure is relatively constant. Thus, no cascade or feedforward is required, although a level-flow cascade may be used. The algorithm selected is a PI with tight tuning, because the level influences the residence time, and zero steady-state offset is desired.

L_2: The process has little or no dead time, and the pump pressure is relatively constant. Thus, no cascade or feedforward is required, although a level-flow cascade may be used. The algorithm is a PI with tight level tuning.

P_1: The process is fast, and the pressure should be maintained at its set point, because it affects safety and the flash product composition. Therefore, a PI controller is selected.

A_2: The concentration of A in the product stream is the key product quality and is affected by the disturbance in T_8. Note that a cascade is not possible, because there is no causal relationship between the valve v_7 and the measured variable T_8. A feedforward controller is possible, because the criteria for feedforward would be satisfied. However, as a preliminary decision, no enhancement will be selected, because of the relatively fast feedback dynamics. This decision will be evaluated at the completion of this study. The feedback controller should have a PI or PID algorithm, depending on the dynamics, fraction dead time, and measurement noise.

Finally, the reactor environment control options are evaluated to determine the best control design. Each is discussed briefly as follows.

1. Design I, shown in Figure 25.7, controls T_5. The reactor temperature is affected by several disturbances. These disturbances influence other measured variables before the reactor temperature measurement responds; thus, the potential for enhancements exists. For example, the measured fresh feed temperature

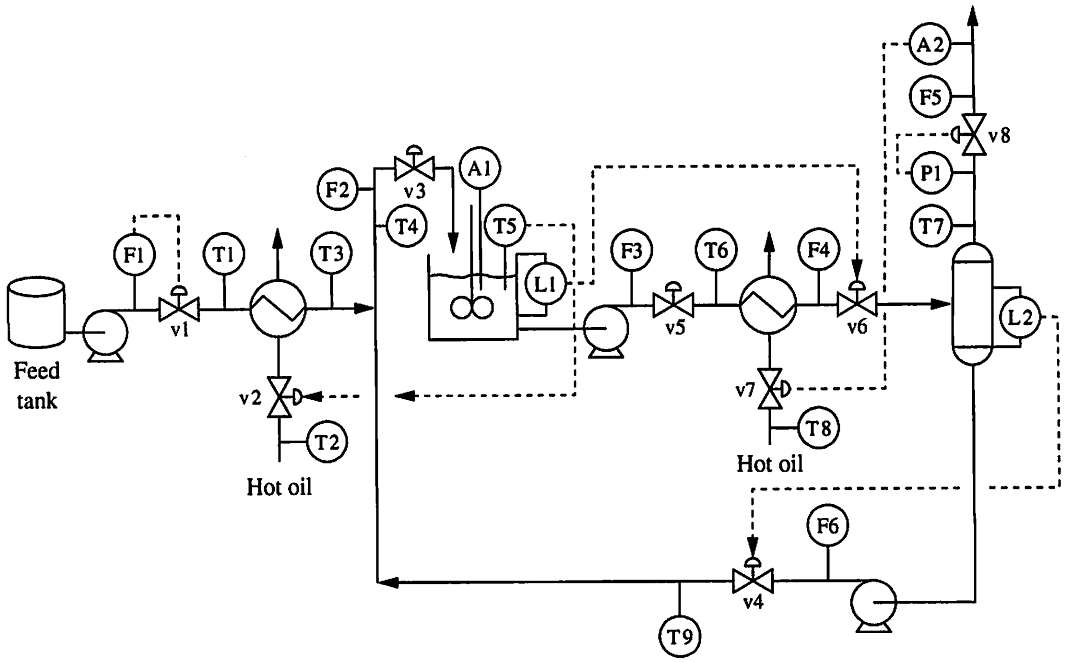


FIGURE 25.7

Control Design I.

T_1 could be a feedforward variable, and the feed temperature T_3 could be a secondary cascade variable. As a preliminary decision, the single-loop design $T_5 \rightarrow v_2$ is chosen with a PI algorithm. The resulting control of F_1, T_5, and A_2 is controllable, as can be verified using the gains in equation (25.2).

2. Design II, shown in Figure 25.8, controls the reactor composition A_1. A more direct measure of the reactor operation is the concentration of A, which can be controlled by adjusting valve v_2, although with slow dynamics. Therefore, the cascade design $A_1 \rightarrow T_4 \rightarrow v_2$ is selected, which gives good responses to temperature disturbances. The resulting control of F_1, A_1, and A_2 is controllable, as can be verified using the gains in equation (25.2).

Since no objectives have been stated for optimization, no further design decisions are needed at the fifth step in the sequence. Also, all manipulated variables have been allocated to control loops, except for v_3 and v_6, which will be held constant. Thus, no further degrees of freedom remain for adjustment.

Some control strategies would be required to ensure safe operation. The enclosed flash drum requires a reliable method for venting on high pressure, and a safety valve must be provided. Also, the objective of preventing an overflow from the reactor could require a safety interlock system (SIS) to stop the feed flow if a high level is detected. If this feature is included, an alternative disposal for the liquid from the flash drum must be provided. The safety controls are not shown in Figures 25.7 and 25.8.

Monitoring and Diagnosis

All processes should be monitored for short-term operation and longer-term performance diagnostics. Shorter-term issues involve alarms for critical variables such

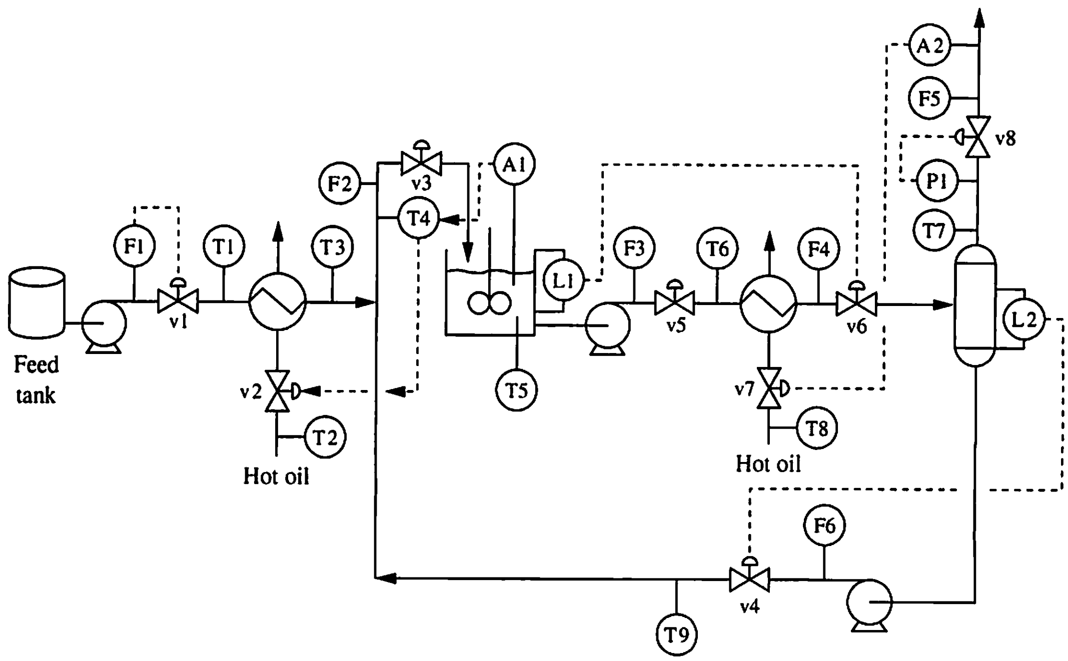


FIGURE 25.8

Control Design II.

as the liquid levels and the flash drum pressure. Some of the longer-term issues involve the reaction rate, which is influenced by impurities in the feed; recognition of poor feed characteristics would enable the engineer to trace the cause of the poor feed and take actions to prevent recurrence of such conditions. To monitor the product rate, the flow measurement F_5 should be accurate. If the density of the stream changes significantly, the conversion of sensor signal to the flow rate should be corrected based on a real-time sensor or on laboratory data of density. Another monitoring goal would involve the performance of the heat exchangers, which might foul over time. The measurements of the flows, temperatures, and valve positions enable some monitoring; for example, if the hot oil valve position increases over time at relatively constant production rate, the heat exchanger is most likely fouling. The lack of hot oil flow measurements prevents a complete check on the data; thus, the addition of flow and temperature sensors might be appropriate so that heat transfer coefficients can be calculated.

Evaluating the Designs

Designs I and II are now complete. To evaluate their performances and select a final design, the dynamic performance of the process with each design was determined. In both cases, the process begins at the same initial steady state and is subjected to a change in feed impurity, which inhibits the reaction by reducing the reaction rate (frequency factor) to 90% of its base-case value.

DESIGN I. The response of Design I is shown in Figure 25.9. The product composition (A_2) and the product flow rate (F_5) experience only small deviations

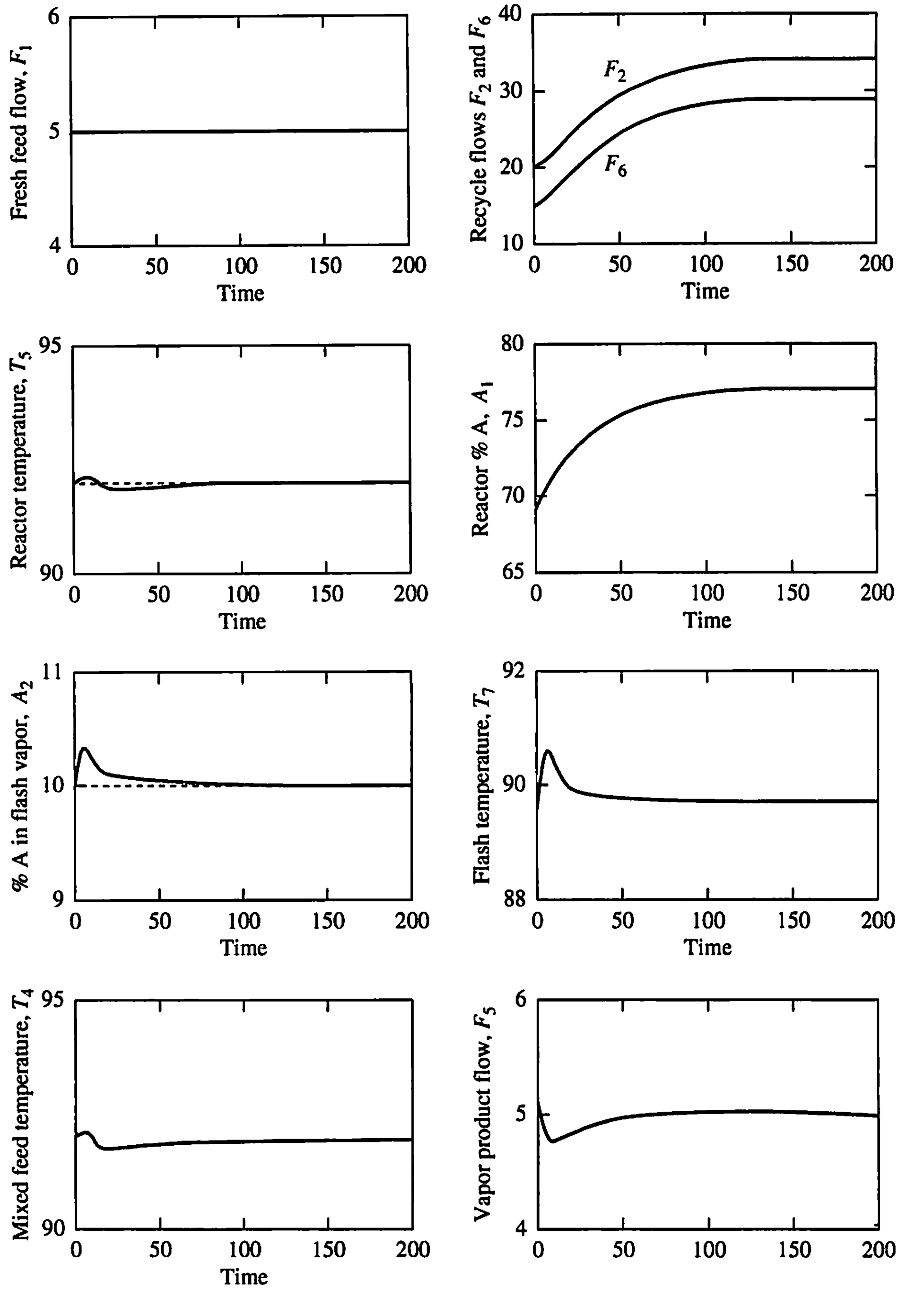


FIGURE 25.9

Transient response to feed impurity disturbance for Design I.

and return quickly to their set points. In spite of the good behavior of these key variables, other variables experience large variations; notice that the recycle flow rate changes dramatically. For this case, the reaction rate disturbance of only 10% requires recycle flow changes of about 75% to achieve a new steady state.

The reason for this large change can be understood by analyzing the dynamic behavior of the total amount of reactant in this recycle system. The amount of reactant A entering the plant is constant because the fresh feed flow is unchanged; also the percent of reactant in the product leaving the plant is controlled. To achieve a new steady-state (i.e., for the system to be self-regulating), the rate of reaction of A must return to its original valve. The rate of reaction of A is $-r_A = Vk_0e^{-E/RT}C_A$ for this continuous-flow stirred-tank reactor. Since the reactor temperature and volume are maintained at their constant set points (in the steady state), the concentration of the reactant must *increase* to compensate for the decrease in k_0 caused by the impurity. As the recycle flow (basically, unreacted A) increases, the single-pass conversion decreases because of the lower space time in the reactor (Fogler, 1997). As the single-pass conversion decreases, the concentration of A, C_A in the reactor increases, and the rate of reaction increases. Ultimately, a new steady-state operation is attained; thus, the amount of reactant in the plant is self-regulating.

Because of the low "single-pass" conversion in the reactor, a large recycle flow rate change accompanies the change in concentration. While this behavior does not negatively influence the product quality or rate, it will require a more expensive plant design. For successful operation, the process equipment, heat exchangers, pumps, pipes, and valves would have to have very large capacities, and the plant design would be costly. The general potential for recycle systems to be highly sensitive to small disturbances has been termed the "snowball" effect by Luyben et al. (1998), who point out that this is fundamentally a steady-state effect.

DESIGN II. The response of Design II is shown in Figure 25.10. Again, the product composition (A_2) and the product flow rate (F_5) experience only small deviations and return quickly to their set points. As discussed above for Design I, the accumulation of reactant A must reach zero for the plant to achieve a new steady state. Also, the flows of A in and out are identical for both the original and final steady states. Therefore, the reaction rates for the original and final operations must be the same. In Design II, the analyzer controller A_1 senses a change in concentration and adjusts the feed preheat (effectively *changing* the reactor temperature) to control the concentration.

After a transient, the process returns to nearly the same flow rates, with the reactor concentration and volume at their initial values. To return the concentration to its set point, the A_1 controller increased the reactor temperature, thus maintaining the production rate of B constant. This response returns to steady state faster, satisfies all performance objectives for F_5 and A_2, and does not require excessive equipment capacity. Based on this analysis, Design II provides better performance for the feed impurity disturbance.

Control Design II should be evaluated for all disturbances in the CDF; these others are discussed briefly here but not plotted. Because of the T_4 temperature controller, it performs well for the +20°C disturbance in T_2, with only very small deviations in the compositions and product flow. The system experiences a rather large, but brief, disturbance when T_8 increases in a step of 20°C. The maximum allowable short-term variations in the product flow F_5 and the product composition A_2 are reached or slightly exceeded. If plant experience indicated that this disturbance occurred frequently, a feedforward compensation for changes in T_8, adjusting v_7 could be added to Design II. Finally, the response of a change in desired production rate, F_5, is rather sluggish, because the feed flow rate is manipulated manually, and the product increases slowly as the recycle system responds, finally

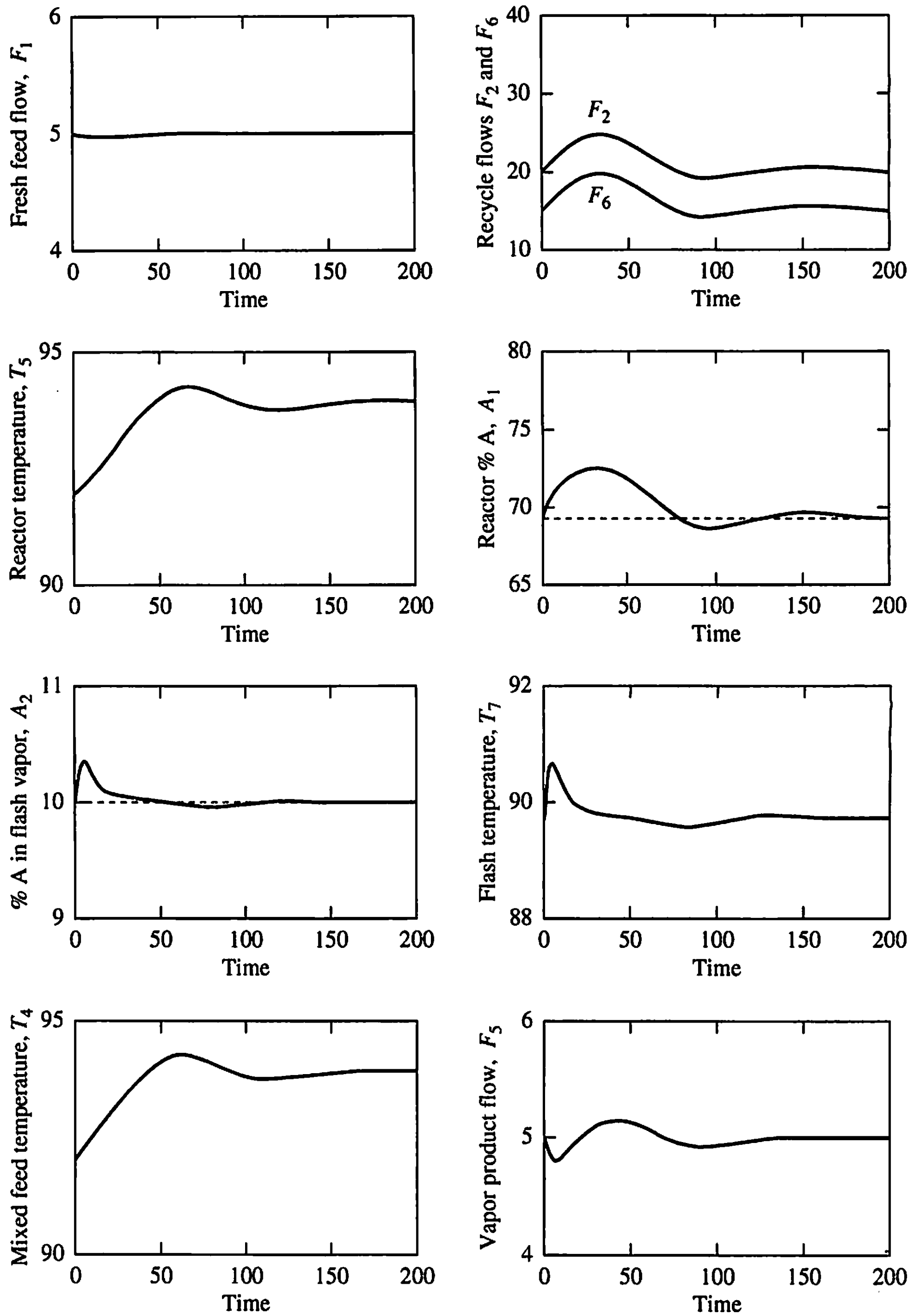


FIGURE 25.10

Transient response to feed impurity disturbance for Design II.

attaining steady state. This is a direct result of the problem definition, because short-term variation in the product rate was stated to have negligible influence on the process performance in the CDF.

The IAE for the product quality variable (A_2) is 7.11 for Design I and 6.62 for Design II for the feed impurity disturbance.

Since Design II has good performance for the key quality variable, has well-behaved dynamics for all variables, satisfies the control objectives, and requires equipment with smaller capacities, it is selected as the better control design for this process.

In performing this analysis, process decomposition was not employed, because of the strong integration, but temporal decomposition was helpful. The conclusion from this section is that the control design procedure was useful in ensuring that all important issues were considered, decisions were made in a reasonable order, and a good control design was completed. Other paths could have led to the same design, but proper shortcuts involve a very quick analysis of the factors covered in this procedure; shortcuts do not involve ignoring potentially important factors. Therefore, using the design procedure builds discipline and competence, enabling the engineer to reach proper decisions in a less time-consuming manner.

25.8 SUMMARY OF KEY DESIGN GUIDELINES

Many useful guidelines have been developed in the preceding chapters for making control design decisions based on fundamental principles. Some of the more important and straightforward are summarized in this section. Before proceeding to the summary, the concept of control performance is reiterated. Here, control performance is defined with respect to the realistic situation of a nonlinear process with changing operating conditions; thus, a nominal linear model of the process used in analysis and tuning cannot be exact, and robustness under likely model uncertainty must be considered. The behavior of all process variables must be considered; this includes the controlled and manipulated variables and may include other "associated" variables, which may become limiting when they deviate too far from normal operation. Also, the possibility of noisy measurements must be considered in estimating performance. Finally, the performance must satisfy the requirements of the plant; thus, certain variables may have overriding influence on safety, product quality, and profit. Therefore, a simple summation of the IAE for all controlled variables often does not represent the process performance. Some controlled variables may be maintained close to their set points, at the expense of others experiencing large transient deviations from their set points. This rich definition of control performance increases the difficulty of the design task, but it represents the realistic situation in most commercial enterprises. All information required to define control performance over specific operating scenarios is reported in the control design form.

The design procedure in Table 25.6 would generally encounter the decisions in the following order.

1. **Degrees of freedom.** A model of the system must have zero degrees of freedom when all external inputs are specified; this is simply requiring the model to be correctly formulated. The number of external manipulated variables (i.e., final elements) must be greater than or equal to the number of variables to

be controlled. Recall that the degrees of freedom must be evaluated using the dynamic model of the process.

2. **Controlled variables.** The engineer next decides which variables are to be measured and controlled in real time. In general, the best designs will use sensors to provide measurements of the variables whose behavior is closely related to the control objectives. This goal is usually possible for flows, pressures, temperatures, and levels. In addition, onstream analyzers can provide measurements of a limited number and type of compositions and physical properties. In many instances, a large number of components exist in product streams and many properties are important for product quality and profit. Even if all of these could be measured, which is not usual, a sufficient number of manipulated variables does not exist. Therefore, the principles of *partial control* are often employed. An inferential variable can be used as a surrogate for unmeasured properties, and a subset of important measured or inferred variables is selected to be controlled. For successful partial control, the *dominant variable(s)* selected should result in all key variables remaining within acceptable limits as disturbances occur.
3. **Operating window.** This is the range of values of process variables for which the steady-state plant operation is acceptable (i.e., physically possible and within safety and product quality limits); it is also referred to as the *feasible operating region.* The window and operating points are typically evaluated using a nonlinear, steady-state model of the process. One or several operating points may be selected within the window to give good plant performance. If a process output variable appears at or near a constraint (frame) of the window, it should be controlled to prevent violations of the limit. If a manipulated variable appears at a constraint (frame), it should be maintained near the limiting value, if possible. Normally, the plant conditions have to be moved "inside" the window, or off the frame, to ensure that no violations occur during operation with disturbances. When important variables change from internal to on a constraint as conditions change, the engineer should anticipate the need for variable-structure control methods.
4. **Interaction and integrity.** The relative gain provides one measure of process interaction. It has limitations since it only represents steady-state behavior and does not indicate strong one-way interaction, but when interpreted properly, it gives useful information. Specifically, pairing control loops which involve negative relative gains result in *poor integrity*, i.e., systems whose stability depends on the manual/automatic status of the loops; thus, designs with such pairings are selected only rarely. Also, pairing on loops with zero relative gains results in systems whose proper functioning depends on the status of many loops, also representing poor integrity. Pairing on zero relative gains is to be avoided, but it may be done if it provides a substantial improvement in control performance. Finally, control designs with loop pairings on relative gains near 1.0 suggest that the PI multiloop tuning should not change significantly between single-loop and multiloop.
5. **Interaction and performance.** The performance of multiloop control systems depends on the type, or direction, of the disturbance. The relative disturbance gain, RDG, was introduced as an approximate indication of whether the inter-

action is favorable or unfavorable. Designs with the product of the disturbance gain (K_d) and relative disturbance gain (RDG) of small magnitude for important controlled variables are generally favored, although evaluation of the dynamics is warranted before final selection.

6. **Feedback process dynamics.** Generally, feedback control performs well when the dynamics in the feedback path are fast, with a short dead time. Also, inverse responses were shown to degrade control performance, and, because multivariable control systems have a parallel structure, the closed-loop systems can experience inverse responses even though each individual input-output dynamic response does not. Improved control performance can be achieved in many cases by selecting from a suite of enhancements that improve dynamic performance, such as cascade, feedforward, adaptive tuning, and process modifications that reduce the feedback dynamics, such as a partial bypass around a heat exchanger.

   Processes that are open-loop stable are preferred. Non-self-regulating levels and all pressures in closed vessels are noted for feedback control when reviewing a process. Also, processes that have significant inherent positive feedback should be evaluated to be sure that they are open-loop stable; if unstable, efforts should be made to modify the process design.

   Processes *with recycle* deserve special attention because of the possibility of positive feedback. When reactants are recovered and returned to a chemical reactor, the possibility of poor self-regulation or instability exists. The control system should be designed to ensure that neither reactants nor inerts accumulate in the system without limit.

7. **Disturbance dynamics.** Additional steps can be taken to reduce the effect of the disturbance; the best action is to eliminate it at the source. Other steps include feedforward control, inventory sizing, and averaging level control, to modulate the rate of change in flow properties, and process operating condition changes, to reduce the sensitivity to a selected disturbance. For multiloop control, the influence of interaction is reduced when interaction dynamics are much slower than the "direct" feedback path; when unfavorable interaction exists, the interaction should be slowed by process equipment modifications and controller detuning.

8. **Tuning guidance.** The control design and tuning should be selected concurrently. For example, certain levels may require averaging or tight level control, and interacting loops should be tuned to increase favorable interaction and minimize unfavorable interaction. These requirements should be documented as part of the control design; later implementation that does not adhere to the proper tuning is likely to be unsuccessful.

The methods used for the control design procedure involve a hierarchical analysis, in which the initial steps establish the feasibility of achieving the desired performance with the process and control designs. These initial evaluations are selected using "open-loop indicators" (Barton et al., 1991), which depend solely on the process and are independent of the control structure, algorithms, and tuning. The operating window, controllability, integral controllability, and relative gain are in this category. In these steps, many inappropriate design candidates are eliminated; also, many insights into the possible strengths and weaknesses in the remaining candidates are developed. Note that most of these evaluations can be based on steady-state models.

For final design of the process and selection of the best control design, the dynamic behavior of the closed-loop system *must* be considered. For example, Skogestad et al. (1990) demonstrate that reliance solely on steady-state analysis can result in the best control design being eliminated from consideration in distillation control. Further, a straightforward example of the importance of dynamics is the pairing of an important controlled variable with a manipulated variable that gives fast feedback dynamics; this can even lead to pairing on a zero relative gain, in extreme situations. In general, the behavior of multiloop systems can be quite complex, with poor designs yielding inverse response even when the process dynamics are well behaved (see Example 21.4). The frequency-dependent relative gain was briefly introduced to evaluate complex interactions, but the best approach is to simulate the final selection(s) to ensure good dynamic behavior. The use of nonlinear dynamic models for this final evaluation provides additional checks on the approximations inherent in the linear analysis methods used at earlier steps in the evaluation.

25.9 ▫ CONCLUSIONS

While no new technology was presented in this chapter, very important methods for managing the design procedure were presented. They enable the engineer to utilize information fully and effectively, to recognize when the problem is or is not fully defined, to apply the simplest decision methods at each stage, and to conclude the design procedure with high probability of success.

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Luyben, W., B. Tyreus, and M. Luyben, *Plantwide Process Control*, McGraw-Hill, New York, 1998.

Skogestad, S., P. Lundstrom, and M. Morari, "Selecting the Best Distillation Control Configuration," *AIChE J., 36,* 753–764 (1990).

ADDITIONAL RESOURCES

Considerable material (including books, videos, and short courses) on HAZOP analysis is available from the Center for Chemical Process Safety of the American Institute of Chemical Engineering.

The concepts in the recycle example have been expanded upon in the following set of publications.

Luyben, W., "Dynamics and Control of Recycle Systems, 1. Simple Open-Loop and Closed-Loop Systems," *IEC Res. 32,* 466–475 (1993); "2. Comparison of Alternative Process Designs," *32,* 476–486 (1993); "3. Alternative Process Designs in a Ternary System," *32,* 1142–1153 (1993); Tyreus, B., and W. Luyben, "4. Ternary Systems with One or Two Recycle Streams," *32,* 1154–1162 (1993).

Additional checklists that can be useful in developing ideas for the control design form can be found in

Marlin, T., J. Perkins, G. Barton, and M. Brisk, *Advanced Process Control Applications, Opportunities and Benefits,* Instrument Society of America, Research Triangle Park, NC, 1987.

The analysis approaches in the last few chapters are complemented by references giving the practice of control design for specific process units. A few examples are cited in Chapter 1 and below.

Balchen, J., and K. Mumme, *Process Control Structures and Applications,* Van Nostrand Reinhold, New York, 1988.

Baur, P., "Combustion Control and Burner Management," *Power, 126,* S-1 to S-16 (1982).

Duckelow, S., *The Control of Boilers* (2nd ed.), Instrument Society of America, Research Triangle Park, NC, 1991.

Liptak, B., "Optimizing Controls for Chillers and Heat Pumps," *Chem. Engr., 90,* 40–51 (October 1983).

Luyben, W. (ed.), *Practical Distillation Control,* Van Nostrand Reinhold, New York, 1992.

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Shinskey, F., *Energy Conservation through Control,* Academic Press, New York, 1978.

Starolesky, N., and L. Ladin, "Improved Surge Control for Centrifugal Compressors," *Chem. Engr., 86,* 175–184 (May 1979).

TAPPI, *Process Control Fundamentals for the Pulp and Paper Industry*, TAPPI Press, Atlanta, 1995.

Leading research and practice in process control relating to process and control system design is presented in many technical meetings, including the following, which are organized periodically.

International Federation of Automatic Control, *Advanced Process Control in the Chemical Industries* (ADCHEM).
International Federation of Automatic Control, *Symposium Series on Dynamics and Control of Chemical Reactors, Distillation Columns, and Batch Processes* (DYCORD).
International Federation of Automatic Control, *Workshop on the Interaction between Process Design and Control.*
Process Systems Engineering (PSE).

> The procedures introduced in this chapter are applied using technology presented throughout the book. Questions for testing your learning are located at the end of Chapter 24. The questions at the end of Chapters 13 and 21 should also provide useful exercises. A few questions are given here that relate to the methods and examples introduced in this chapter.

QUESTIONS

25.1. Answer the following questions on the reactor with separator process.

(*a*) Verify that selected controlled variables in Designs I and II can be controlled with the selected manipulated variables.

(*b*) Check each of the Designs (I and II) to determine whether it is integral-stabilizable.

(*c*) Evaluate the relative gains for the two designs and discuss the implications.

(*d*) Demonstrate that flows F_1 and F_6 can be controlled with v_1, v_3, and v_4. Discuss reasons for selecting two of these three valves.

25.2. Discuss the performance of Designs I and II and propose better alternative designs, if possible, for the following situations. Each situation is to be considered separately, not cumulatively.

(*a*) The reactor temperature, T_5, must be maintained constant to obtain the best product selectivity. Is there an alternative reactor environment variable that can be adjusted? If yes, design a control strategy to meet the objectives.

(*b*) The analyzer for the reactor concentration is quite expensive. Is there another variable that can be used in its place?

(*c*) The control objectives are changed to include tight control of the product flow rate F_5. The disturbances are unchanged. How should the control strategy be changed?

(*d*) The daily total production of product B must be satisfied as close to its target as possible. How can the design be modified to satisfy this requirement?

(*e*) The recycle pump has been replaced with a spare pump of smaller capacity. Modify the control design to produce as much product as possible.

25.3. Using the checklists in Section 25.2, prepare control design forms for the following processes. You should note information that you would need to determine from the plant personnel to complete the form.

(*a*) The distillation process in Examples 5.4, 20.2, 20.4, 20.5, and many examples in Chapter 21.

(*b*) The fired heater in Figure 15.17.

(*c*) The boiler in Figure 2.6.

(*d*) The gas distribution network in Question 24.15.

25.4. A series of processes is represented by the simplified system of flows and inventories in Figure Q25.4. Design a variable structure control system that will maximize the throughput while maintaining all levels within their maximum and minimum limits. The constraint that determines the maximum throughput could be the maximum feed target, the maximum product flow target, or any pump-valve combination in the system. (The targets are specified by the plant personnel.)

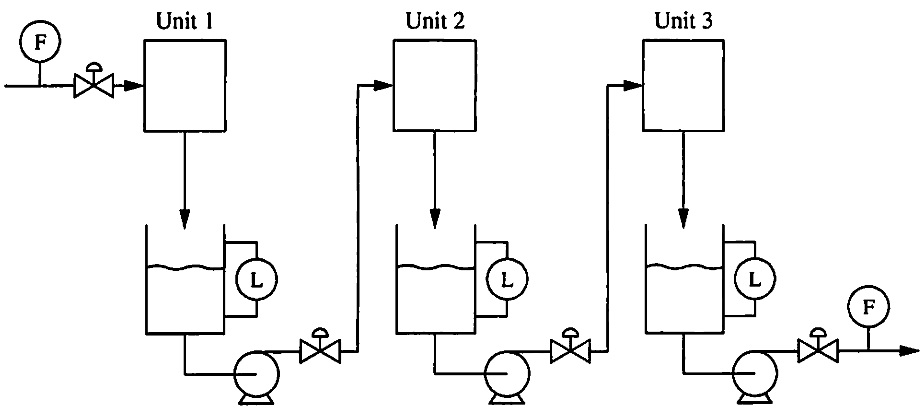

FIGURE Q25.4

25.5. An inverse response (right-half-plane zero) in the feedback process dynamics in a *single-loop* control system was analyzed in Examples I.2 and 13.8. Assume that a two-input–two-output process has monotonic step responses for each input-output relationship. Discuss whether the 2 × 2 closed-loop control system can have an inverse response in one controlled variable, and if so, under what conditions. If yes, discuss how this situation may affect the control performance of the system.

25.6. Discuss the following issues in control design.

(*a*) What is the proper design for systems with more manipulated than controlled variables?
(*b*) How does the design engineer decide at which point(s) the process should be operated within the operating window?
(*c*) Is it impossible to implement feedback control for a system that is not integral-stabilizable, as determined by the Niederlinski index?
(*d*) If a nonzero operating window exists, is the process guaranteed to be controllable within the window?
(*e*) Is it appropriate to design a multiloop control system without giving guidance on tuning the controllers?

25.7. Discuss the following issues in control for safety.

(*a*) Give examples of how control strategies for temporal levels 1 through 3 (flow to product quality) contribute to safe operation.
(*b*) Give examples of how control strategies for temporal levels 1 through 3 can negatively influence the safety of the system. For each example, give a control design decision that would ameliorate the hazard.

25.8. For each of the processes in question 25.3, determine process performance characteristics that should be monitored using real-time data. For each characteristic, define the calculations and sensors required and how the results would be interpreted, and discuss the actions taken when the process performance becomes unsatisfactory.

25.9. A major process design change is being evaluated for the reactor-with-recycle process. The stirred tank reactor can be replaced with a packed-bed reactor, as shown in Figure Q25.9. A new liquid byproduct, component C, is also produced, and it is separated from the recycle A (and B) in a liquid-liquid separator. Sketch a control system design for this process in the figure. You may add valves and sensors as needed.

25.10. In a monograph on plantwide process control, Luyben, Tyreus, and Luyben (1998) discuss the potential accumulation of reactants in reactor-recycle systems. They suggest that one flow rate in the recycle loop should be on flow control, not adjusted by a level controller.

(*a*) Discuss the rationale for this suggestion.
(*b*) Apply this suggestion to the solved example in this chapter (both Designs I and II), and sketch the control designs on copies of Figure 25.6.
(*c*) Discuss the expected performance for the disturbances defined in the control design form.

25.11. In some reaction systems, adjusting temperature can be inappropriate. For the reaction sequence A $\rightarrow$ B $\rightarrow$ C with B the desired product, high temperature might lead to the production of excessive amounts of undesired byproduct C. Answer the following questions for both Designs I and II of the solved example in this chapter.

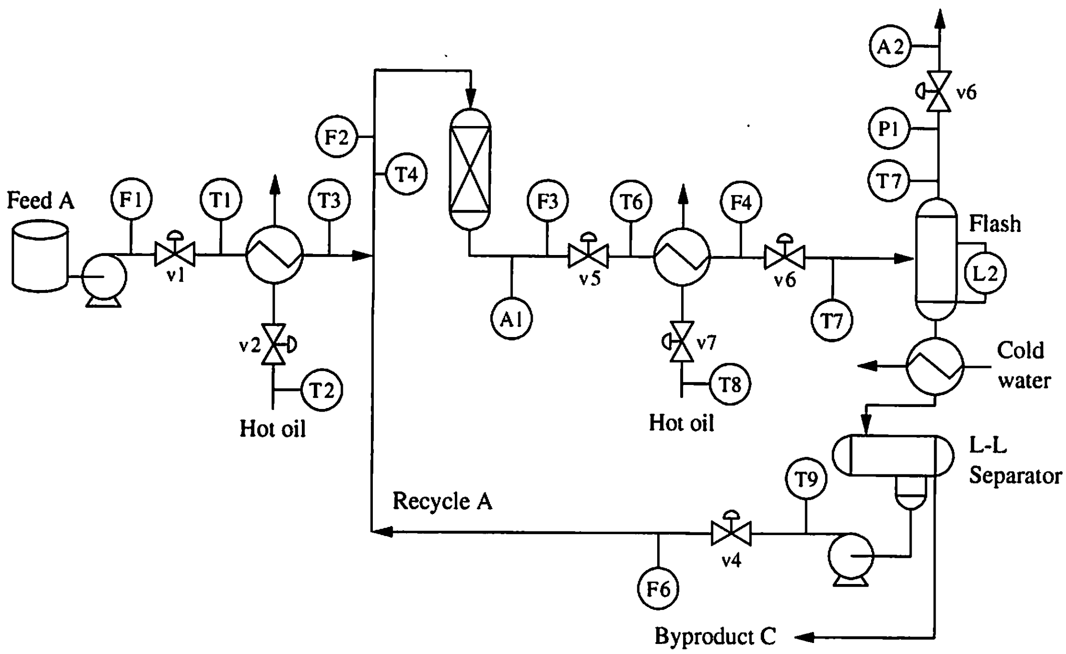

FIGURE Q25.9

(*a*) Discuss all possible dominant variables for the reactor.
(*b*) Select a dominant variable different from the temperature or concentration and sketch the complete design on a copy of Figure 25.6.
(*c*) Discuss the response of the new control design to the disturbances in the control design form.

25.12. The control design form for the worked example in this chapter specified that the product flow rate could deviate from its desired value. Consider a modified problem that requires closer control of the product flow to its set point.

(*a*) Without changing any of the existing controllers in Design I, add one or more controllers to improve the control of the product flow rate. Discuss the performance that you would expect from your new design.
(*b*) Without changing any of the existing controllers in Design II, add one or more controllers to improve the control of the product flow rate. Discuss the performance that you would expect from your new design.
(*c*) Develop a new control design that provides very tight control of the product flow rate, while also achieving the other control objectives.

25.13. The dynamic behavior in the worked example in this chapter was strongly influenced by the material recycle. Consider a modified process without recycle; two feeds are mixed before entering the reactor, and the liquid from the flash separator goes to a tank. (This would be approximately how the plant operated if a very large tank existed in the recycle path.)

(*a*) Consider the performance of Design I for this modified plant.
(*b*) Consider the performance of Design II for this modified plant.
(*c*) Design an improved control system for this modified plant.

25.14. The reactor with recycle process could be modified to achieve better recovery in the separation by replacing the flash drum with a two-product distillation tower. In the modified process, the overhead product would be a vapor stream of mostly component B, and the bottoms product would be liquid recycle to the reactor of mostly component A. Sketch the process and add sensors, valves, and controllers to yield good control performance for the integrated product. You may assume that the separation of the two components can be characterized by a constant relative volatility.