Process Control Design: Definition and Decisions

CHAPTER 24

24.1 INTRODUCTION

Typically, the starting point for control system design and analysis is a preliminary process design, perhaps with some initial control loops, along with a specification of the desired process performance. This amount of initial information is realistic for existing plants, because the equipment is already in operation when an analysis to improve plant performance is carried out. It is also realistic for new plant designs, because a preliminary process structure (or alternative structures) must be available when dynamics and control are first analyzed.

The required information must be recorded concisely, and the control design form described in the next section is proposed as a format for this record. A great advantage for using this form, in addition to giving excellent documentation, is that it provides a way to begin the design analysis. Often, the design problem seems so big and ill-defined that an engineer, especially one new to the technology, is unsure where to begin. By completing the thorough definition, the engineer begins the problem-solving process, and important issues and potential solutions become apparent.

Potential actions required to achieve the desired process performance include (1) defining the control strategy designs, (2) selecting measured variables and instrumentation (i.e., sensors and final elements), (3) specifying the process operating conditions, (4) making minor process changes such as adding a bypass, selecting an alternative manipulated variable, or changing the capacity of some equipment, or (5) making major process structure changes, such as changing from a packed- to fluid-bed reactor. The fifth possibility, involving major process alterations, is excluded from this discussion, because such a major decision would require an analysis of the steady-state and dynamic behavior of an integrated plant involving many units, which is beyond the scope of this book.

The six major categories of decisions made during the design procedure follow in the order covered in this chapter.

- *Measurements:* selecting measured variables and sensors
- *Final elements:* providing final elements with features contributing to good control performance
- *Process operability:* providing good steady-state and dynamic behavior that enables the control performance objectives to be achieved
- *Control structure:* providing the proper interconnection of measured and controlled variables via the control system
- *Control algorithms:* selecting and tuning the proper algorithms for feedback and feedforward control
- *Performance monitoring:* providing measurements and calculations for monitoring and diagnosing the process and control performance

The application of previously introduced technology to achieve a control design is explained in this chapter. All key elements of control design are demonstrated through application to an example design, which is introduced in the following section.

24.2 ☐ DEFINING THE DESIGN PROBLEM

The first step in the design task is the definition of the "problem," which perhaps should be referred to as an *opportunity* to apply our skills. We will retain the term *problem* because it is used commonly to describe the task of addressing complicated issues (e.g., "problem solving"). A complete definition of the design problem may be difficult in the beginning of the analysis, and the need for additional information may become apparent as the problem is analyzed. Therefore, the approach taken here is to provide a comprehensive form in which information can be recorded. The use of a form has several advantages. First, it serves as a convenient checklist so that the engineer is sure to address the important issues at the definition stage. Second, it provides a coherent, readable statement of goals, which can be reviewed by many members of a design team. Third, a form with topics concisely addressed under clear headings provides a structure that is easy to write and to use as a reference. Finally, additional information developed during the design analysis can be added at any time to the original form.

The form used here is referred to as the control design form (CDF). It will be introduced by discussing the initial draft in Table 24.1 for the proposed flash process shown in Figure 24.1. The feed composition, flash temperature and pressure, and product compositions are identical to the example in Section 17.2, and the base-case values for all measured variables are reported in the "Measurements" section of Table 24.1. Note that the equipment in Figure 24.1 may be *incomplete and contain errors*. The control design for this process will be discussed as each major control decision category is introduced in this chapter, and a complete, error-free design will be developed by the end of this chapter.

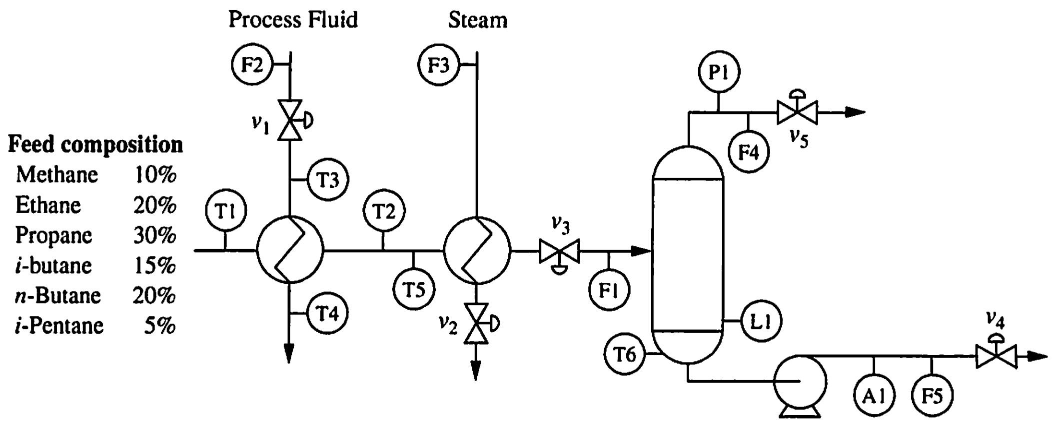


FIGURE 24.1

Preliminary process and instrumentation for the flash process.

As is typical in problem solving, we will start with a definition of the control objectives in the first major heading of the CDF. The control objectives are combined into the seven categories introduced in Chapter 2. The entries in each category must be concise but complete enough to provide the direction for the remaining design decisions. It is especially important to be as quantitative as possible regarding the performance, giving performance criteria for specific scenarios. This type of specification provides the basis for the design, along with a way to test the performance of the design against the objectives. Remember that the control performance should be specified for particular operating conditions and time periods; for example, (1) selected variables must remain within deviation limits from set point for a specified step disturbance; (2) the standard deviation for a variable must be no greater than specified over a day, week, or other interval; (3) a variable may not exceed its limits more than once per day; or (4) very undesirable conditions should not occur "under (essentially) any (conceivable) circumstances." Additional examples are given in Table 24.1 for the flash process.

The second heading contains information on the measurements provided for the control and monitoring system, which are crucial to the success of process control. The location of the sensor is shown in an accompanying drawing (i.e., Figure 24.1), and the physical principle of the sensor and range are given in the CDF. Special features of a sensor, such as the update frequency for a discrete sensor like a chromatograph, should also be recorded.

The final control elements are recorded under the third heading. The maximum capacity of the manipulated variable, typically the maximum flow through a valve, should be noted. Also, nonstandard features should be noted; for example, tight shutoff (i.e., the ability to prevent all flow); a valve that can open quickly; or a final element that has a restricted range (e.g., cannot be closed). The failure mode of the final element is important but is not recorded here, because it is usually indicated on the drawing.

The fourth heading provides a place to document important limitations that could affect the control design. These are typically constraints on equipment and process variables. The limiting values and whether the constraint can be measured, along with the sensor type, should be recorded. The information should clearly

TABLE 24.1

Preliminary control design form for the flash process in Figure 24.1

TITLE: Flash drum	ORGANIZATION: McMaster Chemical Engineering
PROCESS UNIT: Hamilton chemical plant	DESIGNER: I. M. Learning
DRAWING: Figure 24.1	ORIGINAL DATE: January 1, 1994

Control Objectives

1. Safety of personnel
   (*a*) The maximum pressure of 1200 kPa must not be exceeded under any (conceivable) circumstances.
2. Environmental protection
   (*a*) Material must not be vented to the atmosphere under any circumstances.
3. Equipment protection
   (*a*) The flow through the pump should always be greater than or equal to a minimum.
4. Smooth, easy operation
   (*a*) Control all unstable variables (liquid level)
   (*b*) All process variables should remain within reasonable ranges without undue operator actions
   (*c*) One variable should control the production rate with little variation
   (*d*) Control loops should function well independent of manual/automatic status of other loops; that is, the system should have good integrity
5. Product quality
   (*a*) The steady-state value of the ethane in the liquid product should be maintained at its target of 10 mole% for steady-state operating condition changes of
      (i) +20 to −25% feed flow
      (ii) 5 mole% changes in the ethane and propane in the feed
      (iii) −10 to +50°C in the feed temperature
   (*b*) The ethane in the liquid product should not deviate more than ±1 mole% from its set point during transient responses for the following disturbances:
      (i) The feed temperature experiences a step from 0 to 30°C.
      (ii) The feed composition experiences steps of +5 mole% ethane and −5 mole% propane.
      (iii) The feed flow set point changes 5% in a step.
6. Efficiency and optimization
   (*a*) The heat transferred should be maximized from the process integration exchanger before using the more expensive steam utility exchanger.
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	Nominal value and sensor range	Special information
A1	Chromatograph	10, 0–15 mole%	Update every 2 minutes
F1	Orifice	100, 0–200	
F2	Orifice	120, 0–150	
F3	Orifice	100, 0–200	
F4	Orifice	45, 0–90	
F5	Orifice	55, 0–110	
L1	Δ pressure	Range is lower half of drum	
P1	Piezoelectric	5000–15000 kPa	

TABLE 24.1

Continued

Measurements

Variable	Sensor principle	Nominal value and sensor range	Special information
T1	Thermocouple	0, (−)50–100 °C	
T2	Thermocouple	25, 0–100 °C	
T3	Thermocouple	90, 0–200 °C	
T4	Thermocouple	45, 0–200 °C	
T5	Thermocouple	25, 0–100 °C	
T6	Thermocouple	25, 0–100 °C	

Manipulated variables

I.D.	Maximum capacity (at design pressures)
	(%open, maximum flow)
v1	100%, 100
v2	53%, 189
v3	50%, 200
v4	14%, 340
v5	52%, 106

Constraints

Variable	Limit values	Measured/ inferred	Hard/ soft	Penalty for violation
Drum pressure	1200 kPa, high	P1, measured	Hard	Personnel injury
Drum level	15%, low	L1, measured	Hard	Pump damage
Ethane in F5 product	±1 mole% (max deviation)	A1, measured, and T6, inferred	Soft	Reduced selectivity in downstream reactor

Disturbances

Source	Magnitude	Dynamics
Feed temperature (T_1)	−10 to 55°C	Infrequent step changes of 20°C magnitude
Feed rate (F_1)	70 to 180	Set point changes of 5% at one time
Feed composition	±5 mole% feed ethane	Frequent step changes (every 1 to 3 h)

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

Input	Output	Gain	Dynamic model
v1	{see Example 24.6}		
v2			
v3			
v4			
v5			

Additional considerations

Liquid should not exit the drum via the vapor line.

indicate whether the constraint is soft or hard, along with the penalty for exceeding the constraint (e.g., yield loss, energy consumption, or equipment damage). A soft constraint can be violated for a short time and thus does not require the process to be shut down when the constraint is approached. An example would be a stream of material that, when not observing quality specifications, can be recycled or diverted to waste. Naturally, this is to be avoided but can be tolerated. The violation of a hard constraint causes severe safety or environmental hazards or costly equipment damage. Thus, a hard constraint must not be violated, and extreme measures, such as shutting down the process, are appropriate when a hard constraint is approached too closely.

Since the main reason for control is to respond to input changes (disturbances and set points), proper design depends on a good definition of these changes, which are recorded under the fifth heading. Recall that the importance of disturbances was recognized and included in methods presented in previous chapters, such as cascade, feedforward, gain scheduling, inferential control, and multiloop pairing. Therefore, each source of disturbance should be identified, along with its frequency of occurrence and magnitude; this information is useful in evaluating the potential need for and success of various design options. If the disturbance can be measured, that should be noted for possible feedforward and gain scheduling control.

The sixth heading covers dynamic responses between all process inputs (disturbances and manipulated variables), and all outputs (controlled variables and constraints). Naturally, this information is essential for control design. The models at the design stage might be very qualitative (fast, slow), semiquantitative (dominant time constants), or reasonably accurate (transfer function). The level of modelling performed should match the accuracy required for the decisions made during the control design; this design step might be less demanding than control implementation, which can be based on empirical models when the controllers are tuned.

The seventh and final heading provides a location for special information that does not fit under the other headings. For example, perhaps a particular flow should not be adjusted rapidly because of the sensitivity of product quality to flow rate. These special items, which require sound chemical engineering analysis of the process, must be considered in process control design.

This form may seem a bit pedantic, requiring excessive documentation for every decision; in fact, most control designs are performed in practice without such extensive documentation. The form is used here because it provides an excellent structure for beginning engineers who, after gaining proficiency, will often be able to perform the analysis without the form. However, even the most experienced engineers benefit from this type of documentation for complex designs. It is important to recognize:

> Experienced engineers can sometimes bypass the control design form (CDF) documentation, but they must perform a thorough analysis involving all information and issues included in the CDF.

An excellent example of a control design definition is given in the "Tennessee Eastman Industrial Challenge Problem" (Downs and Vogel, 1993). While not organized in the same format as Table 24.1, this definition contains essentially the same information and is sufficiently complete to enable independent teams to design controls and compare results. Again, the definition in this realistic industrial design is complex, and a clear, written presentation is essential.

The subsequent sections of this chapter discuss issues related to the six major design decisions made in control engineering based on the information in the CDF. During the design, the engineer may find that the initial information is not complete and may have to return to enter additional information or enhance the measurements and final elements provided. In fact, the initial performance objectives might not be achievable with the initial process equipment and disturbances, in which case the engineer must reevaluate the objectives and either relax the specifications, alter the process design, or, if possible, reduce the disturbances. Such iterations are a natural part of the design process and do not necessarily indicate poor initial definition and analysis.

24.3 MEASUREMENTS

The success of automatic process control, real-time monitoring, and long-term performance tracking in improving plant performance depends crucially on measurements. The engineer must first determine the process variables to be measured and select a sensor for each. In this section, several important issues in selecting variables and sensors are discussed.

Measurements

- Feasibility
- Accuracy
- Dynamics
- Reliability
- Cost

Measurement Feasibility

When the value of a variable is needed, it can be obtained from at least two real-time methods. First, it can be measured "directly" by a sensor; as an example, a temperature can be measured by a thermocouple, although the actual value sensed is the voltage generated for a bimetallic connection with nodes at two temperatures: the reference and process temperatures. This sensor is called *direct* because the physical principle underlying the measurement is independent of the process application, and the relationship between the sensor signal and process variable is reasonably accurate. Examples of variables that can usually be measured directly are level, pressure, temperature, and flows of many fluids. Also, the compositions and physical properties of some process streams can be determined in real time with on-stream analyzers.

In the second method, the variable cannot be measured, at least at reasonable cost, in real time, but it can be inferred using other measurements and a process-specific correlation. Inferential control is covered in detail in Chapter 17, so procedures for designing inferential variables will not be repeated here, except to emphasize that the acceptability of inferential control must be evaluated on a case-by-case basis. Examples of variables that are often inferred are composition of vapor-liquid equilibria (from temperature and pressure) and chemical reactor conversion (from temperature difference).

Not all variables can be measured or inferred in real time. These variables have to be determined through analysis of a sample of material in a laboratory. When the sample and analysis can be performed quickly, the laboratory measurement value can be used for feedback control. There are many industrial examples of controllers that use laboratory results and are executed every few hours, such as the one described by Roffel et al. (1989). While not providing control performance as good as would be possible with on-stream analysis, this approach usually gives much better performance than not using the laboratory value.

Accuracy

As explained in Chapter 12, the term *accuracy* refers to the error between the true process variable and the sensor signal. The error is a property of the sensor and, usually, its range; thus, the range should be maintained only as large as needed to measure the expected variation of the process variable about its normal operation. An associated property of the sensor is *reproducibility,* which indicates the differences in the sensor signal at different times for the same value of the true process variable. Often, sensors that provide good accuracy cost more than those that provide only good reproducibility; therefore, it is important to recognize which property is most important in a process control design and select the sensor accordingly.

For example, consider the process and control design in Figure 24.2, which includes cascade and feedforward. In determining whether accuracy or reproducibility is required, the key question is, "What is the purpose of the sensor?" For example, the objective of the feedforward controller is to adjust the manipulated variable for *changes* in the measured disturbance; therefore, it acts only on changes in the measured disturbance. In this situation, reproducibility of FI (i.e., reliable indications in the change of the disturbance variable) is more important than accuracy of the actual value. Similarly, the objective of the secondary controller in the cascade, FC, is to respond quickly to disturbances; therefore, reproducibility is again more important than accuracy. In contrast, the objective of the primary feedback controller in the cascade, A1, is to maintain the key output variable at the desired value; therefore, accuracy is required for this measurement. Analyses of

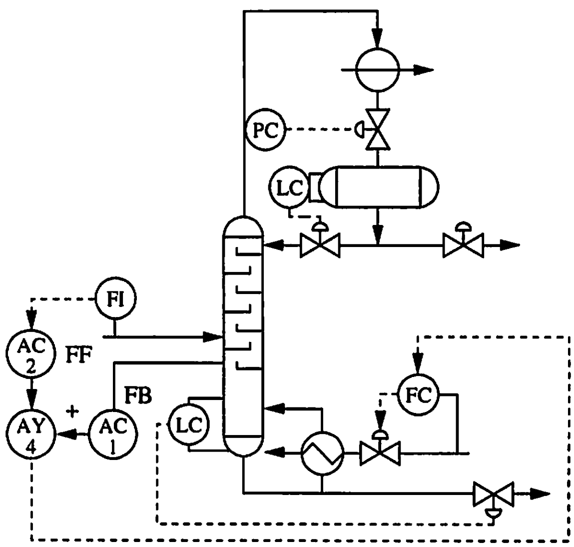


FIGURE 24.2

Example of feedforward-feedback control of a distillation tower product quality.

TABLE 24.2

Measurement objectives for various control structures

Control design	Measurement accuracy required	Only measurement reproducibility required*
Single-loop feedback	Product quality or other key variable	Tight control not important; proper set point can be adjusted infrequently by a person to attain the desired operating condition
Cascade	Primary controller	All secondary controllers
Feedforward-feedback	Measured variable for feedback controller	Measured disturbance for feedforward
Gain schedule	Measured variable used in correlation to determine tuning	
Inventory	Vapor: control must prevent violation of pressure limits for equipment	Liquid: reproducibility is acceptible if the inaccuracy is small with respect to the level range
Production rate	(1) The exact flow rate control is required or (2) The measurement is used to determine the sales volume	(1) Constant flow is important and (2) The goal is the proper average production over a day and (3) The production can be determined by accurate inventory measurement

*This is for control purposes; monitoring may require accuracy.

sensor applications yield the summary of measurement objectives in Table 24.2, which readers should verify for themselves.

Some sensors have inherent inaccuracies that, if significant for a particular application, can be compensated in the input processing phase of the controller execution. As an example, the relationship between the pressure drop across an orifice and the volumetric flow rate is given by the equation

$$F = K\sqrt{\frac{\Delta P}{\rho}} \tag{24.1}$$

When the density of the fluid is not constant, both the density (ρ) and the pressure difference across the orifice plate (ΔP) could be measured and the appropriate calculation made in the control system to yield the corrected "measured" flow rate.

Dynamics

The dynamics of the process and sensors are present in the feedback loop and therefore influence control performance. The first step to improve control is to select a location for the sensor that results in the fastest process dynamics in the feedback system. For example, the analyzer A1 in Figure 24.2 samples the vapor before the large first-order system that would have occurred if the analyzer had been downstream of the liquid inventory.

An estimate of the effect of sensor dynamics can be obtained by performing either a dynamic simulation or a frequency response analysis of the closed-loop system with and without the sensor dynamics. These analyses in Chapter 13 concluded that the sensor dynamics should be fast, certainly much faster than the process dynamics. For common flow, level, pressure, and temperature sensors, the dynamic response of the sensor is not usually a limiting factor in control performance, except for control of fast machinery systems. However, many analyzers are slow, because of (1) their sampling systems, which extract material from the process and transport it to a remote analyzer, and (2) the time for analysis. Thus, these sensors often contribute substantial dynamic delay to the closed-loop system and degrade the control performance. When this situation occurs, a common step to improve the control performance is to use a fast sensor as an inferential variable that can be reset in a cascade design by the slower analyzer controller.

Reliability

Sensors used in control systems must be very reliable, because the failure of a sensor incapacitates the control loop and could lead to an unsafe situation. For example, a failure of the reboiler flow sensor in Figure 24.2, if not identified during input processing, could result in a zero value being used as the value of the controlled variable in the controller calculation. Since the measurement would be below the set point, the controller would rapidly open the reboiler valve completely, which could cause a pressure surge that might damage the trays. Some sensor characteristics that lead to lower reliability are (1) sensors contacting process fluids, (2) poorly designed sample systems that plug or extract an unrepresentative sample, and (3) complex chemical or physical analyses (Clevett, 1986). In many designs these characteristics cannot be eliminated, and the engineer should expect lower reliability.

Cost

The cost of a sensor is the total of equipment purchase, installation, maintenance, and operating costs. Most sensors have small operating costs, perhaps a small amount of electrical power for heating in cold weather; however, a sensor can occasionally contribute substantially to plant operating costs. An example is a flow sensor for gas in a pipe, where the standard orifice meter can be used to measure the flow, but the nonrecoverable pressure drop across the orifice can be large. If compression costs are significant, a sensor that has very low pressure losses (e.g., a venturi meter or pitot tube) could be used. The purchase and installation costs of the alternative meter would be greater than for the conventional orifice, but its total cost over several years would be lower.

Finally, the primary use of the measured value should be considered in selecting a sensor. Fast dynamics would be an important concern for sensors used in feedback control. However, measurements for monitoring, especially longer-term process performance, may be satisfactorily supplied by sensors that are slower or of lower cost.

EXAMPLE 24.1.

In this example, the sensors in the preliminary flash design in Figure 24.1 are considered. First, we notice that the sensors T2 and T5 are redundant and that redundancy is not needed, because this is not a critical measurement. Therefore, sensor T5 is removed. Second, it is noticed that the feed flow measurement F1 is located after the flash valve, where the material is composed of two phases. However, the pressure drop across an orifice meter, which is the sensor principle, does not accurately or reproducibly relate to the actual flow when the fluid has two phases. Therefore, the flow meter location is moved before the first heat exchanger, where the material is always one liquid phase in this example.

Third, the temperature indicating the flash, T6, is in the liquid inventory and will not rapidly respond to changes in the drum inlet temperature. Since this temperature will be used as an inference of composition, minimum feedback dynamics is desired. Therefore, T6 is relocated in the vapor space, which has little inventory. To provide a reliable indication regardless of the flow patterns in the drum, the sensor is located in the pipe leaving the top of the drum.

EXAMPLE 24.2.

For the flash drum example, relate the sensors to the seven categories of control objectives. Present the results in a table similar to the presentations in Chapter 7.

The following table summarizes the relationship between the control objectives and the sensors for the flash process:

Control objective	**Process variable**	**Sensor**
1. Safety	Pressure in the closed vessel	P1
2. Environmental protection		
3. Equipment protection	Liquid level in drum	L1
4. Smooth plant operation	Pressure in the closed vessel	P1
and production rate	Liquid in the drum	L1
	Feed flow rate	F1
5. Product quality	Liquid composition	A1
	Flash temperature (inferential)	T6
6. Profit optimization	Flow of steam	F3
	Flow of process fluid	F2
7. Monitoring and diagnosis	Flow rate of vapor product	F4
	Flow rate of liquid product	F5
	Process fluid exchanger duty and *UA*	F1, F2, T1, T2, T3, T4

Additional sensors will be added in this chapter after new issues related to safety have been introduced.

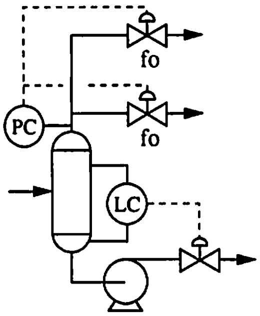


FIGURE 24.3

Example of use of final elements with small and large capacities to expand total range.

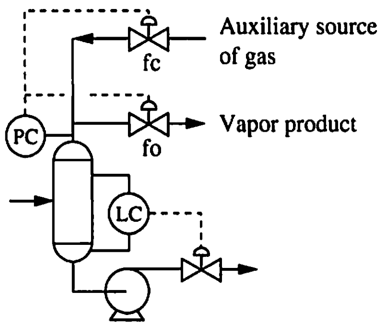


FIGURE 24.4

Example of final elements that allow in or out flow.

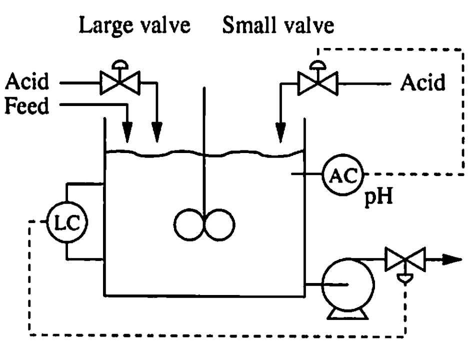


FIGURE 24.5

Example of the use of final elements with large and small capacities to improve accuracy in manipulations.

24.4 ◻ FINAL ELEMENTS

All final elements that are adjusted by an automatic controller or adjusted frequently by plant personnel must be automated. The automation of a final element requires a power source that changes the final element's value, usually the percentage valve opening, as determined by a signal transmitted from the control system. Many other final elements whose values change very infrequently are not automated and require a person to change their values manually at the equipment; thus, plants also contain many "hand valves." Some of the important features for an automated final element are discussed here.

Capacity and Precision

The final element should have the capacity to influence the manipulated process variable over the required range. As an initial guideline, a control valve should be 60 to 70% open at design conditions, so that the valve has considerable additional capacity to allow increased flow during disturbances or operation at increased production rates. However, each control system should be evaluated individually to ensure that the proper capacity exists.

Special designs are required when the range of the manipulated variable is large. For example, the feed to the flash drum in Figure 24.3 can vary from a small to a large amount of light, vaporized material. To accommodate the small, normal flow, a valve with a small capacity could be provided. However, a valve with a much larger capacity is provided to satisfy the infrequent, large vapor flow. The control design, using split range, is shown in the figure. Another example demonstrates the need to consider the sign of the manipulated variable as well as the magnitude. The drum in Figure 24.4 normally has a small vapor product; however, sometimes there is no vapor. To ensure that the pressure can be controlled for both cases, the pressure controller must be able to manipulate the outflow of product vapor or an inflow of a compatible gas. The control design, using split range, is shown in the figure.

A final element has a range over which it can accurately influence the manipulated process variable. For a typical control valve, the range of lowest to highest flows would be on the order of 1:20; thus, the range is quite large. In special cases, the control system might need to make quite small changes accurately when the total flow is relatively large. A two-valve arrangement that achieves this objective for strong acid–strong base pH control is shown in Figure 24.5. Normally, the larger valve is held constant, and the much smaller valve is adjusted by the controller. The larger valve is adjusted only when the smaller valve has reached a maximum or minimum limit. Finally, cascade principles can be employed to improve the valve performance by including a valve positioner, as explained in Chapter 14.

Dynamics

Again, slow dynamic elements in the feedback system degrade control performance. Therefore, the final element response should be much faster than that of other elements in the system. Most valve percent openings are achieved within a few seconds of a change in the signal to the valve, so that the valve dynamics are negligible for all but the fastest process control systems.

Failure Position

The failure position is selected to reduce the hazard to people and environment and damage to equipment when the signal to the final element is lost (i.e., when the signal to the valve attains its lowest value). Most valves are specified to go to either fully open or fully closed upon loss of signal. The proper failure position of a valve must be determined through an analysis of the integrated plant to determine the proper manner for relieving, storing, and venting material during an emergency. Naturally, the integrated plant must have the capacity to process (i.e., condense, combust, or store) material that cannot be vented to the environment.

Final Elements

- Capacity
- Precision
- Dynamics
- Failure position

EXAMPLE 24.3.

In this example, the final elements in the preliminary flash design in Figure 24.1 are considered. First, the valve in the liquid stream, v4, appears to be oversized, since its capacity is about seven times the design flow. Therefore, the valve specification should be changed so that the maximum flow through v4 is changed to 53% opened at design for a maximum flow of about twice its design value.

Second, the valve v2 is located in the condensate line, which means that the heat exchanger is behaving as shown in Figure 24.6*a*. In this design, the heat duty depends primarily on the area for condensation, which has a much higher heat transfer coefficient than the liquid-liquid film. As the valve is closed slightly, the liquid flow decreases, the area for condensation decreases, and the heat duty decreases. This is acceptable from a steady-state perspective; however, the dynamic response of the process depends on the direction of change. Increasing the duty is rapid because the liquid can flow quickly from the exchanger, but decreasing the duty is slow, because the liquid must condense and accumulate in the exchanger to reduce the area. A faster-responding design for both increasing and decreasing the duty is shown in Figure 24.6*b*, in which the steam flow is adjusted. Manipulating the valve in Figure 24.6*b* rapidly influences the steam pressure, and thus the temperature difference for heat transfer, to provide the amount of condensation needed. To complete the water material balance, the liquid condensate is collected in an inventory outside of the exchanger (in a steam trap), from which it is returned to the steam generators. The design in Figure 24.6*b* is preferred and will be used for the flash drum example.

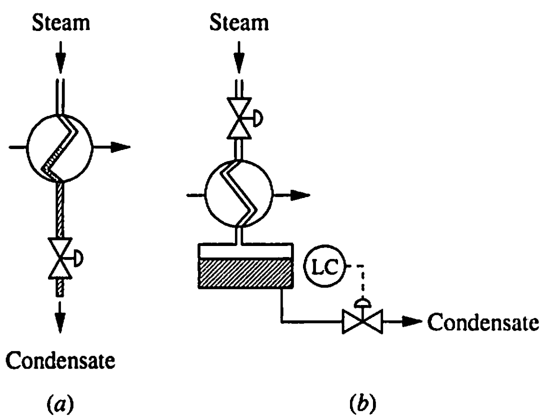


FIGURE 24.6

Alternative process designs for condensing heat transfer.

24.5 □ PROCESS OPERABILITY

One of the most important lessons in this book is that the process design and operating conditions have the most significant influence on control performance. Some processes are easily controlled; others require sophisticated algorithms to achieve satisfactory performance; and some processes cannot perform as required regardless of the type of control technology used. Thus, good control performance is one of the important goals of process design. Often, the ease with which a process is operated and controlled is referred to as *operability.* Some of the important factors that influence operability from the perspective of control performance are discussed in this section. The first topics address the possibility of control, and later topics address the quality of control performance.

Operability

- Degrees of freedom
- Controlled variables
- Controllability
- Operating window
- Feedback dynamics
- Disturbance dynamics
- Inventory

Degrees of Freedom

The process must have sufficient manipulated external (independent) variables to control the specified (dependent) variables; if sufficient manipulated external variables are not provided, the desired control performance will not be achievable. Since the transient behavior is of interest, the degrees of freedom are determined by analyzing the *dynamic* model of the process. As presented in Chapter 3, the degrees of freedom of a system are

$$\mathrm{DOF} = \mathrm{NV} - \mathrm{NE} \tag{24.2}$$

with DOF = number of degrees of freedom, NV = number of dependent variables, and NE = number of linearly independent equations. In modelling, we checked to ensure that the degrees of freedom were zero so that the model was consistent with the exactly defined problem statement. However, an essential part of the design task is to provide a process that can achieve the specified control objectives; therefore,

> The process without the controllers must have zero degrees of freedom when all external variables have been specified. To satisfy the control objectives, the number of manipulated external variables in the process must be equal to or greater than the number of dependent variables to be controlled.

The reason for the first requirement—zero degrees of freedom for the model—was presented in Chapter 3. The second requirement is a *minimum* requirement so that the process has the flexibility needed to satisfy the control objectives. If the number of manipulated variables, i.e., control valves, is smaller than the number of controlled variables, the system is overspecified and cannot achieve all objectives. In other words, an attempt is being made to control more variables than is physically possible for a specific process design. Corrections include reducing the number of variables controlled or adding flexibility to the process by increasing the number of manipulated variables by, for example, adding heat exchangers, bypass flows, and so forth. When the number of manipulated variables is greater than the number of controlled variables, the system is underspecified; it is possible that the control objectives can be achieved by many combinations of manipulated-variable values, subject to further analysis of controllability and dynamic performance.

Since the plant should have a unique operating policy, additional objectives, such as minimizing expensive fuel flows, can be added to the performance objectives.

When the controllers are added, the number of manipulated variables that are externally determined does not change, but the external variables change from the final element positions (for the open-loop system) to the set points (for the closed-loop system).

Selecting Controlled Variables

Process performance is defined in the control design form and generally depends on many variables. It would be the best situation if we could measure and control all of these variables; however, we often cannot. For example, the flash process has two product streams and up to six components in each stream; therefore, the product qualities and profitability of the process depend on many variables. We generally do not measure all components in all streams and usually do not have a sufficient number of manipulated variables to control all of these important variables *independently*. Therefore, we must recognize that we are often implementing *partial control*, in which only a subset of the process variables are measured and influenced by the manipulated variables.

> **Partial control** involves the measurement and control of a subset of the variables important for satisfactory product quality and high plant profitability.

An important control design decision is the selection of variables to be measured and controlled. This selection requires detailed knowledge of product quality specifications and likely plant disturbances, as well as a thorough understanding of process behavior. The selected variables should conform to the description of *dominant* variables given in the following summary.

> When **dominant variables** are maintained at their set points by automatic control, the process achieves acceptable product quality and profitability for the expected range of disturbances.

Many variables are influenced by the manipulations that are made to control the dominant variables. For example, changing a reactor temperature changes all reaction rates, and changing a flash temperature changes all equilibrium compositions. Naturally, the control of dominant variables cannot provide satisfactory process performance over an unlimited range of disturbance types and magnitudes. Thus, the engineer must evaluate candidates using fundamental and empirical models of the process and knowledge of reasonable disturbances.

Two important design decisions are required for successful partial control. The first decision is the choice of dominant variables that can be measured or very accurately inferred from measurements. Since onstream analyzers are costly and less reliable that sensors measuring temperature, pressure, flow, and level, some effort is directed toward finding process environment variables; however, onstream analyzers are often required and can perform well when designed and maintained properly. The second decision is the manipulated variables in the process. These should yield a feedback system that provides controllability, rangeability (large operating window), and good dynamic performance, as discussed in the next few subsections.

This discussion of partial control concentrates on closed-loop automatic control of processes. However, we must also recognize the importance of feedback compensation that is effected through analysis and actions performed by plant personnel at a much lower frequency than automatic control. Thus, the design should provide sufficient measurements, online or laboratory, and adjustable variables for this slow feedback correction. In this case, adjustable variables could be feed composition (through changes to feed-type purchase), catalyst properties (through gradual withdrawal and addition in fluidized beds), and equipment performance such as heat exchanger duties (through mechanical cleaning).

Additional discussions of partial control and dominant variables with many process examples are available (Arbel et al., 1996; Luyben et al., 1998).

Controllability

A process design with the necessary number of manipulated variables is able to satisfy the proper *number* of objectives, but this circumstance is not sufficient to ensure that satisfactory control can be implemented. An additional requirement is that the process must be able to achieve the objectives for the specified controlled variables by adjusting the specified manipulated variables. The requirement to test this feature of the process is *controllability,* which was introduced in Chapter 20 for a multivariable process. The definition of controllability used in this book is repeated here:

> A system is **controllable** if the controlled variables can be maintained at their set points, in the steady state, in spite of disturbances entering the system.

Recall that the system is deemed controllable when the steady-state gain matrix relating the manipulated to controlled variables is nonsingular, that is, when its determinant is nonzero. (If the number of manipulated variables is greater than the number of controlled variables, the gain matrix must have a *rank* equal to or greater than the number of controlled variables. This means that a subset of the manipulated variables can be selected for which the square gain matrix including all controlled variables is nonsingular.)

The controllability criterion was derived using the final value theorem, which requires some limitations to be placed on the process transfer functions $G_{ij}(s)$, basically that each be stable. The use of the final value theorem precludes most liquid levels, which are pure integrators and have transfer functions of the form $G(s) = k/s$. Since most process plants have liquid levels, the method for determining controllability should be extended to levels. To include integrating processes and maintain a simple analysis, we choose to consider the *rate of change of the* *level* as the controlled variable for the controllability analysis. Thus, the controlled variable is $sL(s)$, and the transfer functions between the rate of change of level and the manipulated and disturbance variables are constants and thus stable. Then the final value theorem can be applied, and the test for controllability is valid. In this case, the definition of controllability is modified to include the rate of change of level being returned to its desired value of zero.

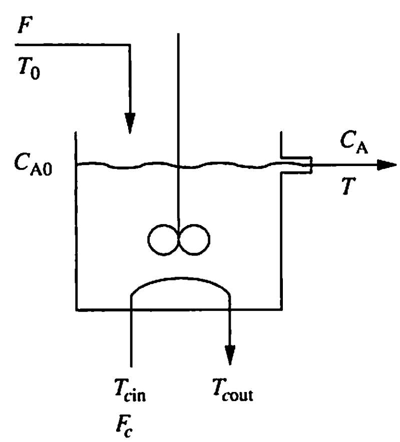
> The analysis of degrees of freedom and controllability evaluates whether the specified variables can be controlled by adjusting the specified manipulated variables in the region for which the linearized model is valid.

This analysis does not indicate the control structure required to achieve stable control or the range of disturbances that can be corrected; nor does it predict the variability of controlled variables from their set points.

Operating Window

The degrees-of-freedom and controllability requirements ensure that for at least some disturbances of very small magnitude, the control system can return the controlled variables to their set points. For practical control performance, the process equipment must have the capacity or range to satisfy the control design objectives for disturbances of expected magnitudes. When analyzing the steady-state performance of a process, the capacity is often represented by an operating window, as presented in Chapter 20. The coordinates are important process variables, and the region of acceptable performance is indicated as a "window" that is surrounded by an "infeasible" region, which represents operation that is either undesirable or not possible. The boundary or "frame" of the window is defined by the constraints in the control design form, and an important function of the control design is to maintain the process operation within the window. To achieve this goal, the manipulated variables must have sufficient capacity.

Equipment sizing is often determined by a steady-state analysis that chooses equipment designs (e.g., heat exchanger area, pump capacity, and distillation tower diameter) to maintain operation within the window for a defined set of expected operating conditions, including disturbances. However, a steady-state analysis is not always sufficient, because a process can exceed the steady-state limits of possible operation during transients, as demonstrated by the following example.



EXAMPLE 24.4.

Consider the nonisothermal, continuous stirred-tank chemical reactor described in detail in Section C.2. The nominal design operating conditions are the same as given in Appendix C, Case I except for the inlet concentration, which in this example has an initial value of 1.0 and experiences a step change to 2.0 kmole/m^3; thus, this exercise investigates a dynamic response returning to the initial conditions.

The dynamic response of the system for the step in inlet concentration is evaluated through numerical solution of the differential equations, and the results are given in Figure 24.7, which shows underdamped behavior. The same data is plotted in Figure 24.8 with concentration and temperature as the coordinates, and

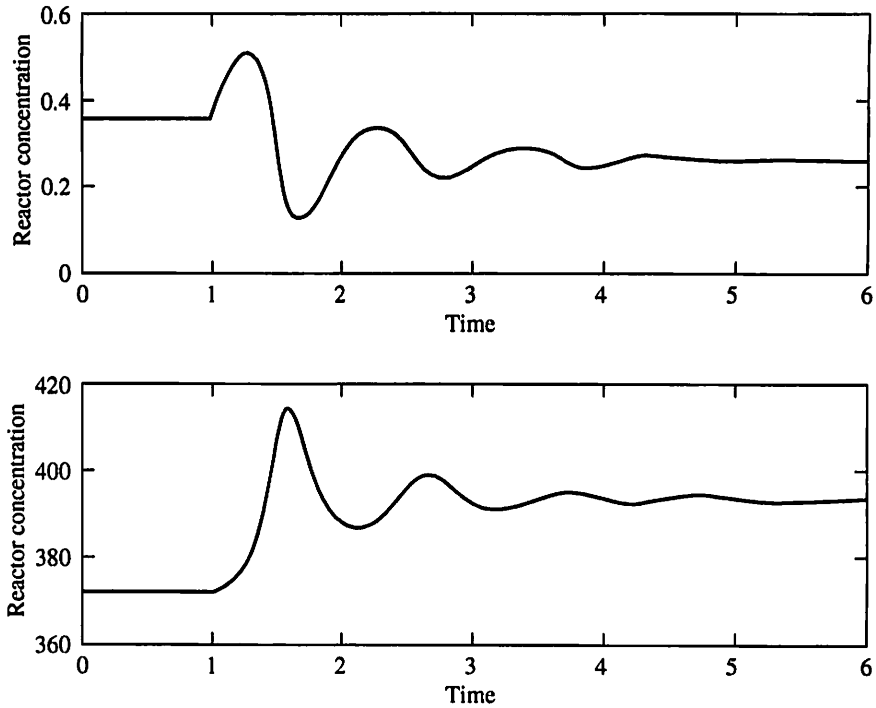


FIGURE 24.7

Dynamic response for Example 24.4.

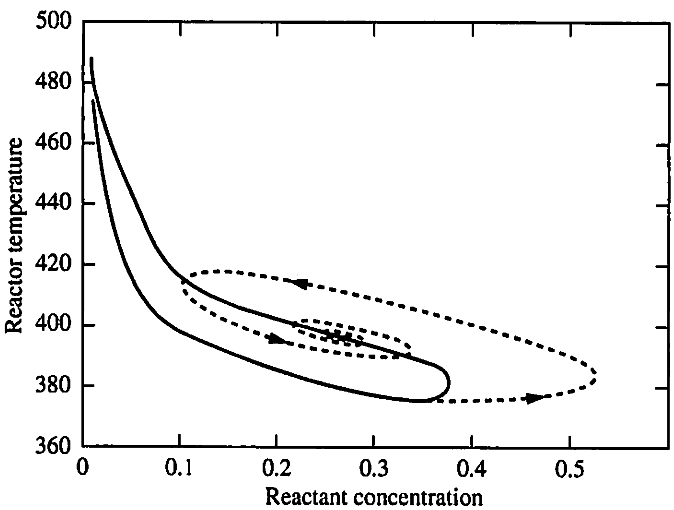


FIGURE 24.8

Steady-state operating window (solid) and dynamic trajectory (dashed) for Example 24.4.

the solid line defines the steady-state operating window: that is, the entire region of possible steady-state operation with $F_c = 0.5$ to 16.0 m^3/min and $C_{\mathrm{Ain}} = 1.0$ to 2.0 kmole/m^3. The trajectory in response to the step in C_{Ain} from 1.0 to 2.0 is shown as a dashed line with the arrows indicating the progression of time. Note that the transient begins and ends *within the steady-state window*, which it must, but that it violates the window by a considerable amount during the transient. This example demonstrates that the dynamic behavior of the process must be analyzed when determining the possible operating conditions that can occur in a process.

Given the importance of maintaining the process variables within an acceptable region and the fact that designing for a steady-state region does not eliminate the possibility of violations during transients, some equipment may have to have a greater capacity than required to meet steady-state demands in order to maintain all variables inside the window during transients (Rinhard, 1982). Failure to consider dynamics could lead to process designs that cannot perform properly during dynamic operation.

After the feasibility of control has been determined from the steady-state analysis, the effect of process dynamics on control performance is evaluated. The dynamic performance of control systems has been addressed throughout the book; here a few of the major conclusions are reiterated. However, this is not a comprehensive summary of important prior results, which would be very lengthy. The highlights are separated into discussions of feedback and disturbance dynamics.

Feedback Dynamics

The first three items in this section addressed the possibility of control; now, the performance issues are addressed. The process typically contributes the dominant dynamics in the feedback system; therefore, improving the process dynamics is especially important in improving control performance, as presented thoroughly in Chapters 13 and 21. Feedback process characteristics that contribute to good control performance include the following:

1. The process should be self-regulatory and open-loop stable, if possible.
2. The process dynamics should be relatively constant as operating conditions change.
3. The process should have fast dynamics with a small dead time and no inverse response.
4. The multivariable process should have favorable interactions.

The first characteristics are not required for good closed-loop control performance; however, stable, self-regulating processes are easier to operate in open loop (i.e., manually). Since all processes are operated manually on some occasions, they are included as good characteristics. The second characteristic of unchanging dynamics allows a controller with constant tuning to provide good control performance. If the dynamics change significantly, methods in Chapter 16 may be applied to compensate partially.

Fast feedback dynamics can be achieved by reducing transportation delays through shortening pipes, reducing (numerous) time constants through decreasing inventories, and speeding thermal processes through lessening the accumulation terms associated with heat exchangers, tank walls, and so forth. These steps improve feedback dynamics and usually also reduce equipment size and cost. However, there is a limit beyond which process equipment cannot be modified, and other approaches are required to improve dynamics. For example, additional improvements can be achieved by selecting the proper manipulated variable from several available; an example of this approach is discussed here with respect to the two temperature control systems in Figure 24.9*a* and *b*. The dynamics between the cooling (or heating) fluid flow and the temperature in Figure 24.9*a* is slow, because the temperature of the fluid and metal in the heat exchanger must be changed to affect the controlled variable. The design in Figure 24.9*b* allows the ratio between the flow through the exchanger and the flow bypassing the exchanger to be adjusted to control the temperature. Thus, the design using the bypass would be preferred when good control performance is required, although the equipment cost would be slightly higher. Note that the engineer must be creative in adding flexibility in the equipment for improved control.

Disturbance Dynamics

The basic objective of process control is to compensate for disturbances; therefore, the process should be designed to reduce the occurrence and effects of disturbances. Previous analysis has established that feedback control is improved when disturbances have (1) small magnitude, ΔD, (2) small gain magnitude, K_d, (3) favorable directions or interaction (small relative disturbance gain, |RDG|), and (4) frequencies much higher than the bandwidth of the disturbance process (where the open-loop amplitude ratio, $|G_d(j\omega)|$, is small) or much lower than the critical frequency of the closed-loop feedback system.

Many disturbances originate externally, such as from feed composition and cooling water temperature. However, the increased use of material and energy integration in process designs has increased the likelihood that variation in the process will negatively affect the dynamic performance of an associated process. As a simple example, consider the chemical reactor with a feed-effluent heat exchanger and exothermic chemical reaction in Figure 24.10*a*. With no temperature control, an upset in the feed temperature affects the reactor inlet, which affects the reactor outlet, which again affects the reactor inlet. Thus, an *energy recycle structure* is created, which heightens the sensitivity to disturbances and could lead to instability for highly exothermic systems. Naturally, the recycle structure could be eliminated by using two exchangers with utility fluids: one to heat the feed and a second to cool the effluent. However, that design modification would lose the energy efficiency advantages of the design in Figure 24.10*a*.

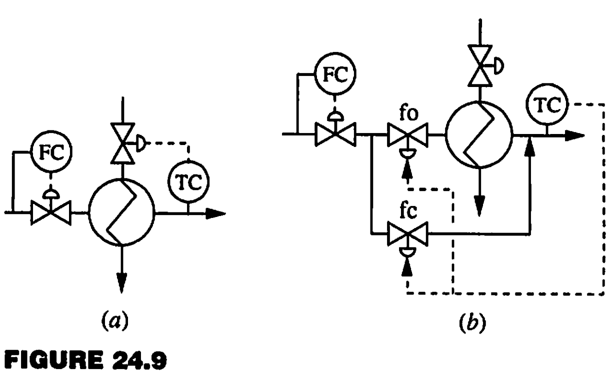
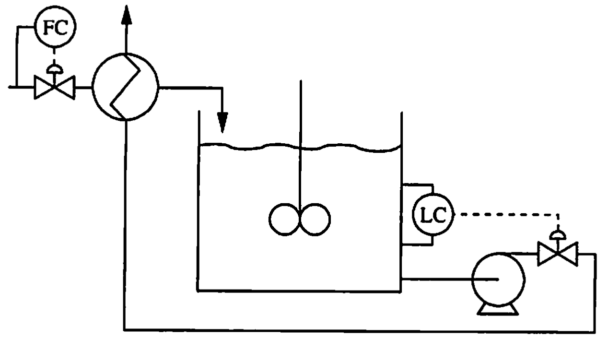
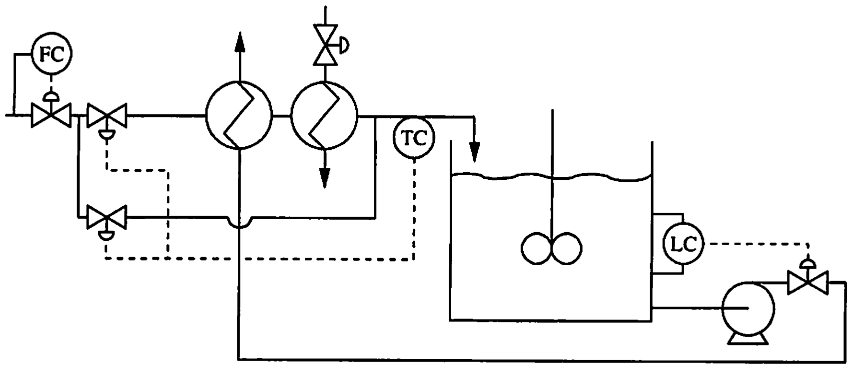


FIGURE 24.9

Alternative heat exchanger control designs.



(a)



(b)

FIGURE 24.10

Example of control design to reduce the effects of process integration.

An alternative way to improve the disturbance response and retain most of the energy savings is to control the inlet temperature so that it is nearly independent of the reactor outlet temperature. The approach requires an additional manipulated external variable, which can be supplied with a bypass placed around the feed-effluent heat exchanger. An additional heat exchanger—which would likely be needed for startup anyway—may be needed to provide the heat duty lost due to the bypass. As shown in Figure 24.10*b*, the reactor temperature could be controlled by adjusting the bypass around the feed-effluent exchanger, and the duty of the utility exchanger could be adjusted so that most of the feed preheat is supplied by the (inexpensive) heat integration.

Two general points demonstrated by this example can be applied to most material and energy recycle systems. First, feedback effects of disturbance propagation due to a recycle can be attenuated by adding an alternative path or source/sink where the recycle occurs. Second, the maximum steady-state benefit of process integration cannot always be achieved because of the poor dynamic behavior; however, most of the benefit can be realized by using the control methods demonstrated here while maintaining good control performance.

Inventory and Flow

Naturally, control of production rates and inventories is essential to good plant performance. The process should have sufficient inventories to ensure uninterrupted

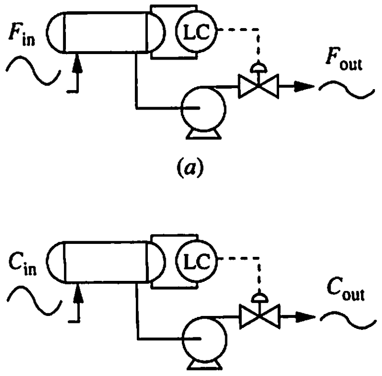


FIGURE 24.11

Use of inventory to improve control performance: (*a*) flow rate attenuation; (*b*) flow property attenuation.

flows to pumps and smooth flow rate variations throughout the plant as shown in Figure 24.11*a*. Good performance depends on the proper combination of inventory size and level control, including a nonlinear feedback algorithm where warranted. A straightforward manner for reducing the effects of disturbances in stream properties, such as temperature and composition, is to locate an inventory between the disturbance source and the controlled variable, but not in the feedback path, as shown in Figure 24.11*b*. However, inventories have disadvantages such as cost and hazards and large inventories are included sparingly—only when absolutely necessary to improve dynamic operation.

The following examples evaluate the possibility of control for the flash example by analyzing the degrees of freedom, controllability, and operating window.

EXAMPLE 24.5. Degrees of freedom

To perform the quantitative aspects of the design analysis in this chapter, a model of the flash process in Figure 24.1 is required. The goal of the model is to represent the dynamic input-output behavior of the system with accuracy adequate to make the design decisions correctly within the mathematical methods consistent with this book. Therefore, the model presented here is simplified to involve algebraic and ordinary differential equations (not partial differential equations) and approximate physical property data. The model is reported in Marlin (1995), and the analysis of the model for control system design is presented in this example.

The physical system in this example is shown schematically in Figure 24.12. The changes in sensors and final elements proposed in previous examples have been included.

Assumptions 1. All volumes are well mixed. 2. Densities, heat capacities, and heat transfer coefficients are constant. 3. Heat losses are negligible. These assumptions are common to all sections of the models.

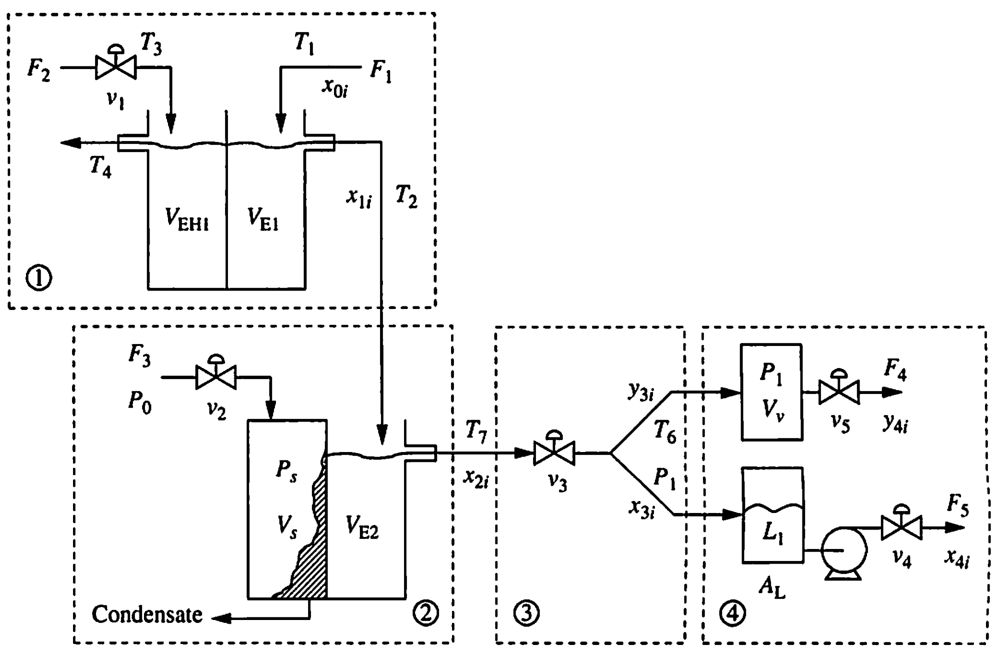


FIGURE 24.12

Approximate system used for modelling the flash process.

TABLE 24.3

Degrees of freedom for the flash process

Section	1	2	3	4	Total
Number of equations	11	13	24	18	65
Number of dependent variables	12	13	24	17	65
Number of external manipulated variables	1	1	1	2	5

Analysis. The analysis begins with a summary of the degrees-of-freedom analysis of the mathematical model, which is summarized in Table 24.3. The table presents the analysis of each section separately; however, the condition of zero degrees of freedom is required only for the complete process, not for any subsection. With all sections considered, the degrees of freedom for the entire system can be determined by summing the variables and equations to give DOF $= 65 - 65 = 0$; thus, the system is exactly specified. Also, the total number of manipulated external variables (valves) is 5; thus, no more than five dependent variables can be controlled.

EXAMPLE 24.6. Controllability

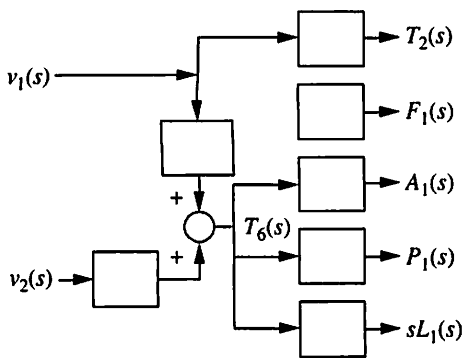
Next, the controllability of the flash system is evaluated. Since five manipulated variables exist, the possibility of controlling five variables is investigated. Controlled variables are selected so that the control system achieves the specified objectives. Typical variables are the process feed flow (F_1) and the liquid product quality (A_1 measures the mole% ethane in the liquid product). The pressure of the flash drum (P_1) should be controlled for safety and product quality, and the unstable liquid level (L_1) should be controlled for smooth operation and to prevent an overflow into the vapor line. Recall that the controllability of the rate of change of level, $sL_1(s)$, is determined, because the level process is an integrating process. Since this system is the same as the flash example in Chapter 17, which demonstrated that the flash temperature is a good indication of the liquid composition, the temperature (T_6) is provisionally selected as a fifth controlled variable.

The linear gains needed for the controllability check could be determined analytically for simple models. In this example they were determined numerically by introducing small changes in each manipulated variable and determining the steady-state value of the variables F_1, P_1, A_1, and T_6 and the rate of change of the level, L_1. The resulting equations are as follows:

$$\begin{bmatrix} F_1 \\ T_6 \\ A_1 \\ P \\ \dfrac{dL}{dt} \end{bmatrix} = \mathbf{K} \begin{bmatrix} v_1 \\ v_2 \\ v_3 \\ v_4 \\ v_5 \end{bmatrix} \quad \text{with } \mathbf{K} = \begin{bmatrix} 0 & 0 & 2.00 & 0 & 0 \\ 0.0708 & 0.85 & -0.44 & 0 & -0.19 \\ -0.00917 & -0.11 & 0.132 & 0 & 0.043 \\ 0.567 & 6.80 & 1.39 & 0 & -5.86 \\ -0.0113 & -0.136 & 0.31 & -0.179 & -0.0265 \end{bmatrix} \tag{24.3}$$

$$\det \mathbf{K} = -6.7 \times 10^{-7} \approx 0.00 \tag{24.4}$$

The result indicates that this 5×5 system is *not* controllable. The reason becomes



- Other manipulated variables and disturbances are not shown.
- v_1 and v_2 have no effect on F_1.

FIGURE 24.13

Block diagram of effects of v_1 and v_2.

apparent when the values of the coefficients in the linearized models for v_1 and v_2 are compared. The first and second columns in the matrix in equation (24.3) are different by only a multiplicative constant, which indicates that these two manipulated variables have the same effect on all of the controlled variables. The lack of independence can be seen clearly in the block diagram of the effects of v_1 and v_2 on the controlled variables in Figure 24.13. Note that both manipulated variables affect the flash temperature, and it is only through the effect on flash temperature that they influence the other controlled variables. Therefore, it is not possible to achieve independent steady-state values for any two controlled variables in equation (21.3) by adjusting v_1 and v_2. As a result, it is concluded that it is not possible to control the five variables by adjusting the five manipulated variables in equation (24.3).

However, it is possible to control a different selection of five controlled variables in this process. For example, it is possible to control variables F_1, P_1, A_1, and T_2 and the rate of change of the level, sL_1, with the five valves v_1 through v_5. This can be seen in Figure 24.13 by the fact that T_2 is affected by v_1 but not by v_2, thus introducing an independent relationship.

Since T_2 is not related to the control objectives, the decision is made to reduce the controlled variables to four and eliminate one manipulated variable. Since no control objective requires a specific behavior for T_6, it is eliminated; also, one of the two manipulated variables in Figure 24.13 must be eliminated: here, v_2 is retained and v_1 is eliminated. When this is done, the 4×4 system is controllable, as follows:

$$\begin{bmatrix} F_1 \\ A_1 \\ P_1 \\ \dfrac{dL}{dt} \end{bmatrix} = \mathbf{K} \begin{bmatrix} v_2 \\ v_3 \\ v_4 \\ v_5 \end{bmatrix} \qquad \text{with } \mathbf{K} = \begin{bmatrix} 0 & 2.0 & 0 & 0 \\ -0.11 & 0.132 & 0 & 0.043 \\ 6.8 & 1.39 & 0 & -5.86 \\ -0.136 & 0.31 & -0.179 & -0.265 \end{bmatrix} \tag{24.5}$$

$$\det \mathbf{K} = -0.126 \neq 0.00 \tag{24.6}$$

EXAMPLE 24.7. Operating window

In addition to ensuring that the system is controllable, which is exact only in a small (differential) region about the steady state, the operating window should be analyzed to ensure that sufficient flexibility exists for expected changes in external disturbances and set point changes. A sample operating window is given in Figure 24.14 for the flash process with the product composition (A1) and pressure (P1) controlled at their set points and the design values for the other external variables, such as feed composition. In this example, the limits to the window are from

1. The minimum external feed temperature, $T_2 = -10$
2. The minimum feed flow, $F_1 = 60$
3. The maximum heating (v_2 fully opened)
4. The maximum flow of product (v_4 fully opened)
5. The minimum heating (v_1 fully closed)

In all of these cases, the frame of the window was selected so that *all control valves* are at least 5% from their limits of 0 through 100%; thus, all controlled variables can be regulated, at least for small disturbances, within the window and on the frame. Additional cases demonstrate that the process can satisfy the requirements specified in objective 5*a* in the CDF. The large operating window involves the cost

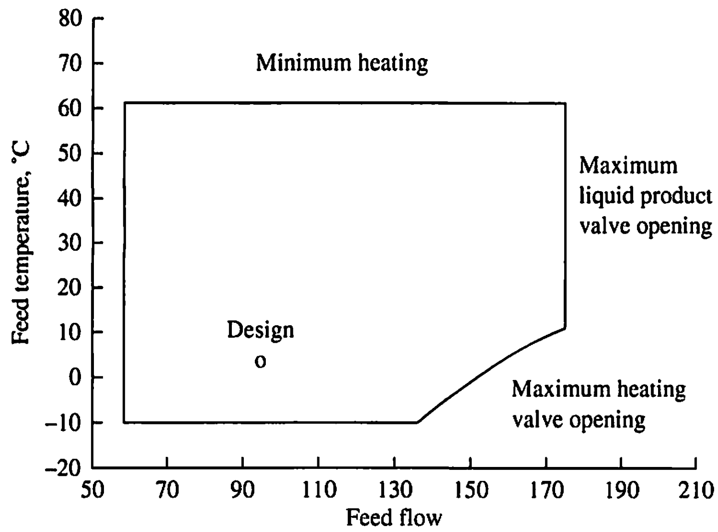


FIGURE 24.14

Operating window for flash process.

of purchasing larger equipment, and the capital costs must be balanced with the advantages of flexibility.

Further development of the control system, including the proper utilization of the T_6 sensor and a strategy for adjusting the additional manipulated variable (v_2), is given in the next section on control structure.

The operating window depends to some extent on the control design. In Example 24.7 the window is determined assuming that both heating valves are adjusted (by the control system) in response to the feed rate and temperature disturbances. After the control design has been completed, we must ensure that this assumption is satisfied; if not, we should reevaluate the operating window actually achieved with the control system. For example, if only one heating valve were manipulated, the size of the operating window would be smaller.

In conclusion, the process design and operating conditions have important effects on control performance that should be carefully analyzed by the control engineer. First, the possibility of control is determined by evaluating the degrees of freedom, controllability, and operating window; if the results indicate that the control objectives cannot be achieved, equipment sizing and process structures would have to be modified. Second, those processes that satisfy the preliminary criteria are evaluated for control performance, which depends on the feedback and disturbance dynamic behavior. Quite simply, feedback dynamics should be fast, and disturbance dynamics should have a small gain and long time constants.

24.6 ▫ CONTROL STRUCTURE

The control system should be designed to give the best performance possible for the process. The comments here refer to multiloop control technology.

Controlled–Manipulated Variable Pairing

The variable pairing should yield a loop with a significant process gain. If the process gain is too small, the controller will not be able to return the controlled variable to its set point when disturbances occur. If the process gain is too large, the controller will be required to adjust the manipulated variable with great accuracy; since such accuracy is usually not possible (for example, because of valve sticking and hysteresis), oscillations will occur. The process gain can be expressed in dimensionless form (scaled), $(K_p)_s$, by relating the variables to their ranges.

$$(K_p)_s = K_p \frac{\text{range of MV}}{\text{range of CV}} \tag{24.7}$$

The typical range of values for this dimensionless process gain is 0.25 to 4.0. Values outside this range are possible but should be evaluated carefully so that satisfactory manipulated-variable capacity and sensor reproducibility are provided. Note that this evaluation requires an estimate of the expected disturbances. The control systems in Figures 24.3 to 24.5 demonstrate approaches to loop pairing with extreme demands on valve range.

The loop pairing should be selected with regard to the effects of interaction in multiloop systems. Analysis methods for multiloop systems were presented in Chapters 20 and 21 (which the reader might review at this point) and are briefly summarized as follows:

1. Automatic control should be provided for all non-self-regulatory or open-loop unstable variables, because if they are not controlled, they will drift out of the acceptable operating region. Manual regulation of such variables is difficult and time-consuming for plant personnel; reliable process operation requires automation.
2. Normally, variables are not paired when their relative gains are negative or zero. This will make the tuning process easier and will result in better performance when some control loops are not functioning (i.e., are in manual or have manipulated variables at their upper or lower limits).
3. The dynamics of the feedback loop pairings should be fast, with small dead times and little inverse response. The most important controlled variables should be paired to give fast feedback loops, even though this might somewhat degrade the performance of some variables of less importance.
4. The pairings should be selected to reduce unfavorable interaction and increase favorable interaction. The relative disturbance gain (RDG) can be used as an indication of how a pairing might affect the control system performance.

Finally, when the system has an unequal number of controlled and manipulated variables, the control structure should be able to alter the pairings to ensure that the objectives are attained. Methods for decentralized multiloop control are split range, signal select, and valve position controllers, which were presented in Chapter 22; a method for centralized multivariable control is Dynamic Matrix Control, presented in Chapter 23.

Disturbances

The effects of disturbances should be reduced through good control design. Two very effective designs that reduce the effects of disturbances are cascade and feed-forward control, covered in Chapters 14 and 15.

EXAMPLE 24.8.

In this example, the control structure in the flash process is considered. First, the inventories should be controlled, and the natural pairings are the drum pressure with the vapor exit valve (v_5) and the drum liquid with the liquid exit valve (v_4). Also, the feed flow rate should be controlled, and valve v_3 should give fast control.

Second, the measurements to be controlled are selected with consideration for the goal of partial control. The performance of this process depends on concentrations of all six components in the two product streams, but all of these compositions cannot be controlled with the process equipment provided. The statement in the control design form indicates that not all variables are required to be constant; only the concentration of ethane in the liquid must be controlled. Therefore, ethane liquid concentration is a *dominant variable* for this process, and we select this as a measured variable to be controlled. Since the onstream composition sensor requires two minutes to analyze each sample, feedback control will be rather slow. We recognize the close relationship between the composition and the process environment variables pressure and temperature, and evaluate each for possible inferential/cascade control. Adjusting the flash drum pressure to achieve acceptable composition would generally require excessive pressure variation (and expensive vessels, pipes, and pumps); therefore, we select temperature. The good inferential relationship between temperature and composition in this process has been thoroughly analyzed in Section 17.2.

This cascade observes the design rules introduced in Chapter 14: the secondary variable is measured, indicates important disturbances, depends in a causal manner on the manipulated variable (v_1), and has faster dynamics than the primary, because of the slow primary measurement. Recall that adding this controller does not change the degrees of freedom, because one external variable (the T_6 set point) becomes a dependent variable, one equation (the controller) is added, and one external variable (the analyzer set point) is added. Also, since the process equipment is unchanged, the operating window is not affected.

Third, the control objectives state that the process fluid flow to the first heat exchanger should be maximized before the steam is used to heat the feed. This is a system with one controlled variable and two manipulated variables with a fixed priority of adjustment. Therefore, a split range control design can be used. The resulting design for the product quality control is shown in Figure 24.15. Again, the split range controller does not violate degrees-of-freedom requirements, because, as discussed in Chapter 22, only one valve is adjusted at a time. The controllability of the system is ensured when either v_1 or v_2 is manipulated, as indicated in Figure 24.13 and as can be verified by evaluating the appropriate gain matrix.

The loop pairing can also be analyzed using methods introduced in Chapters 20 and 21. For example, the relative gain array (RGA) can be applied to ensure that the design does not violate guidelines such as not pairing on negative RGA elements. Following the suggestion of McAvoy (1983), the relative gain is calculated using the self-regulating variables and the rate of change of the integrating level. Thus, the steady-state gains for this 4×4 control system are those in equation

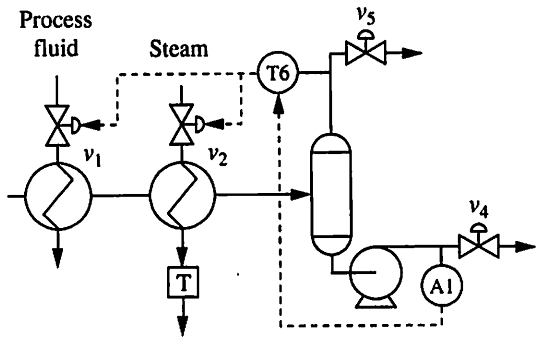


FIGURE 24.15

Control design to speed feedback disturbance response and optimize the use of heating sources.

(24.5). The relative gain array is

$$\text{RGA} = \begin{array}{r|rrrr} & v_2 & v_3 & v_4 & v_5 \\ \hline F_1 & 0 & 1 & 0 & 0 \\ A_1 & 1.83 & 0 & 0 & -0.83 \\ P_1 & -0.83 & 0 & 0 & 1.83 \\ dL_1/dt & 0 & 0 & 1 & 0 \end{array} \tag{24.8}$$

Based on selecting pairings with positive relative gains, the analysis recommends the pairings F1 $\rightarrow v_3$, A1 $\rightarrow v_2$ (which we have selected via T_6 as a cascade), P1 $\rightarrow v_5$, and L1 $\rightarrow v_4$. The analysis confirms the "common sense" selections based on semi-quantitative reasoning.

In conclusion, design of the proper control structure requires considerable knowledge of process dynamics, dominant disturbances, and equipment capacities. The control structure is tailored to satisfy the performance objectives for the process using the appropriate methods in Parts III through V.

24.7 ▫ CONTROL ALGORITHMS

After the control structure has been selected, the algorithms and tuning can be selected to give the best performance for that structure.

Feedback and Feedforward

Feedback control should be used extensively, because it corrects for all disturbances, even unmeasured disturbances, that influence the measured controlled variable. All of the feedback single-loop enhancements, such as cascade and gain scheduling, should be considered to improve the control performance of a feedback system. Feedforward control should be considered as an enhancement to feedback control when the feedback process is difficult to control because of long dead time and unfavorable interaction.

Algorithm

The control algorithm should be matched to the application. In particular, most feedback systems desire zero steady-state offset; therefore, this requirement should be satisfied by including the integral mode in the PID controller or by appropriate considerations in a model predictive controller. Based on its generally good performance and widespread acceptance, the PID controller should be used for most multiloop feedback control systems. Only when another algorithm provides demonstrably better performance should it be chosen over the PID. There are some cases, such as loops with inverse response or very long dead times [and large $\theta/(\theta + \tau)$], where a predictive controller might give better performance.

The feedback controller should be selected to be relatively insensitive to modelling errors, and the associated tuning errors, for the expected range of errors. Most single-loop feedback control algorithms satisfy this requirement. However, sensitivity analysis showed that some multivariable control designs (e.g., decoupling and centralized DMC control) are sensitive to certain model errors when the process has strong interactions (i.e., large elements in the relative gain array).

Tuning

Tuning parameters for all algorithms should be based on a careful analysis of the desired performance of all process variables. Typically, empirical methods are used for determining models for tuning. However, fundamental models are very useful for (1) verifying empirical results, (2) determining how model parameters depend on process operation (e.g., throughput), and (3) providing models for complex, nonlinear processes.

It is important to remember that the manipulated variable in a control system (e.g., steam flow) is another plant variable. The engineers involved with plant design and operations are responsible for ensuring the availability of appropriate utility systems that can be varied to control the process. However, extreme variation in the manipulated variable can cause disturbances in other units in the plant. The typical relationship in feedback systems was covered in Chapter 13, where it was shown that in the region of good tuning, the variability of the manipulated variable increases as the variability of the controlled variable decreases. In most cases, tuning can be selected to reduce the variability in the manipulated variable significantly, with only a small increase in the controlled-variable variability. For this reason, as well as for robustness, the controller is normally tuned to eliminate extreme variability in the manipulated variable.

Often, the tuning parameters do not have to be modified in response to moderate changes in process operation, because the dynamic responses do not change significantly over the range of operation. Recall that 10 to 20% errors in parameters are common. However, if the changes in process operating conditions are large or the process is highly nonlinear, the controller tuning should be adjusted in real time to maintain stability and acceptable performance. Approaches for adapting the tuning, having the goal of maintaining the same stability margin (and relative control performance), were explained in Chapter 16.

Finally, the tuning of multiloop controllers must be performed with consideration of the interaction among loops. This issue, along with tuning guidelines, was discussed in Chapters 20 and 21, where it was shown that the relative gain gives some indication of the extent that tuning must be adjusted to account for interaction. Also, the relative importance of the controlled variables is considered when tuning the controllers, with the tuning selected to reduce the deviation of the most important variables from their set points.

EXAMPLE 24.9.

In this example, a few issues related to tuning the controllers for the flash process are discussed. First, the order of the tuning is important. The level controller tuning can be determined without experimental modelling using the vessel size, and because the level is non-self-regulating, it should be tuned first. No specification is placed on the variability of the liquid leaving the drum, and a proportional-only controller with tight level tuning is selected because of the importance of not having liquid carry over. Also, the pressure in the drum could easily exceed its limits and should be tuned next using the standard methods. Then, the split range controller, T_6, will adjust v_1 and v_2. The dynamics between the valves and the temperature sensor can be expected to be different, and the gain matrix in equation (24.3) shows that the steady-state gains are different by a factor of about 12. Thus, the tuning of the T_6 PI controller should be adapted based on the condition of the split range. Finally, the analyzer measurement is updated only every two minutes; this long execution period for the feedback controller will require some detuning using the guidelines from Chapter 11 ($\theta' = \theta + \Delta t/2$).

24.8 ▫ CONTROL FOR SAFETY

Before completing the discussion on design decisions, safety must be discussed. Safety is addressed in the first control objective, and some control decisions, such as controlling the pressure in the flash process, have been made to satisfy safety requirements. However, special control system features are required, because of the importance of this objective. These features are often implemented in multiple layers, with every layer contributing to the safety of the system by taking actions only as aggressive as required for the particular situation (AIChE, 1993; Crowl and Louvar, 1990).

Control for Safety

- Basic process control system
- Alarms
- Safety interlock system
- Safety relief valves
- Containment

Basic Process Control System (BPCS)

The first layer involves the basic process control approaches discussed in prior sections, which employ standard sensors, final elements, and feedback control algorithms. This first layer maintains the process variables in a safe operating region through smooth adjustment of manipulated variables; this action does not interfere with, but rather usually enhances, the profitable production of high-quality material. However, the basic control system relies on sensors, signal transmission, computing, and final elements, which occasionally fail to function properly. In addition, the process equipment, such as pumps, can fail. Even if all elements are functioning properly, the control system may not maintain the system in the safe region in response to all disturbances; for example, a very large disturbance could cause a deviation of key variables into an unacceptable region.

The basic process control layer can employ standard techniques to improve its response to a fault. For example, the use of several sensors with a signal select reduces the effect of a sensor failure. (An example is the temperature control system in Figure 22.10, which can reduce the likelihood of a temperature excursion due to the failure of a single temperature sensor.) Also, the use of split range control allows a controller to manipulate an additional (e.g., larger-capacity) valve in response to an unusual circumstance. An example of this approach is given in Figure 24.3. However, these techniques do not reduce the likelihood of injury or damage to an acceptably low probability; therefore, additional layers are implemented to improve safety.

Alarms

The second layer involves alarms, which are automatically initiated when variables exceed their specified limits. These alarms involve no automatic action in the process; their sole purpose is to draw the attention of the process operator to a specific variable and process unit. The person must review the available data and implement any actions required. A great advantage of involving operators is their ability to gather data not available to the computer. For example, an operator can determine the values of instruments that display values locally and can check the reliability of some sensors as part of the diagnosis. The operator usually takes action through the process control system; these actions could include placing a controller in manual status and adjusting the manipulated variable to a new value. Since the final element may not be functioning, the operator has the option of going directly to the process and adjusting the valve manually (or having this task performed by another person).

It is good practice for the alarm to be based on an independent sensor, because using the same sensor for alarm and control prevents the alarm from identifying the failure of the sensor to indicate the true value of the process variable. An alarm is shown on a process drawing with a three-letter identification; the second letter is "A" to designate alarm and the third letter is either L (low) or H (high). For example, PAH indicates an alarm when the pressure measurement exceeds its high limiting value. The alarm is usually annunciated by activating a visual indicator (e.g., a blinking light) and an audio signal, beeping horn. These signals continue until the operator acknowledges the alarm; thereafter, the visual indicator remains active (e.g., a nonblinking light) until the variable returns within its acceptable limits. The blinking light indicates the variable involved, its current value, its alarm priority, and whether the variable has exceeded its high or low limit. Alarms can be arranged into three levels, depending on the severity of the potential consequences of the process fault or upset:

LEVEL 1 (HIGH). These alarms are designed to indicate conditions requiring prompt operator action to prevent hazards or equipment damage. Special color and visual displays and a distinct audio tone should be used to alert the operator. Examples of level 1 alarms are high pressure in a reactor; low water level in a boiler; and activation of a safety interlock system that has stopped operation of some processes (see next topic).

LEVEL 2 (MEDIUM). These alarms are designed to indicate conditions requiring close monitoring and operator action to prevent loss of production or other costly (but nonhazardous) situations. The operator typically has some time to analyze the alarm, along with other measurements, and make corrections that can maintain the process in an acceptable region of operation. These alarms should be annunciated in the same general manner as the level 1 alarms, although with distinct colors and tones.

LEVEL 3 (LOW). These alarms identify conditions that are not critical to the operation of the process and require no immediate action by the operator. These can be entered directly into a database for occasional review by the operators and engineers. These alarms should not be annunciated.

Some care must be taken in designing alarms. The major issue is the overuse of alarms. Kragt and Bonten (1983) report that an operator in an industrial processing plant experienced an average of 17 alarms per hour and that the operator took an action after only 8% of these alarms! Most of the alarms were not necessary and needlessly distracted the person. Such poorly designed alarm systems lead to lack of attention by the plant personnel to the occasional, but critical, *important* alarm.

Safety Interlock System (SIS)

The third layer involves automatic feedback control for situations when process variables approach "hard" constraints that should not be exceeded; these could cause injury to people or the environment or damage to expensive equipment. Because of the importance of preventing such situations, the actions taken are extreme and disrupt the process operation; usually, they stop all or part of the process operation by immediately closing (or opening) key valves to move the process to a safe condition. These control systems are termed *safety interlock systems* (SIS) or *emergency shutdown systems* (ESS).

As with alarms, this control layer should use a sensor independent of the basic control system; in addition, this automated system should use a final element independent of the basic control system. The equipment selected for this purpose must be of the highest reliability possible. Depending on the severity of the consequences, this layer may use several sensors and final elements. In some applications, three sensors are used, and the feedback control system bases its decision on the majority of the three; this approach prevents an occasional (individual) sensor failure from stopping process operation, while identifying an actual dangerous condition with high reliability. The control action taken is straightforward and simple to implement. Typically, a solenoid valve, which is normally closed to hold the air pressure to the pneumatic valve at a high value, opens and vents this pressure to atmospheric upon receiving a failure signal, allowing the pneumatic valve to attain its failure position. If this action is taken on a valve that is also used for basic process control, the solenoid valve is placed between the controller output and the valve; thus, under normal circumstances, the controller adjusts the control valve without alteration, whereas a failure signal disconnects the controller output from the valve, which goes to its failure position.

The valve selected for use in an SIS should have a capacity large enough to handle the largest expected flow. For example, a valve to vent a distillation tower may be based on a situation in which the condenser fails. Also, the manipulated variable should have a very fast effect on the key process variable and be able to maintain the process in the safe region; thus, dead times and time constants should be small. The limiting value for the initiation of the SIS is selected to be in the safe region and far enough from the undesired value that the largest expected disturbance will not cause an unsafe condition.

The manner in which a safety interlock system is shown on a process drawing depends on the complexity of the logic. If only one measurement is compared with its low or high limit, a two-letter designation is used, with the second letter being "S" for switch; for example, LS is a switch that changes state based on a level measurement. If the logic is complex, perhaps using many sensors, all measured signals are connected to an "SIS" symbol, and the SIS is connected via signal lines to all manipulated valves (or motors, etc.). Separate documentation is required for the more complex SIS systems.

Safety Valves

The fourth layer involves feedback systems that are self-actuating, that is, which do not require electrical, pneumatic, or hydraulic power sources and have no significant distance of signal transmission. These features contribute to very high reliability. The major application at this layer is the safety valve, which is a valve normally held closed by a spring. When the pressure reaches the preset limit, the force due to the process pressure is high enough to overcome the force of the spring, and the valve begins to open. When the process pressure decreases, the safety valve is designed to close. The engineer must be sure that the material flowing through the safety valve can be either (1) released to the environment safely (e.g., steam), (2) processed to eliminate hazards (e.g., combusting hydrocarbons), or (3) retained in a containment vessel for later processing (e.g., wastewater storage and nuclear plant containment building).

These layers should be carefully designed, properly installed, and meticulously maintained. Through good or poor practices, the high level of safety may be enhanced or compromised. A few of these good practices are given in Table 24.4.

TABLE 24.4

Good Practices in Control for Safety

1. Never bypass the calculation (logic) for the SIS; that is, never turn it off.
2. Never mechanically block a control valve so that it cannot close.
3. Never open manual bypass valves around control and shutdown valves.
4. Never "fix" the alarm acknowledgment button so that new alarms will not require the action of an operator.
5. Avoid using the same sensor for control, alarm, and SIS.
6. Avoid combining high- and low-value alarms into one indication.
7. Evaluate the selection of alarms critically. Do not have too many alarms.
8. Use independent equipment for each layer, including computing equipment.
9. Select emergency manipulated variables with a fast effect on the key process variable.
10. Use redundant equipment for critical functions.
11. Provide capability for maintenance testing, because the systems are normally in standby.

Containment

The final layer involves containment, such as dikes, for major incidents. This layer may not prevent major hazards, but it can prevent their propagation to other sections of a plant and to the surrounding community. Other design issues, such as reliable electrical power supply, are also important for safety control; these are covered in the references.

EXAMPLE 24.10.

In this example, safety controls for the flash process are considered, and the results are shown in Figure 24.16.

There are several issues at the basic process control system layer. First, the pressure in the closed vessel should be controlled, and the valve in the overhead vapor line is a natural choice for the manipulated variable, because it has a very rapid effect on pressure. Second, the liquid level should be maintained within reasonable limits, and the valve in the bottom exit is a natural choice for manipulation. To prevent the liquid flow through the pump from falling below the minimum, the level controller could reset the flow controller set point, with the set point bounded to always be above the limit. Third, the use of a temperature cascade improves the reliability of the product quality control, because the analyzer would be much more likely to fail than the temperature sensor. Finally, the failure positions of the valves are selected to reduce the likelihood of high pressure, high temperature, and an overflow of liquid in the vapor line.

The alarm layer could conceivably include high and low alarms on every variable, but this would lead to excessive interruptions for the operator. Here, alarms will be placed on high pressure and high and low level. The analyzer measurement would normally not be alarmed unless composition variation led to unsafe conditions.

An SIS system would not normally be employed in this process. However, as an example, we will assume that the objective of preventing a liquid overflow in the vapor line from the drum is critical [see Kletz (1980) for an industrial example]. A different type of sensor is used for the SIS; this sensor provides redundancy

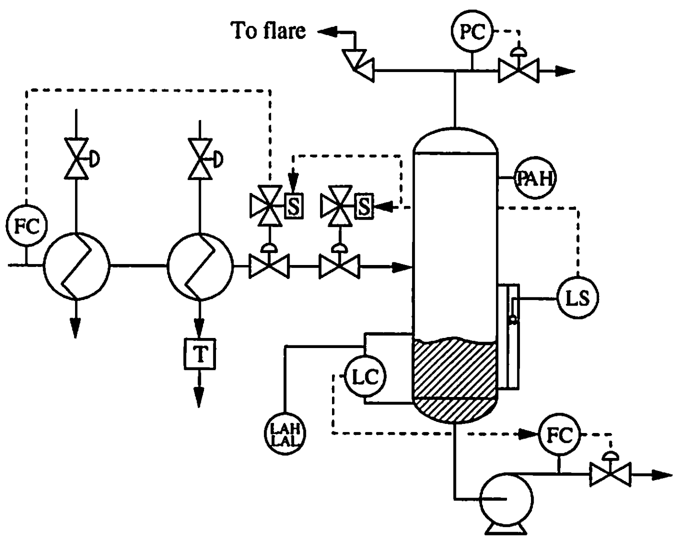


FIGURE 24.16

Safety-related controls for the example flash process.

and diversity at the SIS level. This level sensor is used to determine when the level approaches the limiting value and to activate the emergency action to reduce the feed flow to zero. Both the control valve and an independent valve are used to enforce this SIS. When the safety interlock system activates, the process will experience a major disturbance, and the product will not observe the quality specifications.

The drum can be closed by the (improper) operation of the control valves; thus, a safety valve should be included, as shown in the figure. The combustible material must be contained or processed; typically, it would be diverted to a plant fuel system or combusted in a flare.

The multiple-layer approach described in this section provides excellent protection for most chemical processes. The reader must be aware of the importance of excellent detailed design and construction of equipment for safety control. This section simply presents some introductory concepts and is not meant to teach the practice of safety in design. The novice should refer to the many industrial standards and engage experienced consultants when designing safety control systems.

24.9 ▫ PERFORMANCE MONITORING

Monitoring should be considered at the design stage to ensure that the important performance measures are identified and that the sensors with required accuracy are provided. The most important purpose of short-term monitoring is to enable the plant operator to diagnose incipient problems, preferably before the problems worsen and cause major upsets. The purpose of longer-term monitoring is not only to record the performance but also to diagnose the reasons for good and poor performance. The results of this diagnosis can be used by the engineer as a basis for improving product quality, equipment performance, and profit through changes in operating conditions, control designs, and process equipment.

Real-Time Monitoring for Process Operation

The plant operators are part of the overall "control system"—that is, they are responsible for many feedback control tasks that are not automated, such as switching from one feed tank to another. Also, they are responsible for supervising the process equipment and automatic control system. To perform these tasks, operators require a thorough understanding of the process, along with rapid access to many measured values. The system designer must recognize that because the diagnosis of the control system, including sensors, is an important task, the operators need parallel information on key variables provided by independent sensors. The alarm feature of control systems, discussed in the previous section, can help the operator monitor hundreds of variables by drawing attention to variables that are outside of their normal operating ranges.

Variability of Key Process Variables

Individual measured variables can be analyzed as part of a longer-term monitoring program. The average values of most important variables provide a quick indication of the process performance, and when the average is not close to the desired value, improvement is clearly in order. However, good performance is not ensured when the average conforms to the desired value, as demonstrated in Chapter 2. The variability is important in determining the plant performance, because average process performance depends on the length of time each variable spends at values in the distribution. This concept is shown in Figure 24.17. The average performance can be calculated from the empirically measured distribution without making assumptions concerning the normality of the distribution, and a broad distribution indicates considerable operating time far from the best conditions, even if the average conditions seem acceptable.

The total number of incidents also gives valuable insight into performance. One type of incident is the activation of alarms, with each important alarm monitored separately. Care should be taken in monitoring alarms, because one process disturbance can cause numerous alarms before the plant operation is returned to normal conditions. Other incidents include the number of times important constraints are violated, such as products outside specified quality limits, and the activation of safety interlock systems. Each major incident provides valuable information on the performance of the process and controls, which can be used in designing improvements.

Calculated Process Performance

In many cases, the performance of important process units can be estimated from measured variables. Some of these variables indicate the overall performance of the plant (e.g., energy consumption per kilogram of product sold). These are useful in indicating the overall performance but not usually complete enough to direct

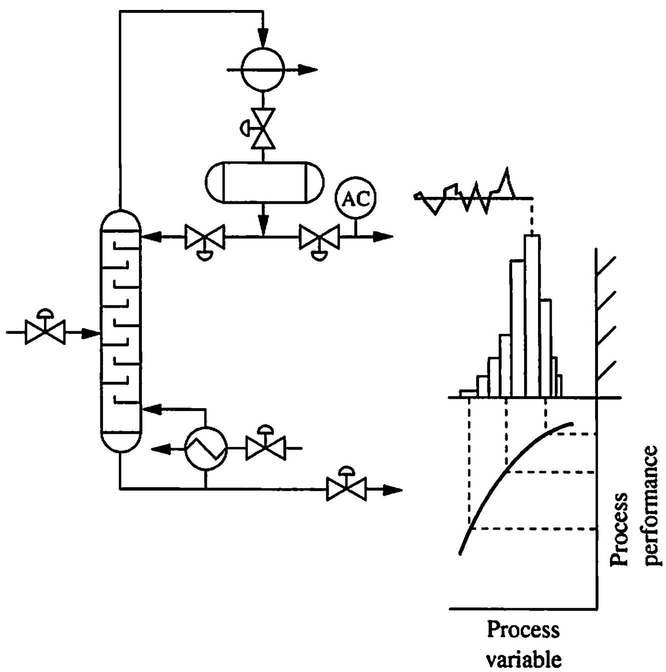


FIGURE 24.17

Schematic of the procedure relating a key process variable to performance.

diagnosing and improving performance. However, monitoring the performance of individual units provides very useful information. For example, the efficiency of a boiler gives insight into the performance of the excess-air control system, as well as other factors such as the heat transfer coefficients. Commonly monitored calculated variables are compressor and turbine efficiencies, heat transfer coefficients, fired-heater efficiencies, and the selectivity of chemical reactions to desired versus undesired products.

Utilization and Performance of Control

The fraction of time that each control system is in operation in automatic should be monitored. Although this information cannot be used to diagnose control performance, a low service factor (time used/time should be used) is a clear indication of unsatisfactory performance, at least in the opinion of the process operators. More information can be determined from dynamic plant data on the performance of the control system. Methods are available for estimating (1) the best possible feedback control, (2) the improvement possible with feedforward control, and (3) likely deficiencies in the existing control system (e.g., feedback controller tuning or feedforward disturbance model). These methods rely on mathematical analysis that is beyond the level of this book, but they require only simple interpretation of graphical results by plant personnel after they are implemented (see the Additional Resources in Chapter 9).

EXAMPLE 24.11.

In this example, monitoring for the flash process in Figure 24.1 is discussed. First, the averages and standard deviations of important process variables should be calculated from real-time data. Typically, the most important variables would be the flow rates, the flash temperature, and the liquid composition. The loss of heavy material in the vapor could be monitored through infrequent samples analyzed in a laboratory. Second, an appropriate sensor should be selected if an accurate measurement of the vapor flow rate is needed, perhaps for a record of sales. If an orifice meter is used in a stream with changing pressure, composition, or both, the density and the pressure drop are required to measure the mass flow rate accurately. Finally, the process performance would depend on the heat transfer coefficient in the process fluid heat exchanger. This could be monitored using measured temperatures and flows. A low value of the heat transfer coefficient, based on process data that satisfied material and energy balances (and thus is considered accurate), along with high steam use would indicate poor performance. Performance could be improved by taking the heat exchanger out of service for a short time to clean the surface.

24.10 ▫ THE FLASH EXAMPLE REVISITED

The original design in Figure 24.1 has been discussed in the examples in this chapter, where the improvements summarized in Table 24.5 were proposed. The final design, which incorporates all improvements except the SIS on high level, is shown in Figure 24.18. This design satisfies the objectives in the control design form and is typical for industrial systems.

TABLE 24.5

Summary of design decisions for the original flash process

Design decision	
Measurements	F_1 moved to one-phase flow region
	T_5 redundant measurement removed
	T_6 moved to vapor space for faster response
Final elements	v_2 changed to steam flow
	v_4 reduced maximum flow
Process	Analyzed degrees of freedom; five manipulated variables exist
	Analyzed controllability to determine that only four (meaningful) variables can be controlled
	Operating window large enough to satisfy objectives
Control structure	Cascade from A_1 to T_6 inferential variable
	Split range to adjust both heating valves
Algorithms	Standard PI control except for P-only level controller
	Adaptive tuning for T_6 when changing the split range
Safety	Basic regulatory control of inventories with minimum liquid flow
	Valve failure modes
	Alarms on pressure and level
	Safety interlock system for high level
Monitoring	Correct vapor flow for density
	Monitor heat exchanger UA
	Monitor product quality (A_1) variance

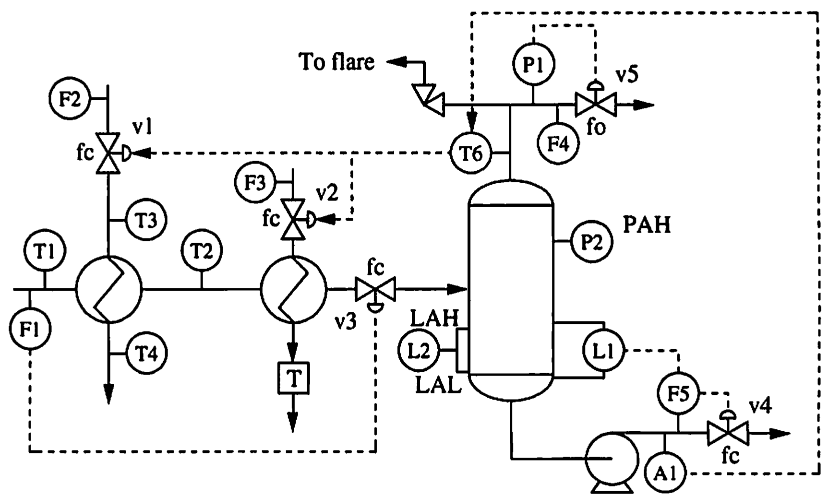


FIGURE 24.18

Modified design for the flash process incorporating improvements in Table 24.5.

EXAMPLE 24.12.

In this example, the dynamic responses for the flash process in Figure 24.18 with recommended control is evaluated for a step change in feed composition, with the

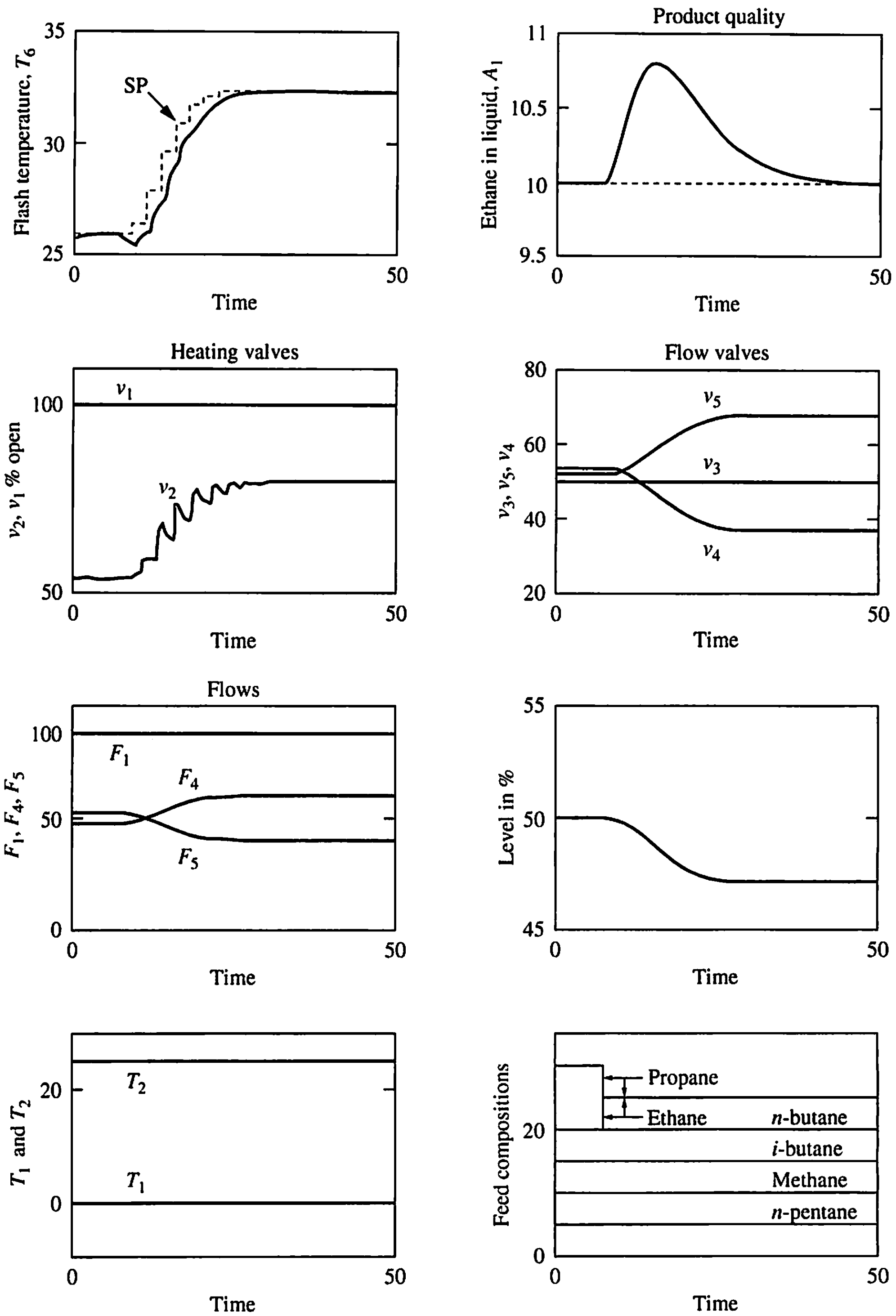


FIGURE 24.19

Transient response for final flash process and control design for the feed composition disturbance.

ethane increasing 5 mole% and the propane decreasing by the same amount. The transient behavior is shown in Figure 24.19. The ethane in the liquid is maintained within the specified limits of 10 ± 1 mole% during the transient and returns to its set point; if the analyzer feedback were removed, the ethane concentration in the liquid product would exceed the acceptable limits during the transient and at the steady state. Other cases demonstrated that the design, including analyzer feedback, could maintain the ethane in the liquid within the limits for the other disturbances defined in the control design form in Table 24.1. Thus, this process and control design is deemed acceptable.

24.11 ▫ CONCLUSIONS

Many issues must be considered in control design. Assuming that the principles in the previous parts of this book have been mastered, the challenges in design are to (1) recognize the issues important for control along with potential results and (2) develop a method for addressing the design task.

This chapter addressed the first challenge by presenting issues in the six categories of control design decisions: sensors, final elements, process design, control structure, control algorithms, and monitoring. Naturally, the issues discussed were not a complete listing of all possible items, but they included the most important issues in typical systems. The analysis of degrees of freedom and controllability were reviewed and their applications to control design demonstrated. Again, we see the importance of the process equipment design and operating conditions on process control, since these determine the operability of the system.

The control design form (CDF) was introduced in this chapter to address the second challenge. By completing the CDF, the engineer can begin the problem-solving task with a complete problem definition without prejudging possible designs. The format provides a helpful checklist with sufficient memory aids to enable the engineer to address all of the important topics.

In addition, the engineer would benefit from a road map for the analysis and decision making during the design process. Is it best to start with the sensors, with the process, or with the algorithms? This important topic is covered in the next chapter, where a sequential design method, with checks for iterations, is presented along with some additional examples.

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Quantitative reliability analysis of equipment and control strategy design is presented in the following.

Fisher, T. (ed.), "Control System Safety," *ISA Transactions, 30,* 1 (special edition), (1991).
Goble, W., *Evaluating Control System Reliability: Techniques and Applications,* Instrument Society of America, Research Triangle Park, NC, 1992.

Industrial experience and design recommendations for safety controls systems are given in the following.

Black, W., "Non-Self-Actuating Safety Systems for the Process Industries," *Hazards X. Process Safety in Fine and Specialty Chemical Plants,* Inst. Chem. Eng., London, 294–307 (1989).
Stewart, R., "High Integrity Protective Systems," *Inst. Chem. Eng. Symp. Ser. No. 34,* Inst. Chem. Eng., London, 99–104 (1971).
Wells, G., *Safety in Process Plant Design,* Wiley, New York, 1980.

The Proceedings from the series of meetings on "Hazards" organized by the Institute of Chemical Engineers, London, is a good source of up-to-date information on control for safety.

Process and control designs are based on analysis of process behavior and control structure performance. This analysis may be analytical, numerical, or semi-quantitative, depending on the type of information available and the necessary accuracy of the results. The problems in this chapter give opportunities for all three.

QUESTIONS

24.1. (*a*) When pumping or compression costs are high, incentive exists for controlling flows at minimum cost. Suggest approaches for controlling flow rates with low pressure drops across sensors and valves.

(*b*) Discuss the response of systems when the sensor, rather than the valve, fails to function properly. How can safety be ensured in such situations?

(*c*) Discuss a quick method for determining the maximum number of variables that can be controlled for a completed process design. Assume that a detailed process schematic (piping and instrumentation drawing) is available, but a detailed mathematical model is not.

(*d*) Discuss why controllability is analyzed with a linear model whereas the operating window is determined based on a nonlinear model.

(*e*) If a system is controllable and has a sufficient operating window, will all possible loop pairings provide stable dynamic performance (assuming proper constant tuning)?

(*f*) If the process dynamics are overdamped, would variable values between two steady states within the operating window remain within the window during the transient response?

24.2. (*a*) Generalize the controllability test for non-self-regulating systems.

(*b*) The definition for controllability employed in this chapter is appropriate for many, but not all, processes. Discuss other definitions (e.g., for batch processing) and define appropriate tests.

(*c*) The test for controllability requires a "square" system (i.e., one with the same number of manipulated and controlled variables). What if the number of manipulated variables is greater than (or less than) the number of controlled variables?

24.3. The design for the mixing process without reaction in Figure Q24.3 is to be analyzed. Additional information is

(1) The inlet stream consists of two components A and S with equal densities.
(2) The pressures P_1, P_2, and P_4 are determined externally.
(3) Pressure P_3 is constant.
(4) All pressure drops occur across the control valves.
(5) The tank is well mixed.
(6) The installed characteristics of the valves are linear.

(*a*) Develop a dynamic model for this process and analyze the degrees of freedom.

(*b*) Based on this drawing, determine the maximum number of variables that can be controlled (without equipment changes).

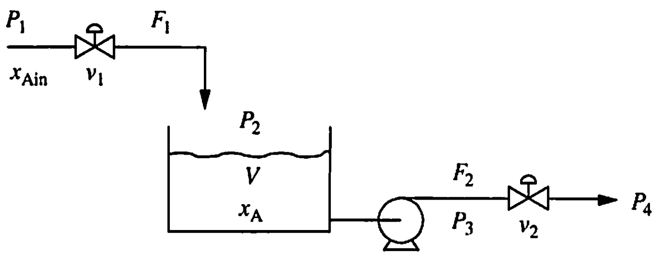

FIGURE Q24.3

(*c*) Determine whether the system is controllable when controlling the feed flow rate and the tank level with the valves shown. Add sensors and sketch the design.

(*d*) Determine whether the system is controllable when controlling the tank level and tank concentration (x_A) with the valves shown. Add sensors and sketch the design.

(*e*) Reconsider part (*d*) above with the following process modification. A new pipe and valve (v_3) are added to inject a flow of pure component A into the feed stream. All three valves are available for manipulation.

24.4. Review the excellent design problem published by Downs and Vogel (1993). Transfer this problem statement into a specification in a control design form. Can you completely translate the statement into the form? Is it easier to read in this form than in the original paper?

24.5. The chemical reactor in Figure Q24.5 has the following properties: well mixed, isothermal, constant volume, constant density. The chemical reaction occurring is A $\rightarrow$ B with the reaction rate $r_A = -kC_A$. The concentrations of the reactant and the product can be measured without delay.

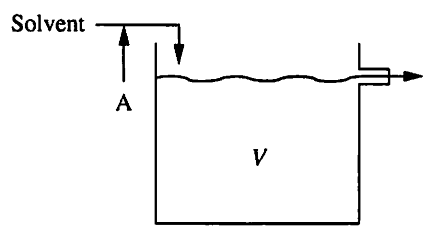

FIGURE Q24.5

(*a*) The total feed flow (F) and the feed concentration (C_{A0}) are the potential *manipulated* variables for the reactor effluent composition control. Construct a regulatory control scheme that will control these two variables (F, C_{A0}) simultaneously to independent set point values, and sketch it on the figure. You may place the sensors and final elements for these variables anywhere you think appropriate.

(*b*) Derive a dynamic model for $C_B(s)/C_{A0}(s)$. Analyze the model regarding (i) order, (ii) stability, (iii) periodicity, and (iv) step response characteristics.

(*c*) Derive a dynamic model for $C_B(s)/F(s)$. Analyze the model regarding (i) order, (ii) stability, (iii) periodicity, and (iv) step response characteristics.

(*d*) Based on the results in (*b*) and (*c*), which of these two manipulated variables would provide the best feedback control for C_B for a set point change using PI feedback control?

24.6. Part of a proposed control design for a blending process is given in Figure Q24.6. In addition, the composition of A is to be controlled by adjusting one or more flow set points. The objectives are to control the product flow tightly and the composition as tightly as possible; disturbances are in component

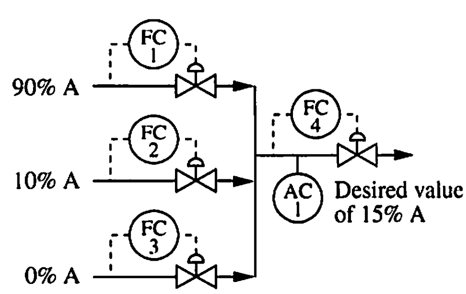

FIGURE Q24.6

compositions and changes to the product flow set point; no constraints are encountered. Critically evaluate the proposed control, make any changes to provide closed-loop flow control, and design the composition feedback control.

24.7. Discuss the need for accuracy or reproducibility for the sensors in the control designs in the following figures: 15.12, 15.13, 17.16, V.2, and 22.10.

24.8. The level process with control design in Figure Q24.8 is proposed to you. Evaluate whether the system can maintain the levels within their limits for changes in the flow from tank 2. Estimate the control performance and make changes, if required, to provide satisfactory performance.

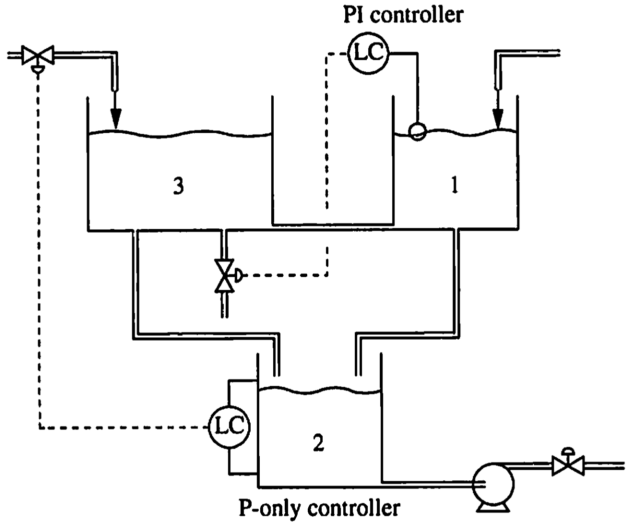

FIGURE Q24.8

24.9. The well-mixed, constant-volume chemical reactor with separator and recycle in Figure Q24.9 is considered in this question. The reactor has a single reaction A $\rightarrow$ B with $r_A = -kC_A$. The separator makes a perfect separation of the product and the pure reactant, which is recycled to the reactor feed, and the separator dynamics and transportation delays are very fast and will be assumed at quasi-steady state.

(*a*) Assume that the fresh feed rate (F_1) is controlled constant. The major disturbance is temperature, which can be taken to be a change in the reaction rate constant. Based on a dynamic model of the process, determine an analytical relationship between the disturbance and (1) the recycle flow and (2) the reactor concentration. Determine how the dynamic behavior is affected by the steady-state conversion, $(C_{A0} - C_A)/C_{A0}$.

(*b*) Discuss the factors that would influence the choice of the best reactor conversion in a typical industrial process.

(*c*) Determine a simple change to the control design that substantially reduces the effect of the disturbance (without controlling temperature).

24.10. If you have not completed questions 15.2 and 18.13, it would be worthwhile to do them now.

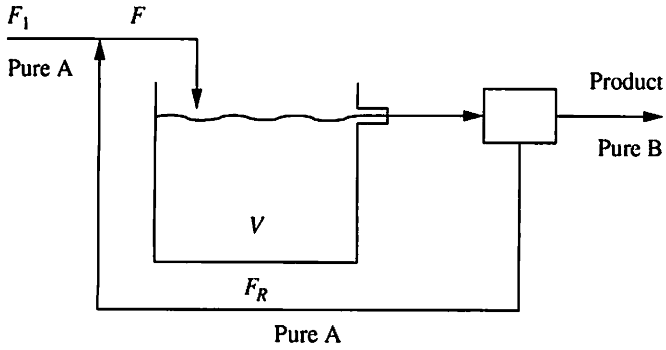

FIGURE Q24.9

24.11. The design in Figure Q24.11 is proposed for an isothermal, well-mixed CSTR with a single reaction, A $\rightarrow$ B with $r_A = -kC_A$. The main disturbance is a steplike disturbance in the feed flow rate, and real-time measurement and control of the compositions is not possible. Evaluate the control performance (i.e., the deviation of the composition) for (*a*) perfect PI control of the level (the level exactly remains at its set point at all times) and (*b*) P-only control of the level. Which approach gives a smaller deviation for the compositions from the initial conditions at the final steady state? Does your answer depend on the tuning of the P-only controller? If so, what is the best value of the controller gain?

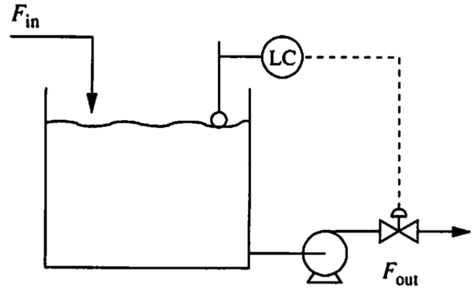

FIGURE Q24.11

24.12. Given the process schematic in Figure Q24.12 and the following data, determine the heat transfer coefficients for the three heat exchangers, and explain the assumptions you made in performing the calculations. If you performed this analysis over a long period of time, what useful information would you determine?

$T_1 = 20, T_2 = 42, T_3 = 45, T_4 = 68, T_5 = 76, T_6 = 88, T_7 = 71,$
$T_8 = 75, T_9 = 31°C$
$F_1 = 50, F_2 = 50, F_3 = 56, F_4 = 150$ m^3/h
$\rho = 0.8 \times 10^6$g/m^3 and $C_p = 0.75$ cal/(g °C) for the streams measured by F_1 and F_2
$\rho = 1.0 \times 10^6$g/m^3 and $C_p = 1.00$ cal/(g °C) for the stream measured by F_3
$\rho = 0.75 \times 10^6$g/m^3 and $C_p = 0.71$ cal/(g °C) for the stream measured by F_4

24.13. Expand the following designs by adding (i) alarms, (ii) safety valves, (iii) final element failure directions, (iv) emergency safety controls, and (v) sensors for monitoring process performance. You may add sensors or final elements as necessary; also, specify whether each sensor should provide a signal that is highly accurate, or merely highly reproducible.

(*a*) The fired heater in Figure 17.17
(*b*) The distillation column in Figure 21.14

24.14. The CST mixing process in Figure Q24.14 is proposed. The goal is to control the effluent composition and temperature. Evaluate the design, suggest

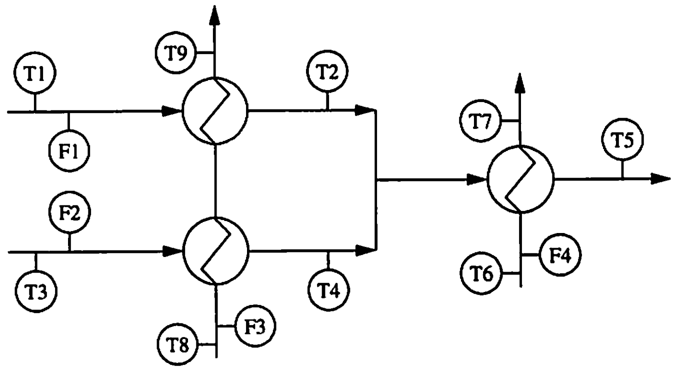

FIGURE Q24.12

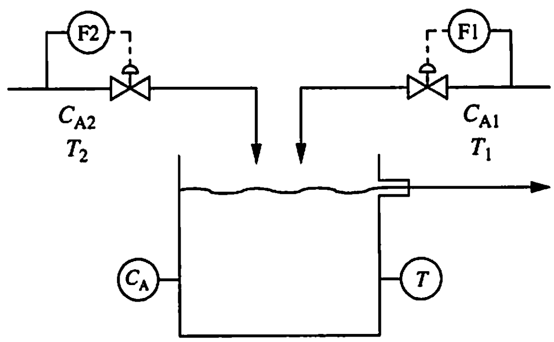

FIGURE Q24.14

a proper loop pairing, and suggest process or control objective modifications, if necessary.

24.15. A gas distribution system for a chemical plant is shown in Figure Q24.15. Several processes in the plant produce gas, and this control strategy is not allowed to interfere with these units. Also, several processes consume gas, and the rate of consumption of only one of the processes can be manipulated by the control system. The flows from producers and to consumers can change rapidly. Extra sources are provided by the purchase of fuel gas and vaporizer, and an extra consumer is provided by the flare. The relative dynamics, costs, and range of manipulation are summarized in Table Q24.15.

(*a*) Complete the blank entries in the Dynamics column in Table Q24.15.

(*b*) Design a control strategy to satisfy the objectives of tight pressure control and minimum fuel cost. You may add sensors as required but make no other changes.

(*c*) Suggest process change(s) to improve the performance of the system. (Hint: Before designing the controls, determine the correct response for all valves as the ratio of producing to consuming gas flows changes from much greater than 1.0 to much less than 1.0.)

24.16. The dynamics of an isothermal, constant-volume, constant-feed flow rate, well-mixed CSTR are to be evaluated for feedback control in this question.

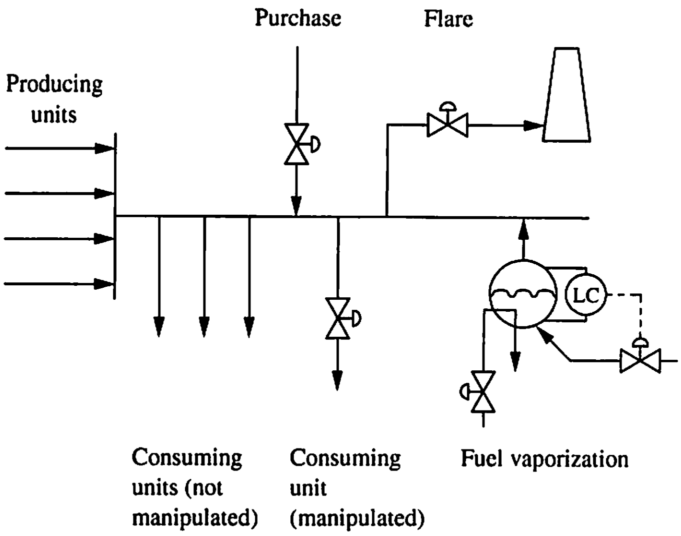

FIGURE Q24.15

TABLE Q24.15

Flow	Manipulated	Dynamics	Range (% of total flow)	Cost
Producing	No	Fast	0–100%	n/a
Consuming	Only one flow	Fast	0–20%	Very low
Generation	Yes		0–100%	Low
Purchase	Yes		0–100%	Medium
Disposal	Yes		0–100%	High

The feed consists entirely of component A, the chemical reaction is

$$A \underset{k_B}{\overset{k_A}{\rightleftharpoons}} B$$

and the rates are first-order for both directions.

(*a*) Derive the dynamic model of the input-output system between C_{A0} and C_A. What conclusions can be determined regarding stability, periodicity, and either overshoot or inverse response for a step input? Describe the expected control performance for a step set point change. What tuning method could be used for a PID controller? Would you recommend feedforward control to improve the performance for a disturbance in temperature?

(*b*) Answer questions (*a*) for the input-output system C_{A0} and C_B.

(*c*) Would you expect that the control performance between $C_{A0} \rightarrow C_A$ would be better, the same, or worse than $C_{A0} \rightarrow C_B$, assuming that the feedback controllers were tuned on the same basis? Base your answer solely on the relative dynamics for the two possible systems. Consider a step set point change.

24.17. The process in Figure Q24.17 is a simplified head box for a paper-making process. The control objectives are to control the pressure at the bottom of the head box tightly and to control the slurry level within a range.

(*a*) Derive a model for the effects of the two inlet flows on the controlled variables.

(*b*) Design a control system by pairing the controlled and manipulated variables. Use the methods introduced in Chapters 20 and 21 as well as this chapter. Discuss the performance of your design and any special features that should be included in the implementation.

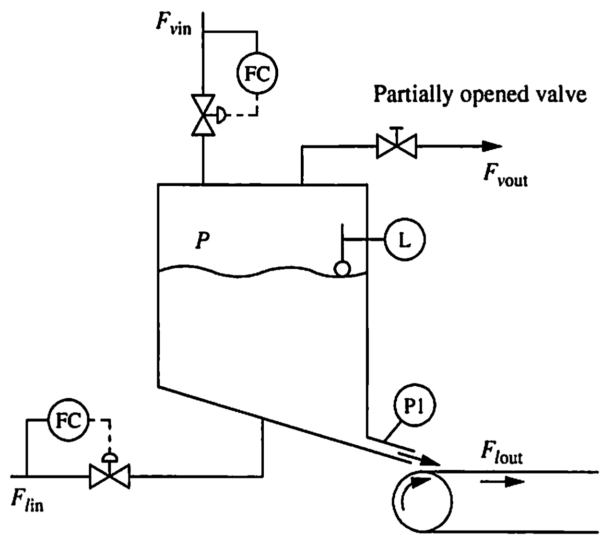

FIGURE Q24.17

24.18. The process in Figure Q24.18 includes a fired heater, chemical reactor, and heat exchangers to recover energy by heat transfer to other processes in the plant. The goals are to have tight flow control (F1), tight control of the reactor outlet temperature (T2), and good control of temperatures T3 and T4 in the integrated processes. The sensors and manipulated variables are shown in the figure. Disturbances are set point changes to the process flow and changes in the heating requirements of the heat-integrated processes.

(*a*) Without changing the instrumentation and process equipment, design a control system to achieve the objectives. Discuss whether all objectives can be achieved and if not, why.

(*b*) By making the minimum changes to the process equipment and instrumentation, design a system that improves the result in (*a*).

24.19. The mixing process in Figure Q24.19 involves a tank to mix components A and B without chemical reaction. The effluent from the mixing tank is blended with a stream of component C, and the flow of this stream is wild; that is, it cannot be adjusted by this control strategy. Note that waste is to be minimized.

(*a*) Using only the equipment shown in the figure, design a control system to tightly control the percentages of A, B, and C in *the blended product*. Can you achieve this and also control the total flow of blended product?

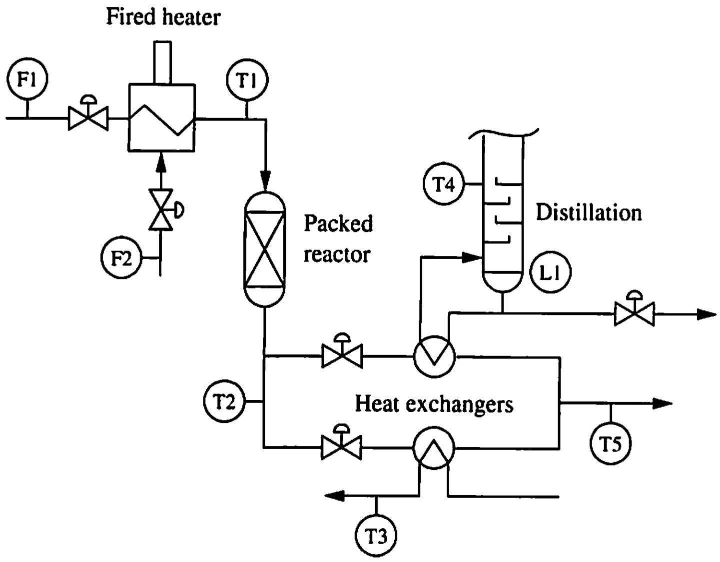

FIGURE Q24.18

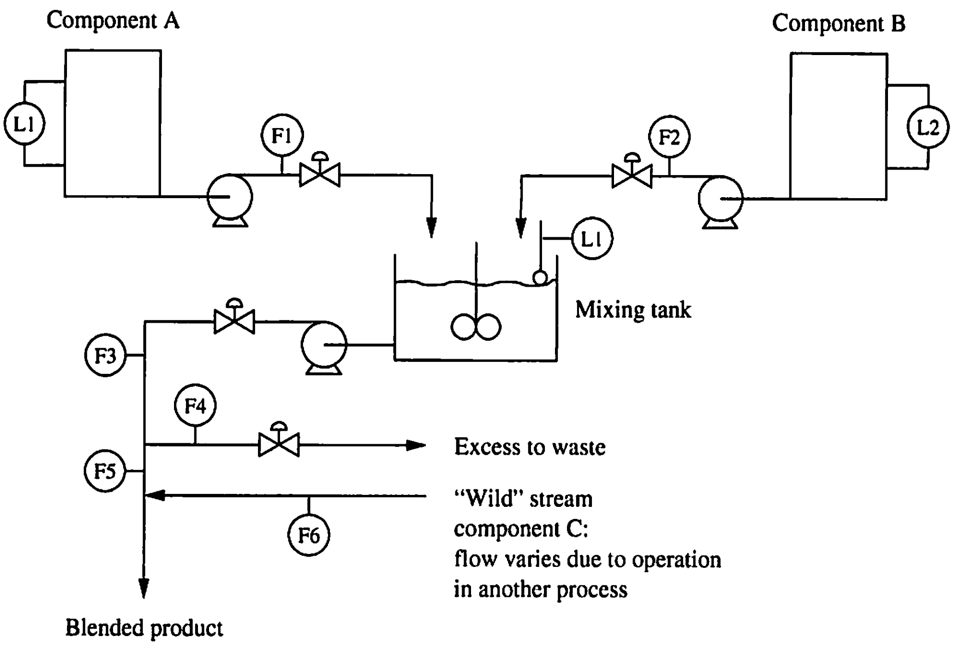

FIGURE Q24.19

(*b*) Improve your result in (*a*) by adding an analyzer that can measure compositions in one stream. Decide the proper location and use it in the control system. Discuss why the analyzer would improve the performance.

24.20. A heat exchanger is shown in Figure Q24.20. The temperature measured by the sensor is to be controlled. Design four different control strategies to control this temperature and discuss the differences. Select the design that would give the best control performance, and discuss the reasons why.

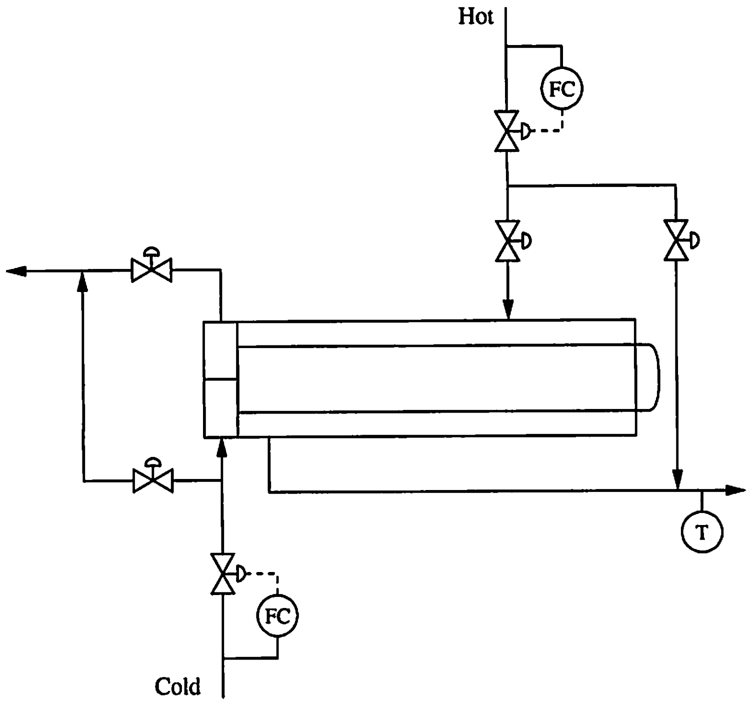

FIGURE Q24.20

24.21. Apply the Niederlinski criterion to the flash control system presented in this chapter. Discuss your results and the interpretation of the control system design.

24.22. Control of the flash process analyzed in this chapter involved partial control. All six components in both product streams would influence product quality and profit, but only one dominant variable, mole fraction ethane in the liquid, was controlled.

(*a*) Discuss the final steady-state deviations of all compositions from their initial steady-state values for the control system developed in this chapter. Consider each of the disturbances in the control design form separately.

(*b*) An alternative dominant variable, the flash temperature (T_6) could have been selected rather than liquid mole fraction ethane. Discuss the advantages and disadvantages of T_6 as a dominant variable (not reset by A_1).

(*c*) Discuss process and control modifications needed to control the percent propane in the vapor product *in addition* to all controlled variables in the original control design.

24.23. The dynamic responses of the heat transfer process in Figure Q24.23 are considered here. The medium in the coil and jackets is heating the fluid in the tanks. The tanks are well mixed, and all transportation delays are small compared with the time constants.

(*a*) Describe the dynamic responses of each temperature to a step change in each valve.

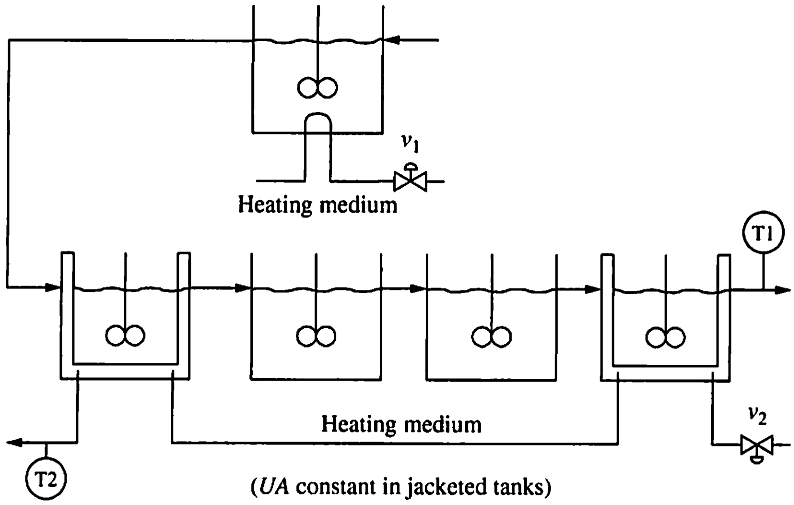

FIGURE Q24.23

(*b*) Discuss the likely control performances for each input-output pair for multiloop control.
(*c*) Assume that one feedback PI controller, with pairing T2 $\rightarrow v_2$, is in operation and that a step change is made to v_1. Describe the dynamic response of both temperatures.
(*d*) Discuss the likely control performance if the other control loop, T1 $\rightarrow v_1$, is closed.

24.24. The series of well-mixed stirred-tank chemical reactors for a first-order chemical reaction with negligible heat of reaction is shown in Figure Q24.24. Each reactor has a mass in the tank, which has the same temperature as the liquid and represents substantial energy accumulation. (This is a simplified representation of a packed bed, with the masses being the catalyst.) The concentration of the effluent from the last reactor is controlled by adjusting the heating medium valve.
(*a*) Discuss the effect of the masses on the control performance in response to feed stream temperature variations.
(*b*) Discuss the effect of the masses on the control performance in response to feed stream concentration variations.
(*c*) Draw general conclusions about the effects of the masses on the disturbance responses in (*a*) and (*b*) and on set point changes.

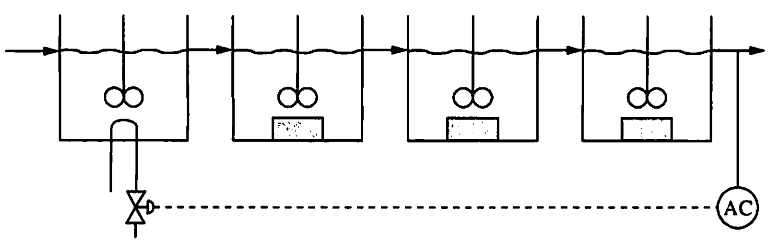

FIGURE Q24.24

24.25. A two-product distillation tower with a single feed is considered in this question.

(*a*) Sketch two types of condensers (showing equipment and valves), describe their physical principles, and be sure to explain how the duty is adjusted for process control.

(*b*) Repeat (*a*) for reboilers.

(*c*) Discuss why distillation towers typically have an overhead liquid accumulator.

(*d*) Discuss why temperatures and pressures are measured on selected distillation trays.

(*e*) Where would you place safety valves and for what maximum flow should they be sized?

(*f*) Identify possible constraints (i.e., items that define the frame of the operating window).

(*g*) Identify potential disturbances.

(*h*) The composition sensors often provide new measurements only every two minutes. Give a reason why this might occur.

(*i*) Identify all inventories in the distillation process, determine which are non-self-regulatory, and describe potential control strategies for each.

(*j*) Discuss options for product quality control and the interaction between inventory and product quality controls.

24.26. The process in Figure Q24.26 involves the chemical reaction with the overall stoichiometry of 3A + B → C taking place in a packed-bed reactor. The inlet temperature has a strong effect on the rate of reaction, and there is no limit to any reasonable value for the bed temperature. The unreacted A is separated in a flash drum and sent to fuel at considerable cost because it cannot be recycled. Also, high temperatures tend to degrade the catalyst. The liquid product has a target of 80% product C. Design a control system to achieve the objectives just described, specify sensors, and sketch the design on the drawing.

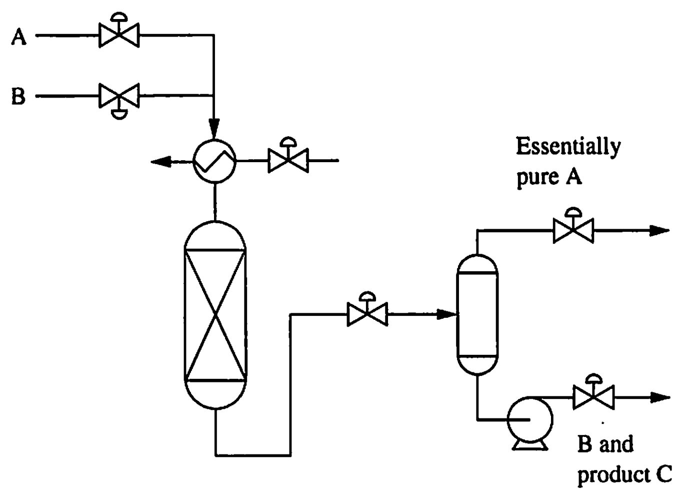

FIGURE Q24.26

24.27. For the fired heater in Figure Q24.27 (i) no change is allowed for the final elements, (ii) sensors must be added, and (iii) the feed rate and product outlet temperature must be controlled. Briefly state a reasonable set of control objectives and design a control strategy.

(*a*) Briefly give the algorithm and purpose for each controller.
(*b*) Sketch the strategy on the diagram.
(*c*) Give the failure positions for the final elements.

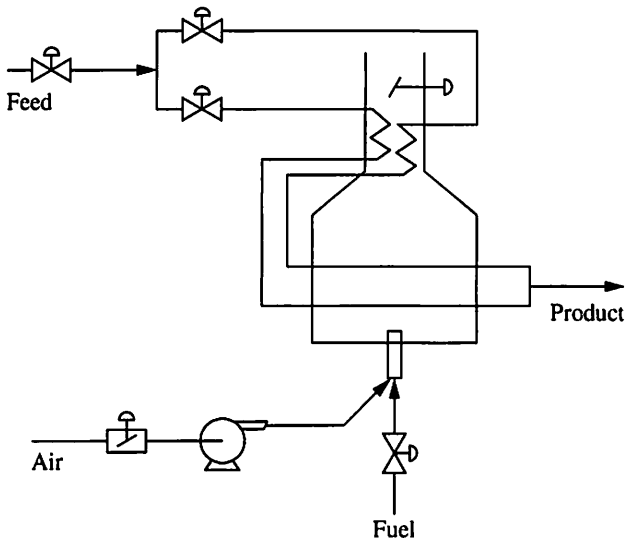

FIGURE Q24.27

24.28. The series of two chemical reactors described in Example I.2 is the initial process upon which this question is based. You may use all results from the modelling in Example I.2 without proving; simply cite the source of the equations.

(*a*) The solvent flow and composition at the inlet to the first reactor are to be controlled by two single-loop controllers. Add sensors and final elements as required and sketch the control system.
(*b*) Given this strategy is functioning perfectly (maintaining C_{A0} *constant*), determine the model between the solvent flow and the concentration of the reactant in the second reactor, C_{A2}, and comment on the expected composition (C_{A2}) control performance using this manipulated-controlled variable pairing.
(*c*) Compare with the control performance in Example 13.8.