Chapter 20

Plantwide Controllability Assessment

20.0 OBJECTIVES

In this chapter, the importance of considering controllability and operability issues early in the design process is demonstrated by showing how controllability considerations can help to differentiate between processes that are easy and processes that are difficult to control. The chapter provides a recommended methodology to initiate the design of attractive plantwide control systems.

After studying this chapter, the reader should be able to:

1. Identify potential control problems in a process flowsheet.
2. Classify and select controlled and manipulated variables for a plantwide control system.
3. Perform a conceptual synthesis of plantwide control structures (pairings) based on degrees-of-freedom analysis and qualitative guidelines.

20.1 INTRODUCTION

The design of a continuous chemical process is usually carried out at steady state for a given operating range, assuming that a control system can be designed to maintain the process at the desired operating level and within the design constraints. However, unfavorable process static and dynamic characteristics could limit the effectiveness of the control system, leading to a process that is unable to meet its design specifications. A related issue is that alternative designs usually are judged based on economics alone without taking controllability into account. This may lead to the elimination of easily controlled but slightly less economical alternatives in favor of slightly more economical designs that may be extremely difficult to control. It is becoming increasingly evident that design based on steady-state economics alone is risky because the resulting plants are often difficult to control (i.e., inflexible, with poor disturbance-rejection properties), resulting in off-specification product, excessive use of energy, and associated profitability losses.

Consequently, there is a growing recognition of the need to consider the controllability and resiliency (C&R) of a chemical process during its design. Controllability can be defined as the ease with which a continuous plant can be held at a specific steady state. An associated concept is switchability, which measures the ease with which the process can be moved from one desired stationary point to another. Resiliency measures the degree to which a processing system can meet its design objectives despite external disturbances and uncertainties in its design parameters. Clearly, it would be greatly advantageous to be able to predict as early as possible in the design process how well a given flowsheet meets these dynamic performance requirements.

Table 20.1 summarizes the four main stages in the design of a chemical process. In the conceptual and preliminary stages, a large number of alternative process flowsheets in the steady state (SS) are generated. Subsequent stages involve more detailed analysis in the SS, followed by the testing of the dynamic (Dyn) performance of the controlled flowsheets. Here, considerably more engineering effort is expended than in the preliminary stages. Therefore, far fewer designs are considered with many of the initial flowsheets having been eliminated from further consideration by screening in the preliminary stages.

The need to account for the controllability of competing flowsheets in the early design stages is an indication that simple screening measures using the limited information available should be employed to select from among the flowsheets. Here, if high-fidelity, closed-loop, dynamic modeling were used, the engineering effort and time required for development and analysis would slow the design process significantly. The right-hand columns in Table 20.1 show that the shortcut C&R tools provide a bridge between SS simulation for process design and the rigorous dynamic simulation required to verify switchability and other attributes of the closed-loop dynamics of the final design. ASPEN PLUS, PRO/II, ASPEN HYSYS, UniSim® Design, and CHEMCAD are commonly used simulation packages, all of which enable both SS and dynamic simulation.

In the following examples, the impact of design decisions on controllability and resiliency is introduced for four processes. The supplement to this chapter (Section 20S in the file Supplement_to_Chapter_20.pdf in the PDF Files folder, which can be downloaded from the Wiley Web site associated with this book) expands upon this introduction and shows how to eliminate the less desirable alternatives and validate the performance of the most promising designs.

Table 20.1 Process Design Stages, Issues, and Tools

| Design Stage	Issues	What Gets Fixed	Tools		
			SS	C&R	Dyn
1. Process creation	Selecting between alternative material pathways and flowsheets	Material pathways			
2. Development of base-case design	Feasibility studies based on fixed material pathways Unit operations selection Heat integration superstructure	Flowsheet structure			
3. Detailed design	Optimization of key process variables Analysis of process sensitivity to disturbances and uncertainties	Optimal flowsheet parameters			
4. Plantwide controllability assessment	Flowsheet controllability Dynamic response of the process to disturbances Selection of the control system structure and its parameters	Control structure and its parameters			

EXAMPLE 20.1 *Heat Exchanger Networks*

The network shown in Figure 20.1a, which was introduced by McAvoy (1983), cools hot stream 1 from 500 to 300°F using cold streams 2 and 3 having feed temperatures of 300 and 200°F and corresponding target temperatures of 371.4 and 400°F, respectively, with the heat-capacity flow rates in MMBtu/hr-°F. Furthermore, the feed rate and temperature of the hot stream are considered as disturbances.

(a)

(b)
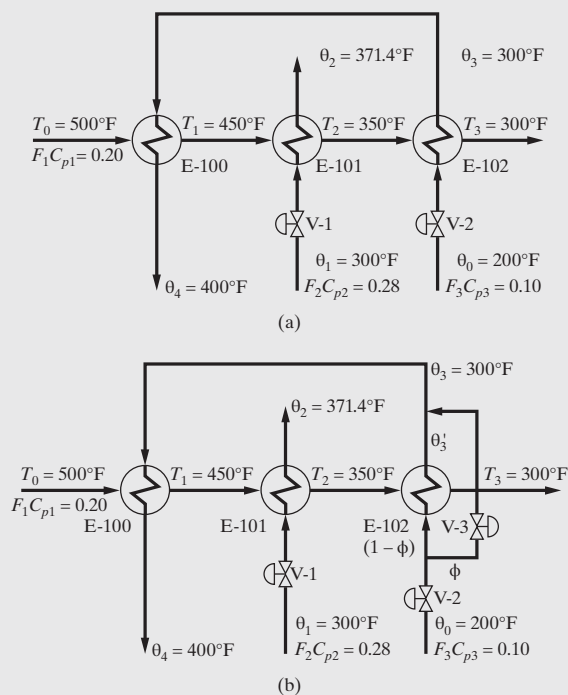
Figure 20.1 Heat exchanger network: (a) original configuration; (b) modification with bypass.

As shown in Figure 20.1a, two of the target temperatures can be controlled by manipulating the flow rates of the two cold streams. This means that one of the target temperatures is left uncontrolled in the face of disturbances in the hot stream. An alternative design, involving a bypass around exchanger E-102, is illustrated in Figure 20.1b. As shown, this simple modification allows all three target temperatures to be regulated. Because the selection of the appropriate bypass flow fraction, ϕ, and of the most effective control configuration is not trivial, controllability analysis should be carried out on the alternative networks and their candidate control structures. This will assist in selecting one of the two designs as shown in the supplement to this chapter.

EXAMPLE 20.2 *Heat-integrated Distillation Columns*

The production of methanol is carried out in a moderate-pressure synthesis loop by the direct hydrogenation of carbon dioxide,

$$CO_2 + 3H_2 \rightleftharpoons CH_3OH + H_2O, \qquad (20.1)$$
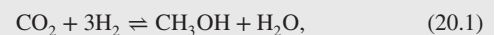
which generates a liquid product that contains a binary mixture of methanol and water in approximately equal proportions. To provide commercial methanol that is nearly free of water, dehydration is achieved commonly by distillation. To reduce the sizable energy costs, three double-effect, heat-integrated configurations shown in Figure 20.2 are commonly considered as alternatives to a single distillation column (SC):

Feed Split (FS)

The feed is split nearly equally ($F_H \approx F_L$) between two columns to achieve optimal operation. The overhead vapor product of the high-pressure column supplies the heat required in the low-pressure column.
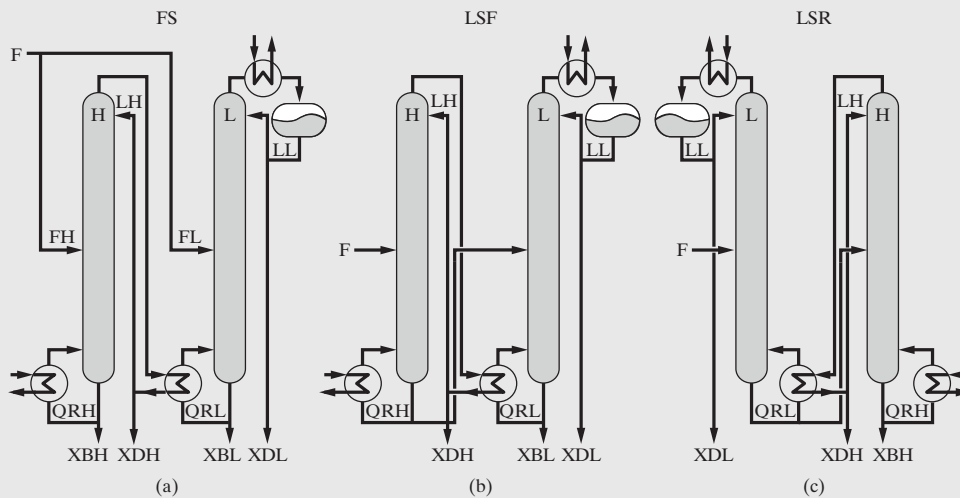
Figure 20.2 Three heat-integrated alternatives to a single distillation column.

Light-split/Forward Heat Integration (LSF)

The entire feed is fed to the high-pressure column. About half of the methanol product is removed in the distillate from the high-pressure column, and the bottoms product is fed into the low-pressure column. In this configuration, heat integration is in the same direction as the mass flow.

Light-split/Reverse Heat Integration (LSR)

The entire feed is fed to the low-pressure column with the bottoms product from the low-pressure column fed into the high-pressure column. Here, heat integration is in the opposite direction to that of the mass flow.

As discussed in Section 11.8, Multiple-Effect Distillation, these configurations reduce the energy costs by using the heat of condensation of the overhead stream from the high-pressure column (H) to supply the heat of vaporization of the boilup in the low-pressure column (L). Although they are more economical, assuming steady-state operation, they are more difficult to control because the configurations (1) are more interactive and (2) have one less manipulated variable for process control because the reboiler duty in the low-pressure column can no longer be manipulated independently.

To show the energy savings, the four flowsheets were simulated on the basis of an equimolar feed of 2,700 kmol/hr, producing 96 mol% methanol in the distillate and 4 mol% methanol in the bottoms product, assuming 75% tray efficiency and no heat loss to the surroundings as well as using UNIFAC to estimate the liquid-phase activity coefficients. The total energy requirements for the four alternatives were computed as follows:

SC: 2.12×10^7 kcal/hr   LSR: 1.23×10^7 kcal/hr
LSF: 1.33×10^7 kcal/hr   FS: 1.23×10^7 kcal/hr

Clearly, the LSR and FS configurations save the most energy, although the energy consumption in the LSF configuration is only 8% higher. Based on steady-state economics alone, one of these three configurations would be selected. However, disturbance resiliency analysis (Chiang and Luyben, 1988; Weitz and Lewin, 1996) shows that either the LSR or LSF configurations are preferred for disturbance

rejection because they provide performance only slightly worse than that of a single column, SC. The FS configuration, on the other hand, does considerably worse. The supplement to this chapter shows how to obtain this information when selecting from among these alternatives.

EXAMPLE 20.3 *Heat Recovery from an Exothermic Reactor*

Often, the heat from an exothermic reactor is used to preheat the reactor feed, thus saving energy, as discussed in Section 6.5. Figure 20.3b shows a configuration using a feed/product heat exchanger that is commonly preferred to the configuration with independent preheat in Figure 20.3a. However, the heat-integrated configuration shares the same disadvantages as the heat-integrated distillation systems discussed in the Example 20.3 (i.e., one less manipulated variable and possibly unfavorable dynamic interactions). Furthermore, the feed-effluent heat exchanger introduces positive feedback and the possibility of thermal runaway.
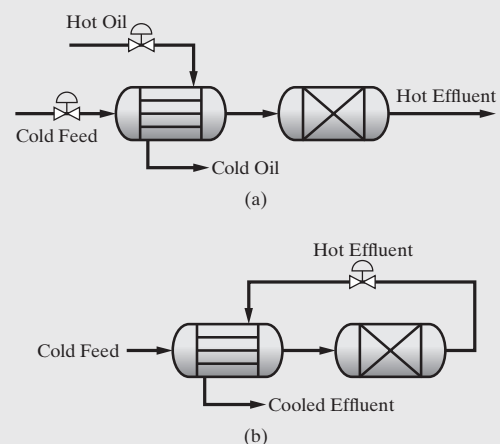

Figure 20.3 Two configurations for an exothermic reactor requiring feed preheating: (a) reactor with independent preheat; (b) heat-integrated (autothermal) reactor.

EXAMPLE 20.4   *Reactor-flash-recycle System*

Reactor design for complete conversion may be impossible thermo-dynamically or undesirable because of reduced yields when byproducts are formed. In such cases, an economic alternative is to design a combined reactor-separator-recycle system as illustrated in the simple example in Figure 20.4. Here, the reaction A → B is carried out in a CSTR whose liquid feed is a stream containing pure A. In the event that B is sufficiently more volatile than A, the separation can be performed using a flash vessel with unreacted A recycled to the reactor. As will be seen in the plantwide control examples at the end of this chapter and in the quantitative analysis in the supplement to this chapter, the presence of the recycle complicates control of the process and requires special attention.
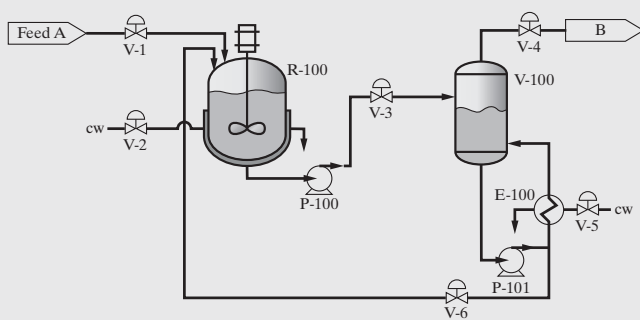

Figure 20.4 Reactor-flash-recycle system for the production of B.

To enable the evaluation of the controllability and resiliency of alternative process configurations, it is important to consider two aspects of the design of plantwide control systems:

1. The classification and selection of controlled and manipulated variables

2. The qualitative synthesis of plantwide control structures based on degrees-of-freedom analysis and qualitative guidelines.

These are examined in the next two sections.

20.2   CONTROL SYSTEM CONFIGURATION

The design of a control system for a chemical plant is guided by the objective to maximize profits by transforming raw materials into useful products while satisfying product specifications, safety constraints, operational limitations, and environmental regulations. All four constraints require special consideration.

1. *Product Specifications*. To satisfy customer expectations, it is important that product quality and production rate meet specifications. This has been the driving force for the implementation of online, optimal process control in the chemical industry. More recently, statistics-based approaches such as six-sigma methodologies have been harnessed for this purpose, as discussed in Chapter 18.

2. *Safety Constraints*. The plant must be operated safely to protect the well-being of plant personnel, the plant itself, and nearby communities. As an example, a typical safety-driven constraint requires that the temperature and pressure of a steel vessel not exceed upper limits dictated by metallurgy. For other examples, see Section 3.6.

3. *Operational Limitations*. Examples of these are upper and lower bounds on the vapor velocity in distillation columns to avoid flooding and weeping, respectively, and upper bounds on the reactor temperatures to prevent degradation of the catalyst or the onset of undesirable side reactions.

4. *Environmental Regulations*. These require that processing plants comply with constraints on air and water quality as well as waste disposal. Many examples are discussed in Section 3.4.

Classification of Process Variables

When designing a plantwide control system, it is common to view the process in terms of its input and output variables. These variables include flow rates of streams entering and leaving process equipment and temperatures, pressures, and compositions in entering and leaving streams and/or within equipment.

Process *output variables* are those that give information about the state of the process. They are usually associated with streams leaving the process or with measurements inside a process vessel. When designing a control system, output variables are usually referred to as *controlled variables*, which are measured (online or off-line).

Process *input variables* are independent variables that affect the output variables of a process. They can be subdivided into two categories: (1) *manipulated variables* (also called *control variables*), which can be adjusted freely by an operator or a control mechanism, and (2) *disturbance variables* (also called externally defined variables), which are subject to the external environment and thus cannot be controlled. These variables are associated typically with the inlet and outlet streams. In a control system, manipulated variables cause changes to controlled variables.

There are three main reasons why it may be impossible to control all of the output variables of a process.

1. It may not be possible to measure online all of the output variables, especially compositions. Even when it is possible, it may be too expensive to do so.

2. By a degrees-of-freedom analysis, described later, there may not be enough manipulated variables available to control all of the output variables.

3. Potential control loops may be impractical because of slow dynamics, low sensitivity to the manipulated variables, or interactions with other control loops.

Qualitative criteria have been suggested by Newell and Lee (1988) to guide the selection of controlled and manipulated variables that are suitable for an initial configuration of a plantwide

control system. These guidelines, which are presented next, are driven by the plant and control objectives and should not be applied without due consideration. When two guidelines conflict, the most important of the two should be adopted. In critical cases, the more reliable quantitative screening approaches discussed in the supplement to this chapter should be considered. Following presentation of the guidelines, examples of the selection of variables are given.

Selection of Controlled (Output) Variables

Guideline 1:   *Select Output Variables That Are Either Non–self-regulating or Unstable.*   A *self-regulating process* is one that is described by a state equation of the form $\dot{x} = f\{x, u\}$ where x is an output variable and u is an input variable. A change in u will result in the process moving to a new steady state. A *non-self-regulating process* is described by $\dot{x} = f\{u\}$. As a result, changes in the input variable, u, affect the process output as a pure integrator. An example of a non-self-regulating output variable is the liquid level of a surge tank whose effluent feeds a pump followed by a control valve. Clearly, if the control valve is left uncontrolled, a positive feed disturbance to the surge drum may cause the vessel to overflow. When the process is *unstable* in the open loop (that is, in the absence of feedback control), a change in the input variable causes the system to become unstable. Clearly, non-self-regulating and unstable process output variables must be selected as controlled variables.

Guideline 2:   *Choose Output Variables That Would Exceed the Equipment and Operating Constraints without Control.*   Clearly, when safety or operational constraints are imposed, it is important to measure and control these output variables to comply with the constraints.

Guideline 3:   *Select Output Variables That Are Direct Measures of the Product Quality or That Strongly Affect It.*   Examples of variables that are a direct measure of the product quality are the composition and refractive index whereas those that strongly affect it are temperature and pressure.

Guideline 4:   *Choose Output Variables That Exhibit Significant Interactions with Other Output Variables.*   Plantwide control must handle the potential interactions in the process. Improved closed-loop performance is achieved by stabilizing output variables that interact significantly with each other.

Guideline 5:   *Choose Output Variables That Have Favorable Static and Dynamic Responses to the Available Manipulated Variables.*   This guideline assists in the selection between alternative output variables that match up on the other guidelines.

Selection of Manipulated Variables

Guideline 6:   *Select Manipulated Variables That Significantly Affect the Controlled Variables.*   For each control loop, select an input variable with as large a SS gain as possible and sufficient range to adjust the controlled variable. For example, when a distillation column operates with a large reflux ratio, that is, values greater than 4 (Luyben et al., 1999), it is much easier to control the level in the reflux drum using the reflux flow rate rather than the distillate flow rate.

Guideline 7:   *Select Manipulated Variables That Rapidly Affect the Controlled Variables.*   This precludes the selection of inputs that affect the outputs with large delays or time constants.

Guideline 8:   *Select Manipulated Variables That Affect the Controlled Variables Directly rather than Indirectly.*   For example, when appropriate for the design of an exothermic reactor, it is preferable to inject a coolant directly rather than use a cooling jacket.

Guideline 9:   *Avoid Recycling Disturbances.*   It is usually better to eliminate the effect of disturbances by allowing them to leave the process in an effluent stream rather than having them propagate through the process by the manipulation of a feed or recycle stream.

Selection of Measured Variables

Both input and output variables may be measured variables with online measurement preferred to off-line measurement. Seborg et al. (1989) discuss the importance of measurements in control and provide three guidelines for the selection of variables to be measured and the location of the measurements.

Guideline 10:   *Reliable, Accurate Measurements Are Essential for Good Control.*   An example of a poorly designed measurement would be an orifice positioned to measure flow rate with an insufficient entry length of piping.

Guideline 11:   *Select Measurement Points That Are Sufficiently Sensitive.*   Consider, for example, the indirect control of the product compositions from a distillation column by the regulation of a temperature near the end of the column. In high-purity distillation columns with almost flat terminal temperature profiles, it is preferable to move the temperature measurement point closer to the feed tray.

Guideline 12:   *Select Measurement Points That Minimize Time Delays and Time Constants.*   Large time delays and dynamic lags in the process limit the achievable closed-loop performance. These should be reduced whenever possible in the process design and the selection of measurements.

Degrees-of-Freedom Analysis

Before selecting the controlled and manipulated variables for a control system, one must determine the number of manipulated variables permissible. As discussed in Section 7.2, the number of manipulated variables cannot exceed the number of degrees of freedom, which are determined using a process model according to

$$N_D = N_V - N_E \qquad (20.2)$$

where N_D is the number of degrees of freedom, N_V is the number of process variables, and N_E is the number of independent equations that describe the process. However, the number of manipulated variables, N_M, is generally less than the number of degrees of freedom because one or more variables may be externally defined (i.e., N_{ED}, the disturbances); that is, $N_D = N_M + N_{ED}$. Consequently, the number of manipulated variables can be expressed in terms of the number of externally defined variables:

$$N_M = N_V - N_{ED} - N_E \qquad (20.3)$$

The number of manipulated variables equals the number of controlled variables that can be regulated. When a manipulated variable is paired with a regulated-output variable, its degree of freedom is transferred to the output's setpoint, which becomes the new independent variable.

Next, degrees-of-freedom analyses are carried out and their implications for control system design are considered for heat exchanger networks, jacketed stirred-tank reactors, a utility system, a flash vessel, and a distillation column.

EXAMPLE 20.5   *Control Configurations for Heat Exchanger Networks (Example 20.1 revisited)*

Referring to Figure 20.1a, the process can be described in terms of 15 variables: F_1, F_2, F_3, T_0, T_1, T_2, T_3, θ_0, θ_1, θ_2, θ_3, θ_4, Q_1, Q_2, and Q_3. Of these, assume that four variables can be considered to be externally defined: F_1, T_0, θ_0; and θ_1. A steady-state model for the process consists of three equations for each heat exchanger. For example, for the first heat exchanger, the following equations apply:

$$Q_1 = F_1 C_{P1}(T_0 - T_1) \qquad (20.4)$$

$$Q_1 = F_3 C_{P3}(\theta_4 - \theta_3) \qquad (20.5)$$

$$Q_1 = U_1 A_1 \frac{(T_0 - \theta_4) - (T_1 - \theta_3)}{\ln\left\{\dfrac{T_0 - \theta_4}{T_1 - \theta_3}\right\}} \qquad (20.6)$$

In these equations, Q_i, U_i, and A_i are the heat duty, heat-transfer coefficient, and heat-transfer area, respectively, for heat exchanger i. Values for the latter two are assumed known, so they are not process variables. Similar equations are written for the other two heat exchangers, making a total of nine equations. Consequently, the number of manipulated variables is computed: $N_M = N_V - N_{ED} - N_E = 15 - 4 - 9 = 2$

Thus, two variables can be manipulated. Two candidates are the flow rates of the two cold streams: F_2 and F_3. Ideally, for the selection of the controlled variables, it would be desirable to regulate

all three target temperatures: T_3, θ_2, and θ_4. However, with only two manipulated variables, only two controlled variables can be selected. The guidelines presented above are insufficient to select which two of the three should be picked because all three provide a direct measure of the product quality (Guideline 3), and there are clearly significant interactions among all three of the variables (Guideline 4). Without quantitative analysis, one cannot gauge which of the three have the most favorable static and dynamic responses to the manipulated variables. If only T_3, θ_2, and θ_4 are considered as potential controlled variables, three possible control systems should be investigated. As an illustration, Figure 20.5 shows one possible configuration of two control loops. One loop adjusts the flow rate, F_2, to control θ_2, and the other loop adjusts F_3 to control θ_4. An alternative configuration with reversed pairings (i.e., $\theta_4 - F_2$, $\theta_2 - F_3$) is unstable as shown in Case Study 20S.2.

The design in Figure 20.1b involving a bypass on exchanger E-102 permits the regulation of all three target temperatures. In this case, the number of variables is increased by 2 (the bypass flow fraction, ϕ, and the temperature, θ_3'), giving a total of 17 variables with the same four disturbance variables. For constant heat capacities and no phase change, the process is modeled by one additional energy balance for the mixer,

$$\theta_3 = (1 - \phi)\theta_0 + \phi\theta_3' \qquad (20.7)$$

where ϕ is the E-102 bypass flow fraction and θ_3' is the temperature leaving heat exchanger E-102. Because $N_M = N_V - N_{ED} - N_E = 17 - 4 - 10 = 3$, three variables can be manipulated, namely, F_2, F_3, and ϕ. The flow rate of the second cold stream, F_3, affects two of the three heat exchangers whereas F_2 affects only the second one directly, and ϕ affects T_3 directly (Guidelines 6, 7, and 8). The control structure shown in Figure 20.6 is the most resilient and controllable regulatory structure as is demonstrated in Case Study 20S.2.
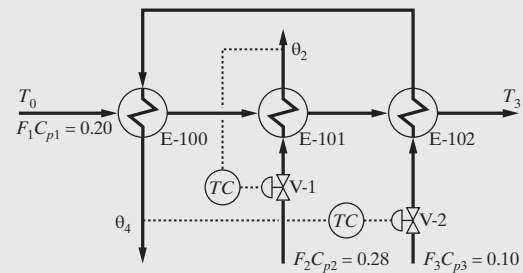

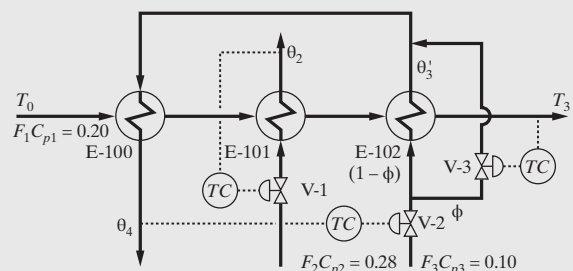**Figure 20.5** Control system for original heat exchanger network.

Figure 20.6 Control system for the modified heat exchanger network.

EXAMPLE 20.6 *Control Configuration for a Jacketed CSTR*

Consider the control of a jacketed continuous-stirred-tank reactor (CSTR) in which the exothermic reaction A → B is carried out. This system can be described by 10 variables as shown in Figure 20.7: $h, T, C_A, C_{Ai}, T_i, F_i, F_o, F_c, T_c$, and T_{co}, three of which are considered to be externally defined: C_{Ai}, T_i, and T_{co}. Its model involves four equations, assuming constant fluid density.

1. Overall mass balance:

$$A\frac{dh}{dt} = F_i - F_o \tag{20.8}$$

2. Mass balance on component A:

$$A\frac{d}{dt}(hC_A) = F_iC_{Ai} - F_oC_A - Ah \cdot r\{C_A, T\} \tag{20.9}$$

3. Energy balance on the reacting mixture:

$$A\rho c_p \frac{d}{dt}(h \cdot T) = F_i\rho c_pT_i - F_o\rho c_pT$$
$$- Ah \cdot r\{C_A, T\}(-\Delta H) - UA_s(T - T_c) \tag{20.10}$$

4. Energy balance on the jacket coolant:

$$V_c\rho_c c_{Pc}\frac{dT_c}{dt} = F_c\rho_c c_{pc}T_{co} - F_c\rho_c c_{pc}T_c + UA_s(T - T_c) \tag{20.11}$$

where A is the cross-sectional area of the vessel, h is the liquid level in the reactor, A_s is the area for heat transfer, U is the overall heat transfer coefficient, C_{Ai} and C_A are the inlet and reactor concentrations of A, T_i and T are the inlet and reactor temperatures, F_i and F_o are the inlet and outlet volumetric flow rates, ρ is the fluid density, F_c is the coolant volumentric flow rate, ρ_c is the coolant density, T_{co} and T_c are the inlet coolant and jacket temperatures, V_c is the volume of fluid in the cooling jacket, r is the intrinsic rate of reaction, ΔH is the heat of reaction, and c_p and c_{pc} are the specific heats of the reacting mixture and coolant, respectively.

Here, the number of variables that can be manipulated independently is $N_M = N_V - N_{ED} - N_E = 10 - 3 - 4 = 3$.

Selection of Controlled Variables. C_A should be selected because it affects the product quality directly (Guideline 3). T should be selected because it must be regulated properly to avoid safety problems (Guideline 2) and because it interacts with C_A (Guideline 4). Finally, h must be selected as a controlled output because it is non-self-regulating (Guideline 1).
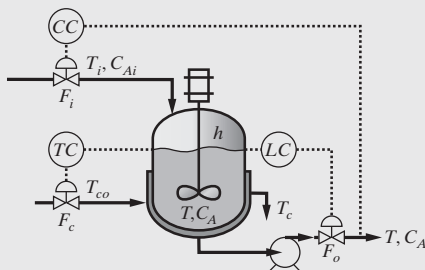
Selection of Manipulated Variables. The volumetric feed flow rate, F_i, should be selected because it directly and rapidly affects the conversion (Guidelines 6, 7, and 8). Using the same reasoning, F_c is selected to control the reactor temperature, T; and the flow rate of the reactor effluent, F_o, is selected to control h. This configuration, which is shown in Figure 20.7, should be compared with other pairings using the quantitative analysis presented in the supplement to this chapter, it being noted that there are several opportunities for improvement.

EXAMPLE 20.7 *Control Configuration for a Utilities Subsystem*

Often, the contents of a chemical batch reactor are heated initially to achieve ignition and then cooled to remove the heat generated in reaction. In such cases, it is common to install a jacket supplied with both cooling and heating utility streams as shown in Figure 20.8. The utilities subsystem involves eight variables: $P_{cf}, T_{cf}, F_{c1}, T_{c1}, F_{c2}, T_{c2}, F_c$, and T_{co}. Of these, two are externally defined and constitute disturbance variables: P_{cf} and T_{cf}. Four material and energy balances relate the subsystem variables: (1) an energy balance for the cooling branch, (2) an energy balance for the heating branch, (3) an energy balance for the mixing junction, and (4) a mass balance for the mixing junction. Hence, the number of variables to be manipulated independently is $N_M = N_V - N_{ED} - N_E = 8 - 2 - 4 = 2$. This is also the number of subsystem variables that can be controlled independently.

Selection of Controlled Variables. The guidelines presented earlier are not helpful because no output variable has a direct effect on the product quality, all are self-regulating, and none is directly associated with equipment or operating constraints. Nonetheless, F_c and T_{co} are obvious choices for the controlled variables because the objective of this subsystem is to control the temperature and flow rate of the utility stream fed to the reactor jacket.

Selection of Manipulated Variables. The two obvious candidates are F_{c1} and F_{c2} because both affect the two outputs directly and rapidly (Guidelines 7 and 8). However, linear and nonlinear combinations of these flow rates are also possible. As shown in Example 20S.4, a quantitative analysis is needed to make the best selection.
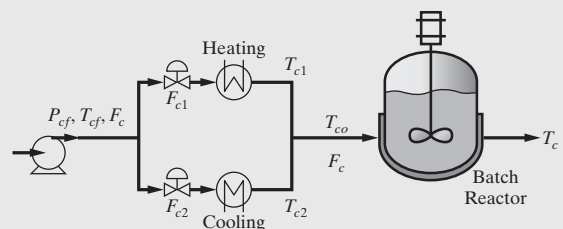

Figure 20.7 Control system for a jacketed CSTR.

Figure 20.8 Utilities subsystem for a batch chemical reactor.

EXAMPLE 20.8   *Control Configuration for a Flash Drum*

The flash drum in Figure 20.9 illustrates a situation in which a stream containing a binary mixture of two components, A and B, is flashed through a valve and separated in a flash drum into an overhead vapor stream and a residual liquid product stream. External heat exchange with liquid recycle is provided with a fixed recycle ratio. This process is modeled with 11 variables: F_i, T, C_A, F_W, P_f, h, T_f, F_V, y_A, F_L, and x_A. Two variables are considered to be externally defined, T and C_A. The model involves five equations: a total mass balance, a mass balance for component A, an overall energy balance, and a vapor–liquid equilibrium equation for each component. Thus, the number of variables to be manipulated independently is $N_M = N_V - N_{ED} - N_E = 11 - 2 - 5 = 4$.

Selection of Controlled Variables. P_f is selected because of the potential safety problems (Guideline 2) and because it affects the product concentrations (Guideline 3). T_f should be selected because it directly affects the product quality (Guideline 3). The liquid level in the drum, h, must be selected because it is not self-regulating (Guideline 1), and F_i is selected because it controls the product flow rate directly, one of the overall control objectives (Guideline 3). Note that all of these outputs exhibit significant interaction.

Selection of Manipulated Variables. F_i is designated as the *production handle*, that is, the manipulated variable selected to maintain the desired production rate. It is adjusted to achieve its setpoint (Guideline 8). F_V has a rapid, direct effect on the vessel pressure, P_f, but almost no effect on any other output (Guidelines 7 and 8). For similar reasons, F_L is selected to control the liquid level, h. F_W is selected because it directly controls the flash temperature, T_f (Guideline 8).
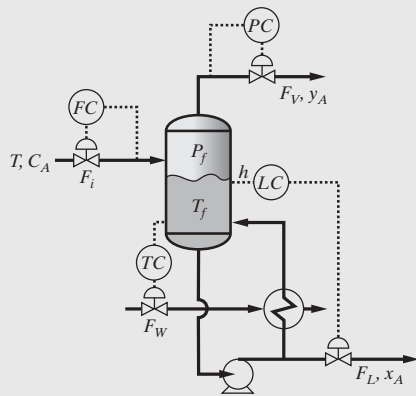

Figure 20.9 Control configuration for a flash drum.

EXAMPLE 20.9   *Control Configurations for a Binary Distillation Column*

This analysis of the distillation operation in Figure 20.10 is based on the following assumptions and specifications: (1) constant relative volatility, (2) saturated liquid distillate, (3) negligible vapor holdup
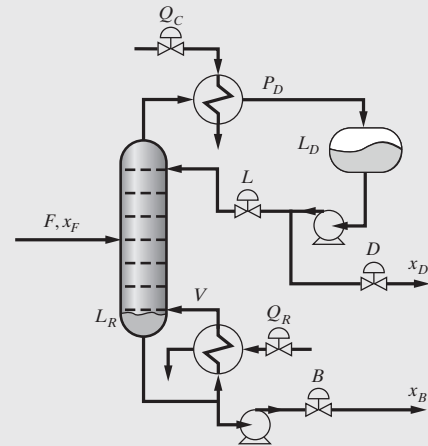

Figure 20.10 A distillation column with two liquid products.

in the column, (4) constant tray pressure drops (Luyben, 1990), and (5) negligible heat losses except for the condenser and reboiler. A column consisting of N_T trays is modeled in terms of the following $4N_T + 13$ variables:

Vapor and liquid compositions on each tray	$2N_T$
Tray liquid flow rates and holdups	$2N_T$
Reflux drum holdup and composition	2
Reflux and distillate flow rates	2
Column sump vapor and liquid compositions	2
Column sump liquid holdup	1
Bottoms product and reboiler steam flow rates	2
Feed flow rate and composition	2
Condenser pressure	1
Condenser duty	1

The system is described by $4N_T + 6$ equations:

Species mass balances (trays, sump, reflux drum)	$N_T + 2$
Total mass balances (trays, sump, reflux drum)	$N_T + 2$
Vapor–liquid equilibrium (trays, sump)	$N_T + 1$
Tray hydraulics for tray holdup	N_T
Total vapor dynamics	1

Assuming that the feed flow rate and composition are externally defined, the number of variables to be manipulated independently is $N_M = N_V - N_{ED} - N_E = 4N_T + 13 - 2 - (4N_T + 6) = 5$.

Selection of Controlled Variables The condenser pressure, P_D, should be regulated because it strongly affects the product compositions (Guidelines 3 and 4). The reflux drum and sump liquid inventory levels, L_D and L_R, need to be regulated because they are not self-regulating (Guideline 1). This leaves two additional variables that can be regulated. When distillate and bottoms streams are product streams, their compositions, x_D and x_B, respectively, are often selected as controlled variables (Guideline 3). Because significant delay times
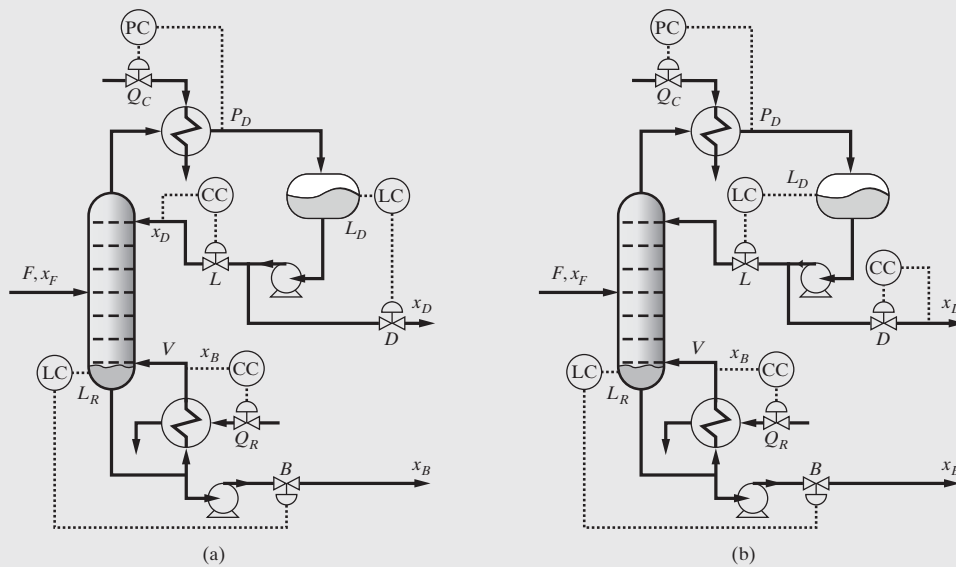
Figure 20.11 Two control configurations for a binary distillation column: (a) LV; (b) DV.

compositions (Guideline 12). In this regard, temperatures must be measured on trays that are sensitive to column upsets (Guideline 11).

Selection of Manipulated Variables. As shown in Figure 20.10 by labels on the valves, the five manipulated variables are the flow rates of the reflux, distillate, and bottoms streams L, D, and B, respectively, and the cooling and heating duties, Q_C and Q_R, respectively. It is most common to control P_D in columns with a liquid overhead product by manipulating Q_C. This leaves two compositions and two liquid inventories to control. A commonly employed configuration is to use the reflux, L, to control the top composition, x_D, and the reboiler duty, Q_R (which is closely related to V), to control the bottoms composition, x_B, in the so-called LV-configuration shown in Figure 20.11a. This leaves the distillate and bottoms streams, D and B, to control the reflux and sump liquid inventory levels, L_D and L_R, respectively. In columns operating with large reflux ratios (above 4), the flow of distillate will be insufficient to adequately regulate the reflux level, and in such cases, it is advisable to use L to control L_D (Guideline 6). Then, D regulates x_D, in the so-called DV-configuration, shown in Figure 20.11b, noting that the dynamic performance of this scheme is inferior to the LV-configuration. Alternative configurations involve ratios of manipulated variables intended to decouple the control loops by reducing the interaction between them (Shinskey, 1984; Luyben et al., 1999). The dynamic performance of a column control system should be verified using the quantitative methods described in the supplement to this chapter.

20.3 QUALITATIVE PLANTWIDE CONTROL SYSTEM SYNTHESIS

As pointed out by Luyben et al. (1999), the design of a plantwide control system should be driven by the objectives of the overall process rather than by considerations of the individual processing units as in the preceding section. Their strategy for control system design utilizes the available degrees of freedom to achieve these objectives in order of importance by adopting

a "top-down" approach in common with successful programming practice. Alternatively, in a simpler "bottom-up" approach (Stephanopoulos, 1984), the process is divided into subsystems with each subsystem often composed of several process units that share a common processing goal. Then, a control system is formulated for each subsystem by relying on the qualitative guidelines in Section 20.2 or the quantitative analysis to be described in the supplement to this chapter. Finally, an integrated system is synthesized by eliminating possible conflicts among the subsystems. The main disadvantage of this bottom-up approach is that good solutions at the subsystem level may not satisfy the process objectives. This can occur when manipulated variables are assigned to meet the control objectives of a subsystem, leaving less attractive inputs to satisfy those of the overall process. As will be demonstrated, interactions among subsystems, such as those resulting from heat integration and material recycle, are not addressed in this decomposition approach, which often leads to unworkable solutions.

The qualitative design procedure for plantwide control by Luyben et al. (1999) consists of the following steps:

Step 1: ***Establish the Control Objectives.*** As mentioned, the control objectives are related closely to the process objectives. For example, one may wish to impose a given production rate while ensuring that the products satisfy the quality specified by the market and guaranteeing that the process meets environmental and safety constraints.

Step 2: ***Determine the Control Degrees of Freedom.*** In practice, the degrees-of-freedom analysis in Section 20.2 may be too cumbersome for the synthesis of plantwide control systems. In a more direct approach, the number of control valves in the flowsheet equals the degrees of freedom (Luyben et al., 1999). As the valves are positioned on the flowsheet, care must

be taken to avoid the control of a flow rate by more than one valve. When the degrees of freedom are insufficient to meet all of the control objectives, it may be necessary to add control valves, for example, by adding bypass lines around heat exchangers as shown in Example 20.5 or by adding trim heaters or coolers serviced by utility streams.

Step 3: *Establish the Energy-management System.* In this step, control loops are positioned to regulate exothermic and endothermic reactors at desired temperatures. In addition, temperature controllers are positioned to ensure that disturbances are removed from the process through utility streams rather than recycled by heat-integrated process units.

Step 4: *Set the Production Rate.* This is accomplished by placing a flow control loop on the principal feed stream (referred to as *fixed feed* or *fresh feed*) or on the principal product stream (referred to as *on-demand product*), noting that these two options lead to different plantwide control configurations. Alternatively, the production rate is controlled by regulating the reactor operating conditions, for example, by controlling reactor hold-up.

Step 5: *Control the Product Quality and Handle Safety, Environmental, and Operational Constraints.* Having regulated the production rate and the effect of temperature disturbances, secondary objectives to regulate product quality and satisfy safety, environmental, operational, and process constraints are addressed in this step.

Step 6: *Fix a Flow Rate in Every Recycle Loop and Control Vapor and Liquid Inventories (vessel pressures and levels).* Process unit inventories, such as liquid holdups and vessel pressures (measures of vapor holdups), are relatively easy to control. Although vessel holdups are usually non-self-regulating (Guideline 1), the dynamic performance of their controllers is usually less important. In fact, level controllers are usually detuned to allow the vessel accumulations to dampen disturbances in the same way that shock absorbers cushion an automobile as demonstrated in the supplement to this chapter. Less obvious is the need to handle plantwide holdups in recycle loops. As will be shown qualitatively in several examples that follow and quantitatively in the supplement to this chapter, failure to impose flow control on each recycle stream can result in the loss of control of the process.

Step 7: *Check Component Balances.* In this step, control loops are installed to prevent the accumulation of individual chemical species in the process. Without control, chemical species often build up, especially in material recycle loops.

Step 8: *Control the Individual Process Units.* At this point, the remaining degrees of freedom are assigned to ensure that adequate local control is provided in each process unit. Because this step comes after the main plantwide issues have been handled, it often requires no additions to the control system.

Step 9: *Optimize Economics and Improve Dynamic Controllability.* When control valves remain to be assigned, they are utilized to improve the dynamic and economic performance of the process.

The above procedure is demonstrated next on three processes of increasing complexity: (1) an acyclic process, (2) the reactor-flash-recycle process in Example 20.4, and (3) the vinyl-chloride process discussed in Sections 2.3 and 2.5. The second and third examples feature recycle loops.

EXAMPLE 20.10   *Plantwide Control System Configurations for an Acyclic Process*

The chemical process shown in Figure 20.12 is based on an example by Stephanopoulos (1984). It consists of a CSTR in which species A reacts to form B in an exothermic reaction. The reactor effluent is fed to a flash vessel where the heavier product B is concentrated in the liquid stream, and unreacted A is discarded in the vapor stream. A preheater recovers heat from the hot reactor effluent with a so-called trim heater installed to ensure that the liquid reactor feed is at the desired temperature. To ensure that the reactor temperature remains on target, the CSTR is equipped with a jacket fed with cooling water to attenuate the heat released.
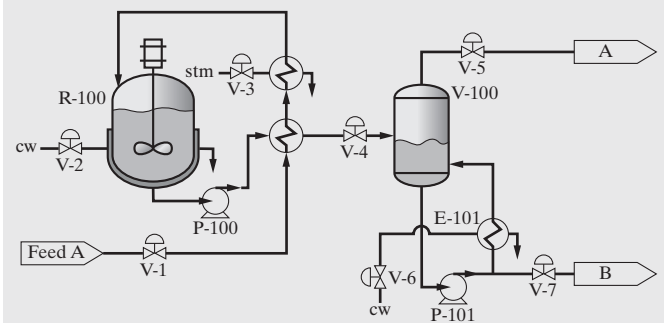

Figure 20.12 Process flowsheet for the acyclic process.

Applying the nine-step design procedure for plantwide control by Luyben and co-workers (1999), and using when possible the guidelines of Section 20.2:

Step 1: *Set Objectives.* The control objectives for this process are as follows:

1. Maintain the production rate of component B at a specified level.

2. Keep the conversion of the plant at its highest permissible value.

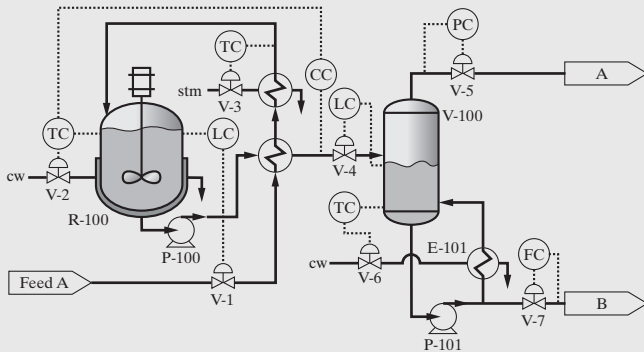3. Achieve constant composition in the liquid effluent from the flash drum.

Figure 20.13 Control structure for *on-demand product* in the acyclic process.

The structure of the plantwide control system depends on the primary control objective, that is, to maintain a desired production rate. The two possible interpretations of this goal are to ensure (1) a desired flow rate of the product stream by flow control using valve V-7, which leads to the *on-demand product* configuration shown in Figure 20.13 or (2) a desired production level by *fixed feed* using valve V-1, which leads to the control configuration shown in Figure 20.14. The *on-demand product* configuration is considered first.

Step 2: *Define Control Degrees of Freedom.* As shown in Figure 20.12, the process has seven degrees of freedom for manipulated variables. Having decided to design a configuration for *on-demand product*, the valve controlling the B product flow rate (V-7) is reserved for independent flow control (i.e., it directly controls the flow rate).

Step 3: *Establish the Energy-management System.* The critical energy management for the CSTR is handled next because loss of control of the reactor would have serious plantwide consequences. Using the guidelines for controlled and manipulated variable selection, the reactor feed and effluent temperatures are identified as critical for safety (Guideline 2) and quality assurance (Guideline 3). The obvious choices for valves to control these two temperatures are V-3, the steam valve for the trim heater, and V-2, the jacket coolant valve, both of which have a direct effect (Guidelines 6 and 7). These are assigned to temperature control loops.

Step 4: *Set the Production Rate.* As mentioned, the B product valve, V-7, is assigned to a flow controller whose setpoint directly regulates the production rate.

Step 5: *Control Product Quality, and Meet Safety, Environmental, and Operational Constraints.* The product quality is controlled by maintaining the operating temperature and pressure in the flash vessel at setpoints (Guideline 3). The former is regulated by adjusting the coolant water flow rate through V-6, and the latter is controlled by adjusting the vapor flow rate through overhead valve V-5. These valves are selected because of their rapid and direct effect on the outputs (Guidelines 6, 7, and 8). In addition, these two control loops satisfy the third control objective, that is, to provide tight product-quality control.

Step 6: *Fix Recycle Flow Rates and Vapor and Liquid Inventories.* The liquid inventories in the flash vessel and reactor are non-self-regulating and, therefore, need to be controlled

(Guideline 1). Because the liquid product valve from the flash vessel has been assigned to control the product flow rate, the inventory control must be in the reverse direction to the process flow. Thus, the reactor effluent valve, V-4, controls the flash vessel liquid level, and the feed valve, V-1, controls the reactor liquid level. Both of these valves have rapid, direct effects on the liquid holdups (Guidelines 6, 7, and 8). The vapor product valve, V-5, which has been assigned to control the pressure in V-100, thereby controls the vapor inventory.

Step 7: *Check Component Balances.* With the controllers assigned above, A and B cannot build up in the process, and consequently, this step is not needed.

Step 8: *Control the Individual Process Units.* Because all of the control valves have been assigned, no additional control loops can be designed for the process units, nor are they needed because both are already adequately controlled.

Step 9: *Optimize Economics and Improve Dynamic Controllability.* While a temperature control system for the CSTR is in place, its setpoint needs to be established. To meet the second control objective, which seeks to maximize conversion, a cascade controller is installed in which the setpoint of the reactor temperature controller (TC on V-2) is adjusted to control the concentration of B (CC) in the reactor effluent. If the reaction is irreversible, conversion is maximized by operating the reactor at the highest possible temperature, making this controller unnecessary. This completes the control system design for the on-demand product configuration in Figure 20.13. The performance of the control system needs to be verified by using controllability and resiliency assessment and by applying dynamic simulation as described in the supplement to this chapter.
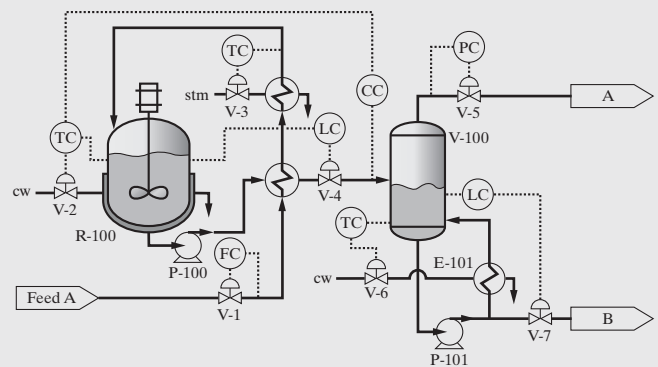

Figure 20.14 Control structure for *fixed feed* in the acyclic process.

As an alternative to the *on-demand product* configuration, the production level can be maintained by fixing the feed flow rate, which leads to the control system shown in Figure 20.14. This control configuration is derived using the same procedure as with Figure 20.13 with the only difference being in Step 6 where the liquid levels are controlled in the direction of the process flow because here, valve V-1 is in use as the production handle, and valve V-7 is free for use in inventory control. For the fixed feed configuration, reaction kinetics may dictate that the reactor holdup be manipulated in concert with throughput changes. In this case, it may be necessary to coordinate the reactor level setpoint with the feed flow rate.

EXAMPLE 20.11 *Plantwide Control System Configuration for Reactor-flash-recycle Process (Example 20.4 revisited)*

For the reactor-flash-recycle process introduced in Example 20.4, Figure 20.15 shows a control system with the control objectives to:

1. Maintain the production rate of component B at a specified level.
2. Keep the conversion of the plant at its highest permissible value.

The control configuration consists of six control loops: (1) production rate controlled using valve V-1 on the fresh feed stream, (2) temperature control using valve V-2 to ensure isothermal operation of R-100, (3) level control in R-100 holdup using valve V-3, (4) level control in V-100 using valve V-6, (5) pressure control in V-100 using the vapor-product valve, V-4, and (6) temperature control in V-100 (controlling product quality) using the coolant valve, V-5. This control system results from using a unit-by-unit design approach with each vessel inventory controlled by manipulation of its liquid effluent flow. Although the control pairings are acceptable for each process unit in isolation, the overall control system does not establish flow control of the recycle stream. Consequently, a change in the desired feed rate that keeps the reactor inventory constant with level control causes an excessive increase in the reactor effluent flow, which is transferred rapidly to the recycle flow by the flash level controller. This undesirable positive feedback is referred to as the "snowball effect" by Luyben and co-workers (1999) and is the consequence of not ensuring flow control of the recycle stream.

Since Luyben identified the snowball effect (Luyben, 1994), the sensitivity of reactor-separator-recycle systems to external disturbances has been the subject of several studies (e.g., Wu and Yu, 1996; Skogestad, 2002). Recent work by Bildea and co-workers (2000; Kiss et al., 2002) has shown that a critical reaction rate can be defined for each reactor-separator-recycle system using the Damköhler number, *Da* (dimensionless rate of reaction, proportional to the reaction rate constant and the reactor holdup). When the Damköhler number is below a critical value, Bildea et al. show that the conventional unit-by-unit approach in Figure 20.15 leads to the loss of control. Furthermore, they show that controllability problems associated with exothermic CSTRs and PFRs are often resolved by controlling the total flow rate of the reactor feed stream.

The extent of the snowball effect is shown next by analysis of the controlled process in Figure 20.15. The combined feed of pure A and recycle is partially converted to B in reactor R-100 by the isothermal, liquid-phase, irreversible reaction A → B, which has first-order kinetics. The reactor effluent is flashed across valve V-3 to give a
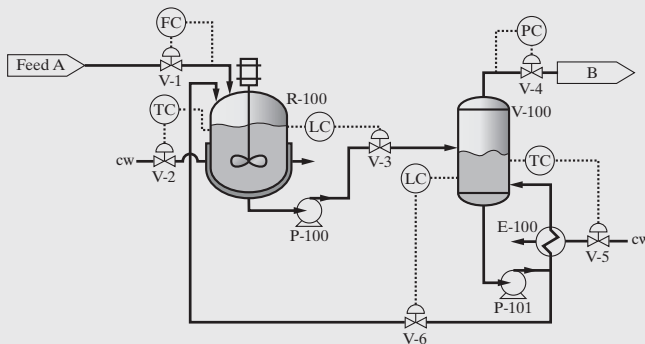

Figure 20.15 Control structure for reactor-flash-recycle process based on unit-by-unit design approach.

vapor-product stream assumed to be pure B and a liquid-product stream assumed to be pure A. The liquid stream is recycled to the reactor where it is mixed with fresh feed A to give the combined feed. What happens when the fresh feed flow rate changes? The equations that apply are:

Combined molar feed to the CSTR	$F_0 + B$
Molar material balance around the flash vessel	$F_0 + B = D + B$
Overall molar material balance	$F_0 = D$

where F_0, D, and B are the molar flow rates of the feed, flash vapor, and flash liquid streams. Finally, the rate of consumption of A in the reactor is:

$$r_A = kc_A \tag{20.12}$$

where r_A is the intrinsic rate of reaction, k is the first-order rate constant, and c_A is the molar concentration of A in the reactor effluent. Defining c_{total} as the total molar concentration and x_A as the mole fraction of A in the reactor effluent, Eq. (20.12) becomes:

$$r_A = kx_A c_{total} \tag{20.13}$$

The molar flow rate of B in the reactor effluent is:

$$(1 - x_A)(F_0 + B) = kx_A c_{total} V_R \tag{20.14}$$

where V_R is the volume of the reactor holdup. Then, substituting $c_{total} V_R = n_T$:

$$(1 - x_A)(F_0 + B) = kx_A n_T \tag{20.15}$$

where n_T is the total molar holdup in the reactor. Rearranging Eq. (20.15) for the flash liquid stream flow rate (recycled to the reactor), B:

$$B = \frac{x_A(F_0 + kn_T) - F_0}{1 - x_A} \tag{20.16}$$

With the reactor temperature and holdup fixed, a change to the fresh feed flow rate by a disturbance causes the mole fraction of A in the reactor effluent to change. Therefore, to obtain the effect of a change in F_0 on B, x_A must be eliminated from Eq. (20.16). For a perfect separation, an overall balance on the disappearance of A gives:

$$F_0 = kx_A n_T \tag{20.17}$$

Rearranging Eq. (20.17) for x_A and substituting in Eq. (20.16) gives:

$$B = \frac{F_0^2}{kn_T - F_0} \tag{20.18}$$

Equation (20.18) indicates that B increases by more than quadratically with increases in F_0. As an example, consider F_0 in the range of 50 to 150, with $kn_T = 200$. Then, Eq. (20.18) gives the recycle rate, B, as a function of F_0:

F_0	B
50	16.7
75	45.0
100	100
125	208
150	450

Thus, when the feed rate is tripled from 50 to 150, the recycle rate increases by a factor of $450/16.7 = 27$. This result assumes a fixed

kn_T. A more general result relies on reformulating Eq. (20.18) in terms of the Damköhler number, $Da = kn_T/F_0$, giving:

$$B = \frac{F_0}{Da - 1} \qquad (20.19)$$

Equation (20.19) shows that for values of Da much larger than unity, no snowball effect is expected. The snowball effect occurs as Da approaches a critical value of 1 and is eliminated by controlling the recycle flow rate as shown next.

To generate a workable plantwide control system as shown in Figure 20.16, the design procedure for plantwide control by Luyben and co-workers (1999) is applied:

Step 1: *Set Objectives.* To achieve the primary control objective, the production level is maintained by flow control of the feed stream using valve V-1.

Step 2: *Define Control Degrees of Freedom.* The process has six degrees of freedom.

Step 3: *Establish the Energy-management System.* The reactor temperature, which affects the process yield and stability (Guidelines 2 and 3), is controlled by adjusting the coolant flow rate, using valve V-2.

Step 4: *Set the Production Rate.* As stated previously, the feed valve, V-1, is assigned to a flow controller, whose setpoint regulates the production rate.

Step 5: *Control Product Quality, and Meet Safety, Environmental, and Operational Constraints.* A conventional pressure and temperature control system is set up for the flash vessel as in the previous example.

Step 6: *Fix Recycle Flow Rates and Vapor and Liquid Inventories.* To eliminate the snowball effect, the recycle flow rate must be controlled by installing a flow controller either on the reactor effluent or on the flash liquid effluent. As shown in Figure 20.16, the second option forces the reactor effluent valve, V-3, to control the flash vessel liquid inventory in the absence of other alternatives. Then, to regulate the reactor inventory, a cascade control system is designed in which the reactor level controller (LC) adjusts the setpoint of the feed flow controller (FC on V-1). This does not conflict with the objective to set the production rate by fixing the feed flow rate because in stable operation, the reactor level and feed flow rate vary proportionally through higher conversion in the CSTR. The vapor product valve, V-4, which has been assigned to control the pressure, thereby controls the vapor inventory.
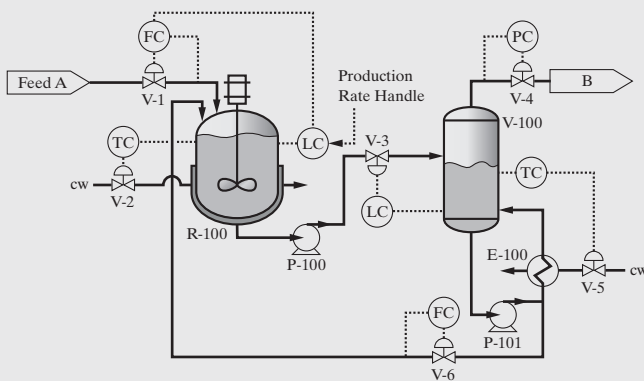

Figure 20.16 Workable plantwide control structure for reactor-flash-recycle process.

Steps 7 and 8: *Check Component Balances and Control Individual Process Units.* As in Example 20.10, the controllers assigned thus far prevent the buildup of A and B in the process. Both of the individual units are adequately controlled already, and in any case, no valves are left unassigned.

Step 9: *Optimize Economics and Improve Dynamic Controllability.* To maximize conversion, a cascade controller is installed as in the previous example in which the setpoint of the reactor temperature controller (TC on V-2) is adjusted to control the concentration of B in the reactor effluent. Again, for an irreversible reaction, it is enough to operate the reactor at the highest possible temperature.

EXAMPLE 20.12   *Plantwide Control System Configuration for the Vinyl-chloride Process*

For the vinyl-chloride process synthesized in Section 2.3 and shown in Figure 20.17, a preliminary design of its plantwide control system helps to assess the ease of maintaining the desired production level. This is achieved following the design procedure of Luyben and co-workers (1999):

Step 1: *Set Objectives.* Note that nearly 100% conversion is achieved in the dichloroethane reactor (R-100). Assuming that the conversion in the pyrolysis furnace (F-100) cannot be altered, the production level can be maintained by flow control of the ethylene feed flow rate using valve V-1.

Step 2: *Determine the Control Degrees of Freedom.* Twenty control valves have been positioned in the PFD, as shown in Figure 20.17.

Step 3: *Establish the Energy-management System.* The coolant valve, V-3, in the overhead condenser of the exothermic dichloroethane reactor, R-100, is used for temperature control. The yield in the pyrolysis furnace, F-100, is controlled by maintaining the outlet temperature at 500°C using the fuel gas valve, V-6. To attenuate the effect of temperature disturbances, the flow rates of the utility streams are adjusted to regulate effluent temperatures in the evaporator, E-101 (using V-5); the quench tank, V-100 (using the cooler E-102 and manipulating V-7); the partial condenser, E-103 (using V-8); and the recycle cooler, E-108 (using V-20). All of these valves act rapidly and directly on the controlled outputs (Guidelines 6, 7, and 8). Note that the temperature control loops using utility exchangers ensure that temperature disturbances are not recycled (Guideline 9).

Step 4: *Set the Production Rate.* As stated previously, the feed valve, V-1, is assigned to a flow controller, whose setpoint regulates the production rate.

Step 5: *Control Product Quality and Meet Safety, Environmental, and Operational Constraints.* The overhead product compositions in both distillation columns are regulated by adjusting the reflux flow rates using valves V-11 and V-16, both of which provide fast, direct control action (Guidelines 6, 7, and 8). The bottoms product compositions are controlled using the reboiler steam valves, V-13 and V-18. Thus, the control systems for both columns are in the LV-configuration. Valve V-9 is used to regulate the pressure in the feed to T-100. Because V-14 is needed for sump level control in T-100, it is not available for regulation of the feed pressure of T-101, and the addition of a pressure regulator should be considered for this purpose.
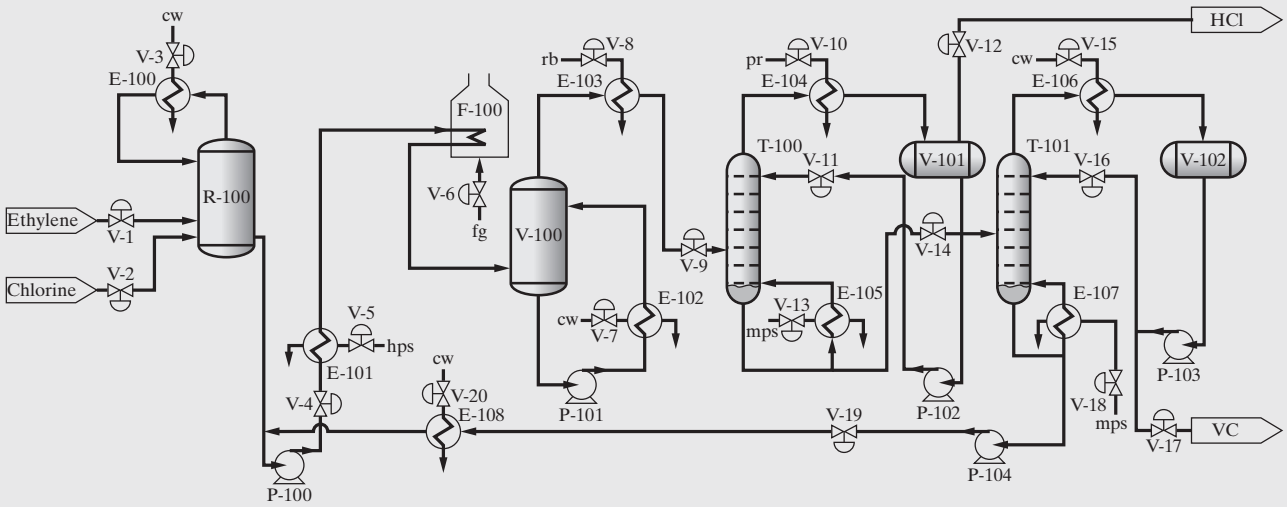
Figure 20.17 Control valve placement for the vinyl-chloride process.

Step 6: *Fix Recycle Flow Rates and Vapor and Liquid Inventories.* Inventory control of T-101 is assigned first. The bottoms flow rate in T-101, adjusted by valve V-19, is used for sump level control. The liquid level in reflux drum V-102 is controlled by manipulating the distillate valve, V-17. Inventory control for T-101 is completed by controlling the overhead pressure using the coolant valve, V-15. Turning to the HCl column, T-100, the bottoms product valve, V-14, is assigned to control the sump level. Because the overhead product is vapor, the condenser pressure is regulated using the distillate valve, V-12. Inventory control of T-100 is completed by assigning the condenser coolant valve, V-10, to regulate the reflux drum liquid level. The recycle flow rate must be held constant by flow control, and since V-19 is not available, a flow controller is installed to fix the combined recycle and feed flow rates using V-4. The level controller for R-100 is cascaded with the ethylene flow controller, making the level setpoint the production handle as in Example 20.11 (see Figure 20.16).

Steps 7 and 8: *Check Component Balances and Control Individual Unit Operations.* At this point, all but one of the valves (V-2) has been assigned. To ensure a stoichiometric ratio of reagents entering reactor R-100, the chlorine feed is adjusted to ensure complete conversion of the ethylene using a composition controller on the reactor effluent.

Step 9: *Optimize Economics and Improve Dynamic Controllability.* As in the previous example, to improve the range of production levels that can be tolerated, the setpoint of the recycle flow controller is set in proportion to the feed flow rate, suitably lagged for synchronization with the propagation rate through the process.

The complete control system is shown in Figure 20.18. Many of the qualitative decisions need to be checked by quantitative analysis or by simulation. For example, the interaction between the control systems of the two columns may require careful controller tuning. These refinements are discussed in the supplement to this chapter.
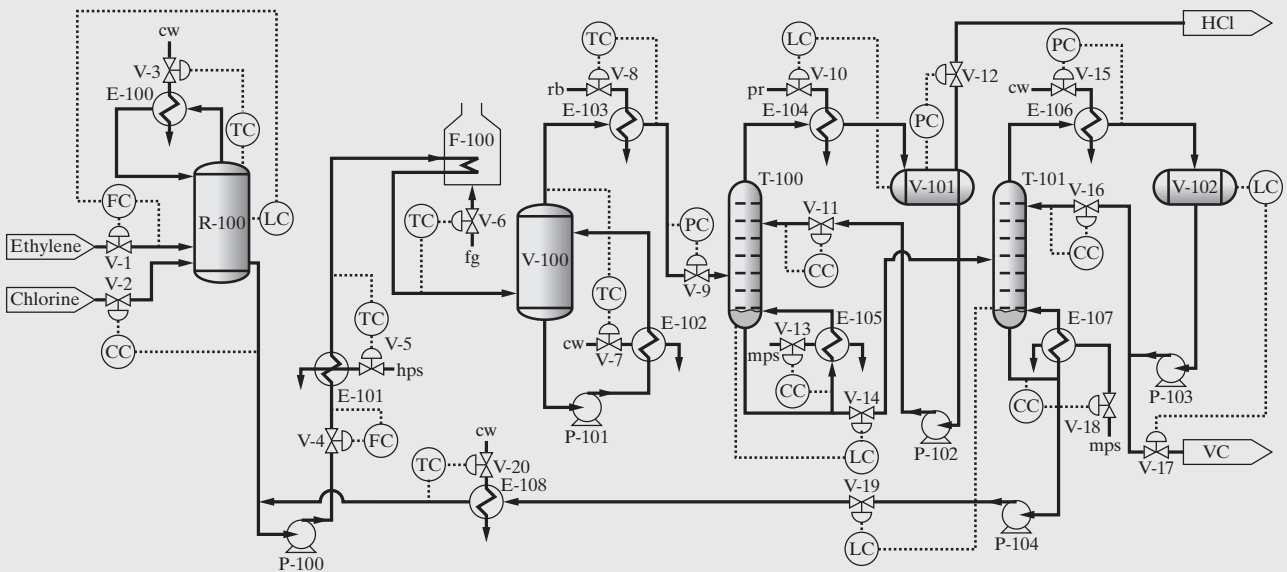

Figure 20.18 Control system for the vinyl-chloride process.

20.4 SUMMARY

This chapter has introduced the importance of considering plantwide control early in the design process. A qualitative control synthesis method combining the approaches suggested by Newell and Lee (1988) and Luyben and co-workers (1999) was presented to show how to generate alternative control configurations. The limitations of this qualitative approach have been highlighted, and the need for the quantitative approach presented in the supplement to this chapter, which involves analysis and dynamic simulation, has been established.

After studying this chapter, the reader should be able to:

1. Identify potential control problems in a process flowsheet.

2. Classify and select controlled and manipulated variables for a plantwide control system.

3. Perform a conceptual synthesis of plantwide control structures (pairings) based on degrees-of-freedom analysis and qualitative guidelines.

SUPPLEMENT TO CHAPTER 20—FLOWSHEET CONTROLLABILITY ANALYSIS

A supplement to Chapter 20, Flowsheet Controllability Analysis, is provided in the PDF Files folder, which can be downloaded from the Wiley Web site associated with this textbook. (See the file Supplement_to_Chapter 20.pdf.) The contents of this supplement are:

20S.0 Objectives

20S.1 Generation of Linear Models in Standard Forms

20S.2 Quantitative Measures for Controllability and Resiliency
Relative-gain Array (RGA)
Properties of Steady-state RGA
Dynamic RGA (McAvoy, 1983)
The RGA as a Measure of Process Sensitivity to Uncertainty
Using the Disturbance Cost to Assess Resiliency to Disturbances

20S.3 Toward Automated Flowsheet C&R Diagnosis
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Generating Low-order Dynamic Models
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20S.4 Control Loop Definition and Tuning
Definition of PID Control Loop
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20S.5 Case Studies
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Case Study 20S.3 Interaction of Design and Control in the MCB Separation Process

20S.6 MATLAB for C&R Analysis

20S.7 Summary
References
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EXERCISES

20.1 Perform a degrees-of-freedom analysis for the noninteracting exothermic reactor shown in Figure 20.3a. Suggest an appropriate control structure. Carry out the same exercise for the heat-integrated reactor shown in Figure 20.3b. Compare the results.

20.2 Consider the mixing vessel shown in Figure 20.19. The feed stream flow rate, F_1, and composition, C_1, are considered to be disturbance variables. The feed is mixed with a control stream of flow rate F_2 and constant known composition, C_2. To ensure a product of constant composition, it is also possible to manipulate the flow rate, F_3, of the product stream. Perform a degrees-of-freedom analysis and suggest alternative control system configurations. Note that unsteady-state balances are required.
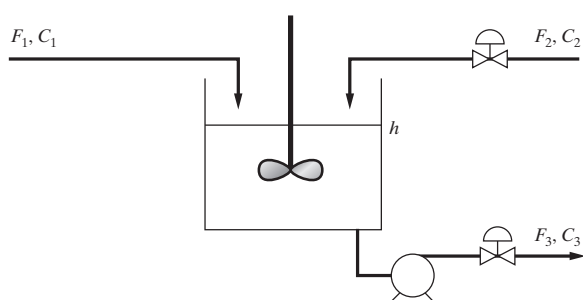
Figure 20.19 Mixing vessel.

20.3 Consider the FS two-column configuration for the separation of methanol and water in Figure 20.2 and (a) determine the number of degrees of freedom for the overall system, (b) determine the number of controlled and manipulated variables, and (c) select a workable control configuration using qualitative arguments.

20.4 Repeat Exercise 20.3 for the LSF configuration in Figure 20.2.

20.5 A control system is suggested for the exothermic reactor in Figure 20.7. Suggest alternative configurations and compare them with the original configuration.

20.6 Figure 12.20 shows a process for the isothermal production of C from A and B (A + B → C). The two reagents are fed to a CSTR, R-100, where complete conversion of B is assumed. The reactor effluent stream consisting of C and unreacted A is separated in a distillation column,
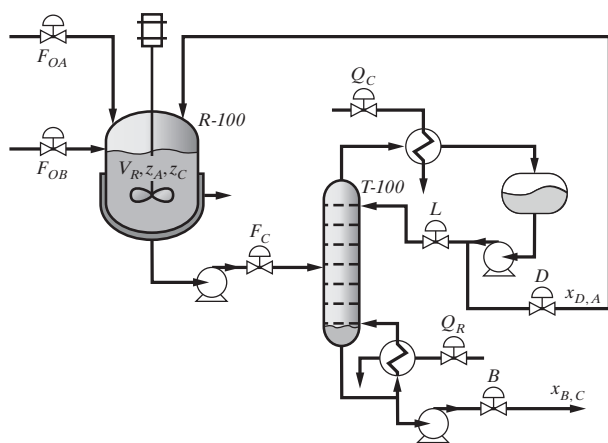

Figure 20.20 Process flowsheet for Exercise 20.6.

T-100, where the more volatile A is withdrawn in the distillate and recycled, and product C is withdrawn in the bottoms stream. Your task is to devise a conceptual plantwide control system for the process.

Hint: It may be helpful to reposition the feed stream of A.

20.7 Figure 20.21 shows the flowsheet for a reactive distillation column for the production of ethylene glycol (EG) from ethylene oxide (EO) and water (Al-Arfaj and Luyben, 2002):

$$EO + H_2O \rightarrow EG$$

Note that the reaction proceeds to 100% conversion in the column with part of the EG undergoing a secondary (undesired) reaction to diethylene glycol (DEG):
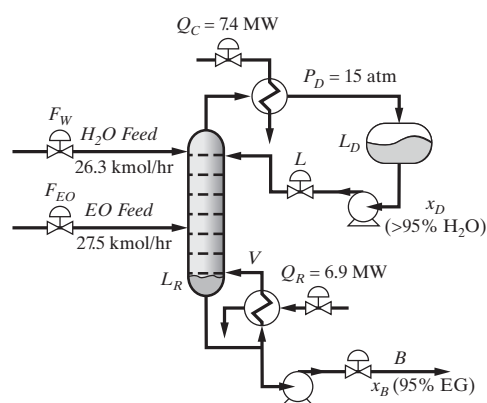
$$EO + EG \rightarrow DEG$$

Figure 20.21 Ethylene glycol reactive distillation column.

For this reason, the EO is fed to the column in slight excess. The EG product is withdrawn as the bottoms stream, and almost pure water concentrates at the top of the column. Your task is to use the procedure of Luyben and co-workers, showing all steps, to devise a conceptual plantwide control system for the process with the following objectives:

(a) Control production rate.

(b) Ensure the EG product is at the required concentration.

20.8 Figure 20.22 shows the monochlorobenzene separation process introduced in Section 7.4. The process involves a flash vessel, V-100; an absorption column, T-100; a distillation column, T-101; a reflux drum, V-101; and three utility heat exchangers. As shown in Figure 20.22, most of the HCl is removed at high purity in the vapor effluent of T-100. However, in contrast with the design shown in Chapter 7, the design in Figure 20.22 does not include a "treater" to remove the residual HCl; instead, it is purged in a small vapor overhead product stream in T-101. The benzene and monochlorobenzene are obtained at high purity as distillate and bottoms liquid products in T-101. Note that the 12 available control valves are identified. Your task is to design a conceptual control system to ensure that the process provides stable production at a desired level while meeting quality specifications.

20.9 Figure 20.23 shows a heat-integrated process for the manufacture of vinyl chloride. This design sharply reduces the utilization of utilities in Figure 20.17 without requiring additional heat exchangers. Design a conceptual control system for the same control objectives in Example 20.12.

Hint: To provide sufficient degrees of freedom, it may be necessary to add heat exchanger bypasses and/or trim utility exchangers.
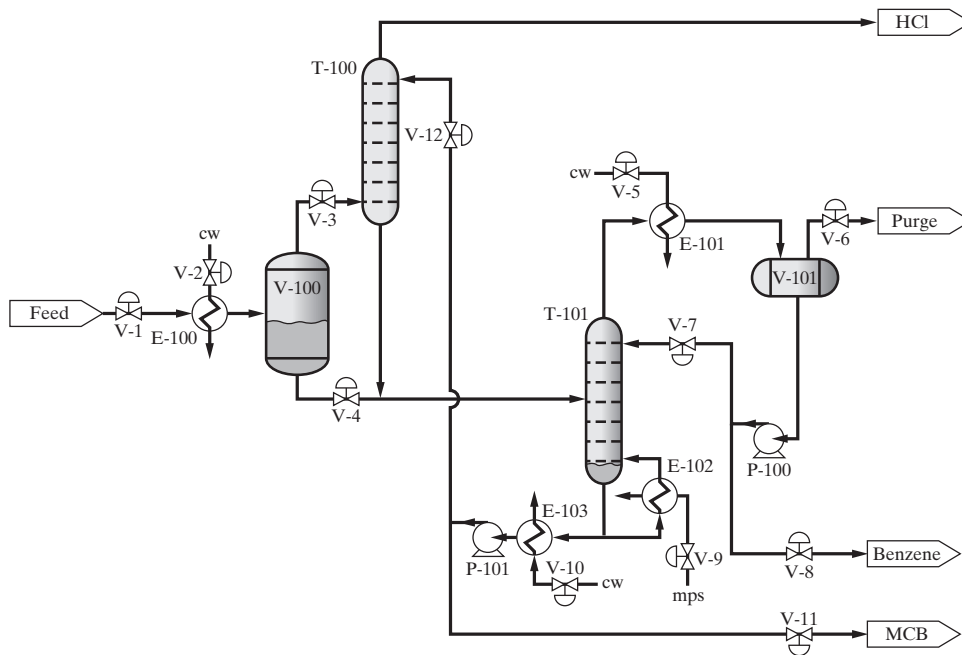
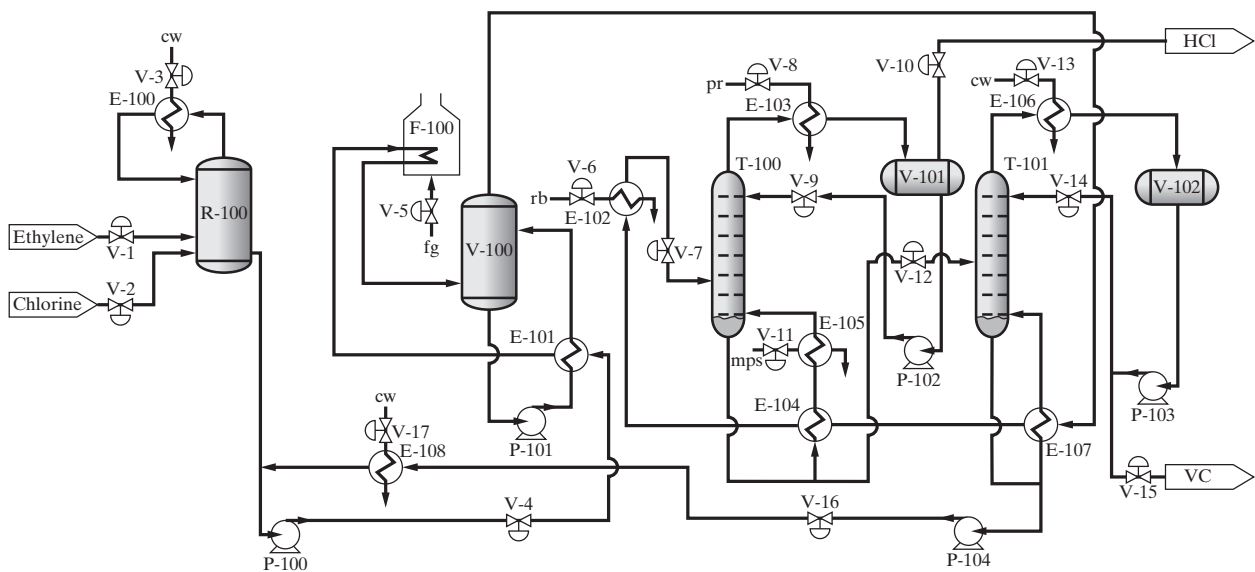**Figure 20.22** Process flowsheet for the MCB separation process.

Figure 20.23 Process flowsheet for the heat-integrated vinyl-chloride process.

20.10 How would the control configuration for the vinyl-chloride process in Figure 20.18 change if the primary control objective is to provide on-demand vinyl-chloride product?

20.11 Figure 20.24 shows the flowsheet for a process for the production of methanol from synthesis gas (a mixture of hydrogen, carbon monoxide, and carbon dioxide and a small quantity of residual methane, which is inert in this process). In the process, the feed stream undergoes compression from 13 to 48 bar in compressor K-100, is then mixed with the recycle, and then is further compressed to 50 bar in compressor K-101. Both of the compressors are driven by turbines fed with high-pressure steam (hps). The high-pressure combined feed is

heated to ignition temperature in heat exchanger E-100 and is then fed to the adiabatic reactor R-100, where it undergoes partial conversion to methanol and water as limited by the equilibrium conditions. The hot reactor effluent is used to preheat the feed in E-100, and is then further cooled in E-101 before being flashed in V-100. The vapor stream product from V-100 is recycled after a fraction is purged (to remove the inert methane fed in the process feed). The pressure of the liquid product stream from V-100 is reduced to 10 bar in pressure regulator V-7 and is then fed to distillation column T-100, where it is separated into a water/methanol mixture in the distillate and high-purity water in the bottoms. The distillate from T-100 undergoes further pressure reduction to 2 bar in pressure regulator V-13 and is then fed to a second
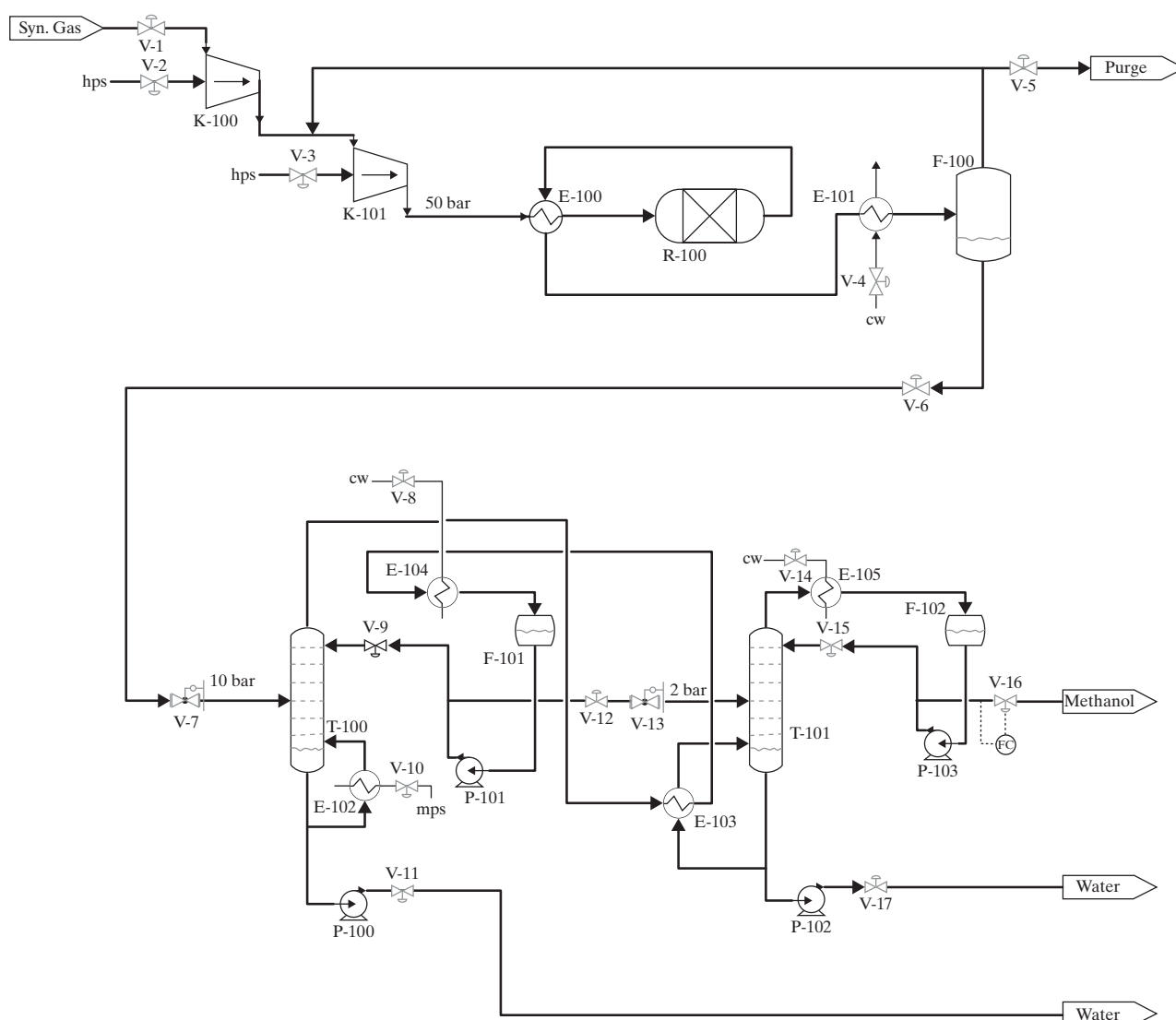
Figure 20.24 Process flowsheet for the production of methanol from synthesis gas.

column, T-101, which produces high-purity methanol as the distillate and high-purity water as bottoms. This arrangement enables the partial condensation of the overhead stream in T-100 to provide all of the heat of evaporation needed in the reboiler of T-101 in heat exchanger E-103, with the remaining condensation duty provided by the cooler E-104.

You are requested to suggest a plantwide control system to enable stable operation of the process in Figure 20.24, while satisfying the following requirements:

(a) Methanol production on demand; thus, valve V-16 is already assigned to accomplish flow control of the product stream as shown in Figure 20.24.

(b) Methanol product composition on specification.

(c) Maximum methanol recovery from the process.

(d) Maximum conversion to methanol in R-100.

Note: As indicated in Figure 20.24, the dispositions of three valves have already been resolved: V-16 is assigned to flow control for on-demand product, and both V-7 and V-13 are assigned for pressure regulation of column feed streams. Your solution should follow the

procedure of Luyben et al. and include the positioning of all control loops in the PFD of Figure 20.24. You are allowed to add control valves to those already in place in the PFD, but only if these are absolutely necessary to meet the requirements.

20.12 Figure 20.25 shows the flowsheet for a process for the production of synthesis gas (a mixture of hydrogen, carbon monoxide, and carbon dioxide and a small quantity of unreacted residual methane) from methane, steam, and oxygen. The feed stream of methane is mixed with steam that is generated using waste heat generated in the process, heated in E-100 and then fed to the reformer R-100, where partial conversion of the methane to hydrogen and CO takes place. The heat demanded by this endothermic reaction is provided by the combustion of fuel gas (fg) in R-100. The hot reformer effluent stream first feeds E-100 to preheat the reformer feed as stated and is then further cooled in E-101, mixed with the oxygen feed stream, and then fed to the adiabatic oxidation reactor, R-101, where the remaining methane reacts to produce more hydrogen and CO_2. The effluent from R-101 is cooled in E-102 and then fed to the adiabatic shift reactor, R-102, where it is possible to control the ratio of $CO : CO_2$ (depending on the feed temperature to R-102). The effluent stream from R-102 is cooled in E-103 and then fed to the flash vessel
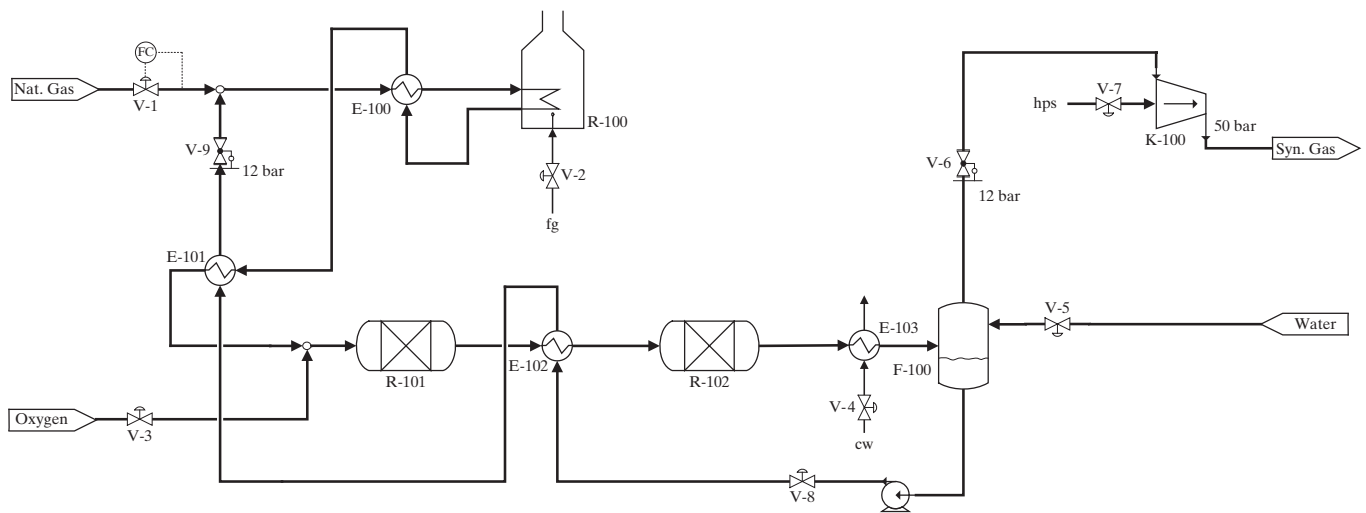
Figure 20.25 Process flowsheet for the production of synthesis gas from natural gas.

F-100, where the water in the synthesis gas condenses and is drawn off as the liquid effluent stream, which is recycled back to the process feed after being converted to saturated steam by heat exchange with hot process streams in E-102 and E-101. The vapor effluent from F-100 is compressed in K-100 to 50 bar and is the desired synthesis gas.

You are requested to suggest a plantwide control system to enable stable operation of the process in Figure 20.25 while satisfying the following requirements:

(a) *Fixed feed* methane consumption; thus, valve V-1 is already assigned to accomplish flow control of the methane feed stream as shown in Figure 20.25.

(b) Controlled hydrogen production rate in the reformer by regulation of the methane/steam ratio in the feed to R-100.

(c) Control of the $H_2 : CO_2$ ratio in the oxidation reactor (R-101) effluent.

(d) Control of the $CO : CO_2$ ratio in the shift reactor (R-102) effluent.

(e) Minimization of the energy consumption in the reformer.

Note: As indicated in Figure 20.25, the dispositions of three valves have already been resolved: V-1 is assigned to flow control for fixed-feed configuration, V-6 is assigned for pressure regulation of F-100 vapor flow, and V-9 is used to maintain the desired pressure of the saturated steam generated using the water recycle stream. Your solution should follow the procedure of Luyben et al. and should include the positioning of all control loops in the PFD of Figure 20.24. You are allowed to add control valves to those already in place in the PFD, but only if these are absolutely necessary to meet the requirements.

20.13 Figure 20.26 shows the flowsheet for a process for the production of G from A. The feed stream of A (entering through valve V-1) is mixed with a makeup stream of B (entering through valve V-2) and a recycle stream composed mostly of B, and the resulting stream is heated in E-100 and then fed to adiabatic PFR, R-100 in which occurs the exothermic reaction: $A + B \rightarrow C + D$. The R-100 effluent is separated in column T-100 into a bottoms stream rich in D, and a distillate rich in C. The distillate is mixed with a makeup stream of F (entering through valve V-8) and an F-rich recycle stream, and the resulting stream is

heated in E-103 and then fed to adiabatic PFR, R-101 in which occurs the exothermic reaction: $C + F \rightarrow E + B$. The R-101 effluent is separated in column T-101 into a bottoms stream rich in E and a distillate rich in B, which is recycled. The bottoms from T-101 (rich in E) is mixed with the bottoms from T-100 (rich in D) and a makeup stream of E (entering through valve V-14), and the resulting stream is heated in E-106 and then fed to adiabatic PFR, R-102 in which occurs the exothermic reaction: $D + E \rightarrow F + G$. The R-102 effluent is separated in column T-102 into a bottoms stream rich in G and a distillate rich in F, which is recycled. The relative volatilities of the participating components are in the order: $\alpha_A > \alpha_B > \alpha_C > \alpha_D > \alpha_E > \alpha_F > \alpha_G$.

You are requested to suggest a plantwide control system to enable stable operation of the process providing G on demand. Your solution should follow the procedure of Luyben et al. and should include the positioning of all control loops in the PFD of Figure 20.26. You are allowed to add control valves to those already in place in the PFD, but only if these are absolutely necessary to meet the requirements.

20.14 Figure 20.27 shows the flowsheet for a process for the production of gasoline (mainly octane) from an olefins feed (propane, propene, and butene). The feed to the process is heated in E-100, and then E-101, and then fed to a cascade of three PFRs operating at 500 psia where it is partially converted to gasoline (mainly octane), with temperature control affected using cold shots of propane. The hot reactor effluent stream is used to preheat the reactor feed in E-100 and is then fed to the first column, T-100, after pressure reduction to 200 psia by the pressure regulator, PRV-1, where the propane is removed as the distillate with the remaining components leaving as bottoms. Part of the propane is recycled at 500 psia for use as cold shot in the reactor cascade. The bottoms from T-100 are reduced to 90 psia by the pressure regulator, PRV-2, and then fed to the second column, T-101, which separates it into butane as distillate and gasoline as bottoms. All of the product streams are cooled using cooling water.

You are requested to suggest a plantwide control system to enable stable operation of the process providing gasoline on demand. Your solution should follow the procedure of Luyben et al. and should include the positioning of all control loops in the PFD of Figure 20.27. You are allowed to add control valves to those already in place in the PFD, but only if these are absolutely necessary to meet the requirements.
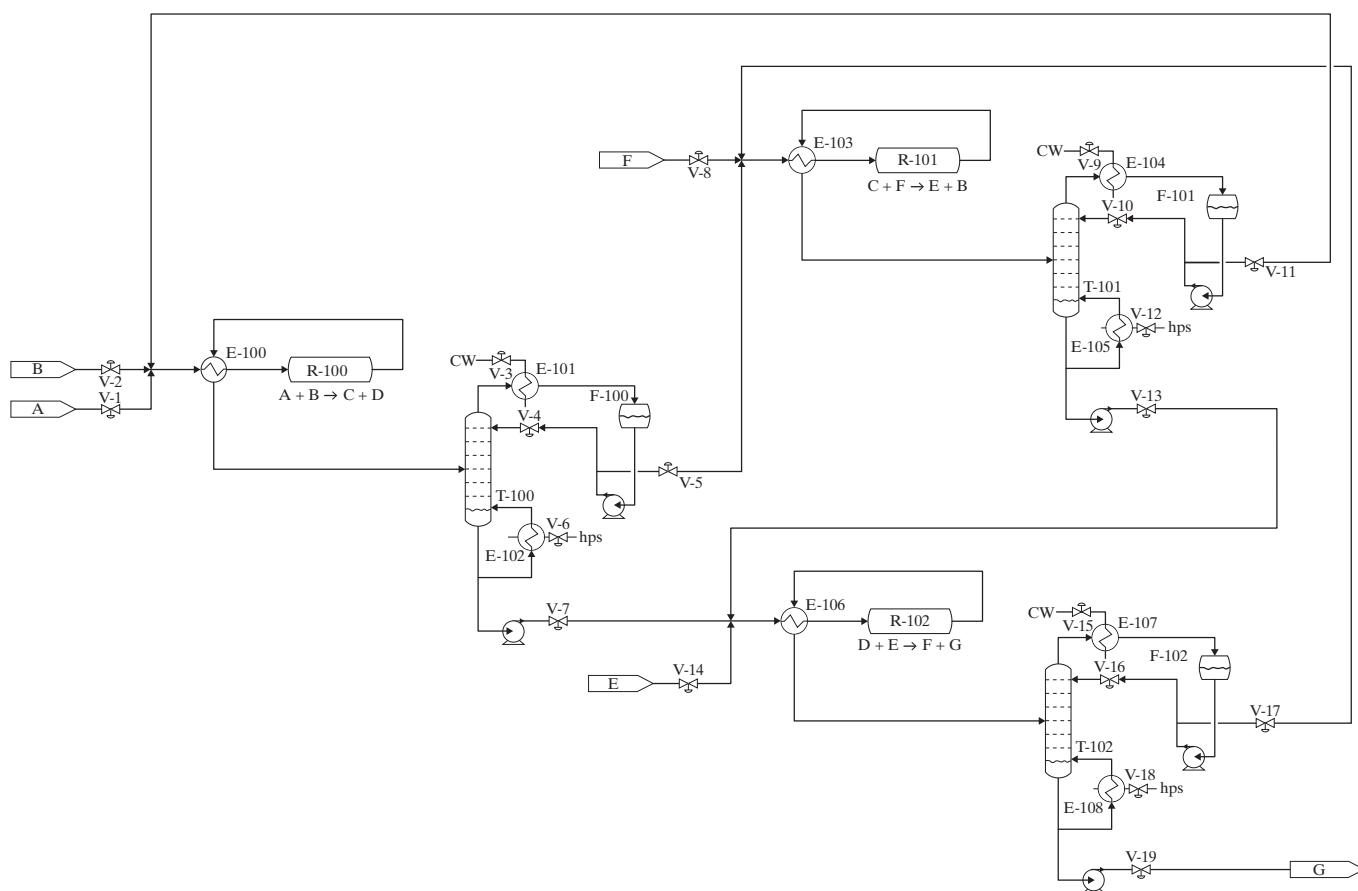
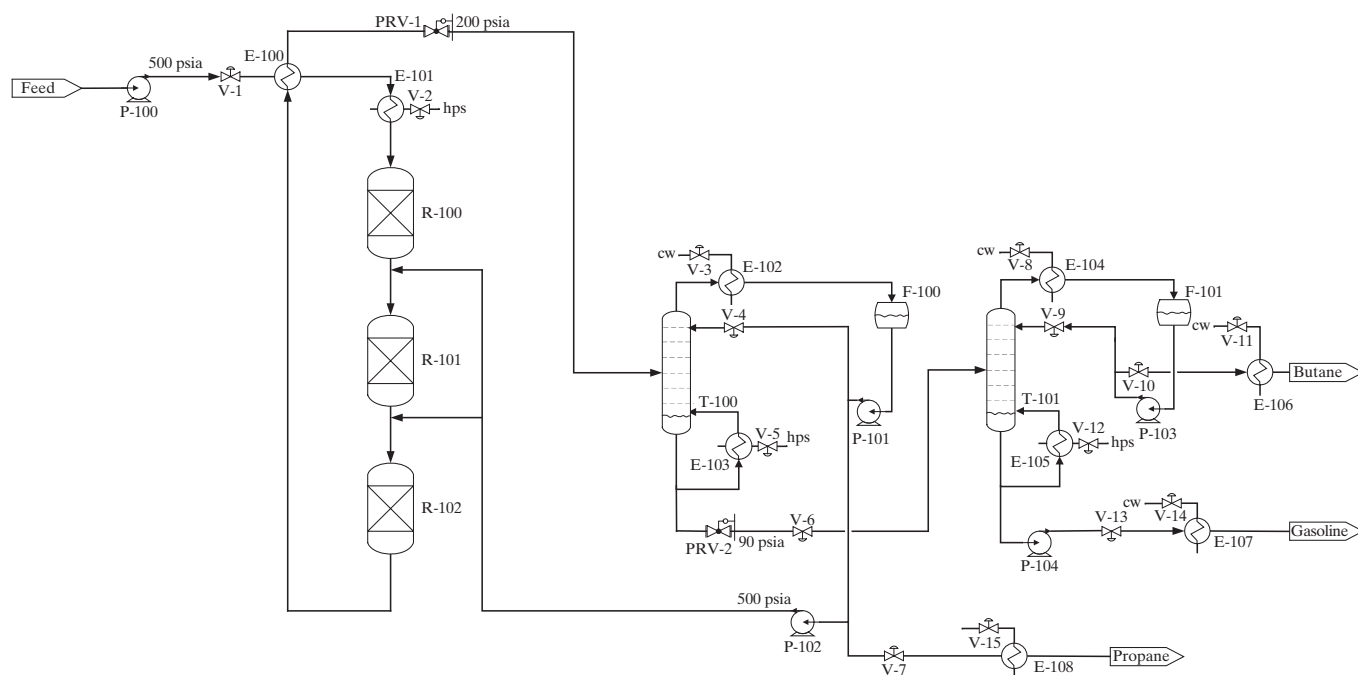**Figure 20.26** Process flowsheet for the production of G from A.

Figure 20.27 Process flowsheet for the production of gasoline from olefines.
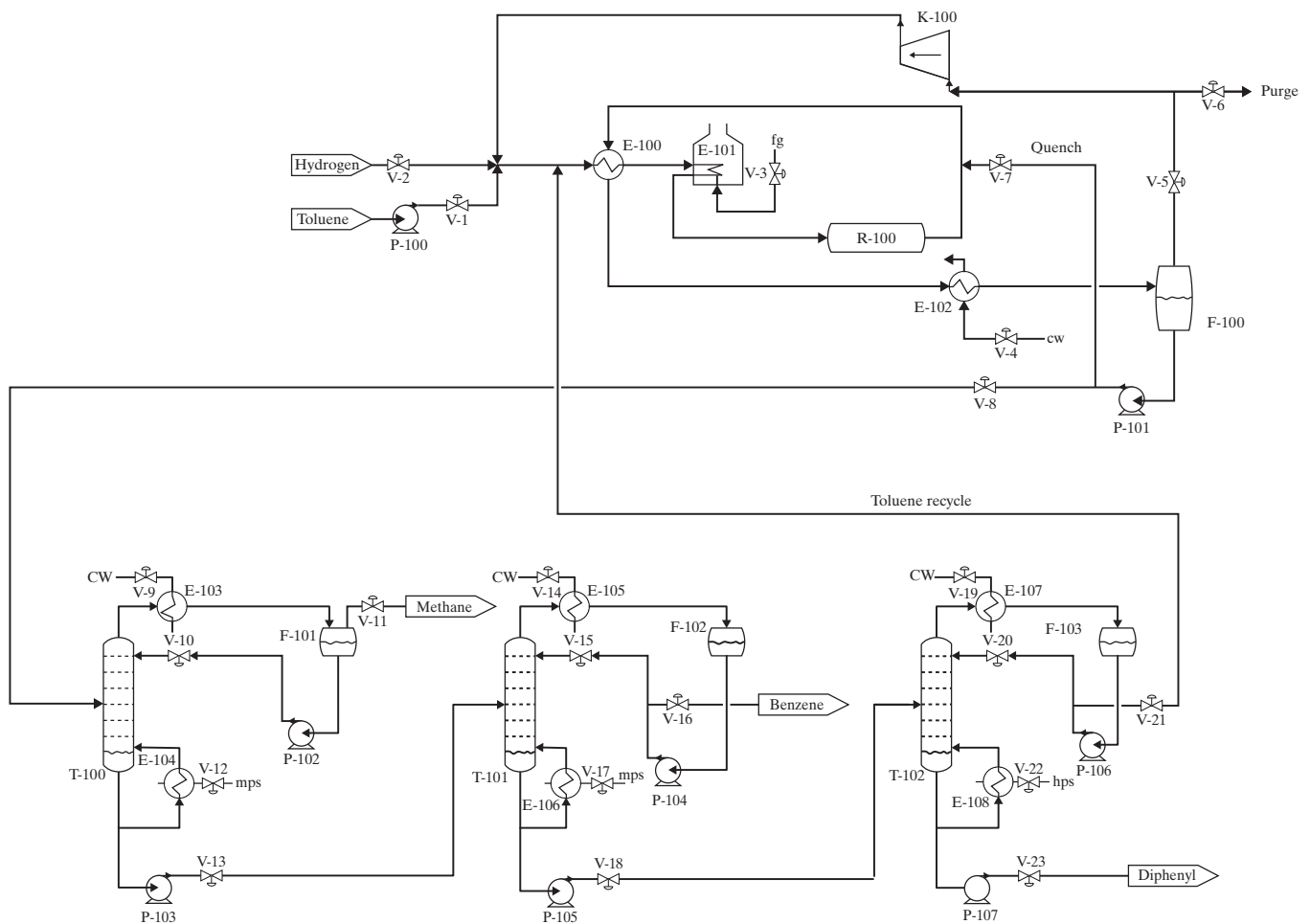
Figure 20.28 Process flowsheet for the HDA process.

20.15 Figure 20.28 shows the flowsheet for the HDA process for the production of benzene from toluene and hydrogen. Feed streams of toluene (entering through valve V-1) and hydrogen (entering through valve V-2) are mixed with a liquid recycle stream rich in toluene and a vapor recycle stream rich in hydrogen, and the combined stream is preheated first in E-100, and then vaporized and superheated in furnace E-101. The vaporized mixture of toluene and hydrogen is fed to the adiabatic PFR, R-100, where they mostly undergo partial conversion to the desired main products, benzene and methane, as well a small portion to the undesired byproduct, diphenyl. The hot reactor effluent is cooled by heat exchange with the reactor feed in E-100 and then by cooling water in E-102 before being flashed in F-100. A portion of the hydrogen-rich vapor product stream from F-100 is purged through valve V-6 to attenuate the methane content and then compressed in K-100 and recycled. The compressor is operated at maximum capacity to enable production rate to be maximized. A portion of the liquid product stream from F-100

is recycled and mixed with the reactor effluent stream to cool it before it enters E-100, but most of it is fed to a separation system consisting of three columns in series.

In the first column, T-100, the reactor products are separated into a methane-rich distillate with the other components exiting as a feed stream to the second column, T-101, which produces a benzene-rich distillate and a mixture of toluene and diphenyl as bottoms. This is fed to the last column, T-102, which separates it into a toluene-rich distillate, which is recycled, and a bottoms stream consisting largely of diphenyl.

You are requested to suggest a plantwide control system to enable stable operation of the process, while regulating the production rate. Your solution should follow the procedure of Luyben et al. and should include the positioning of all control loops in the PFD of Figure 20.28. You are asked not to add control valves to those already in place in the PFD or any additional utility heaters and coolers.