Appendix F

Introduction to Plantwide Control

Previous chapters have generally concentrated on the analysis and design of simple control systems, from single loops (such as reactor temperature control) to single processing units with multiple loops (e.g., distillation column control). However, most industrial applications involve larger problems with multiple processing units that interact with each other. The subject of *plantwide control* deals with unit-to-unit interactions through the choice of measured and manipulated variables in each unit and the selection of a control strategy—namely, how to pair controlled and manipulated variables in individual loops, where to use multiloop controllers (Chapter 16), where to use multivariable controllers such as MPC (Chapter 20), and so on.

For a new plant, the problem of designing the control system can be quite difficult as a consequence of unit-to-unit interactions. Thus, understanding the potential sources of these interactions and finding ways in which they can be substantially mitigated are important to achieve effective plant operations. In this chapter, we introduce several key concepts in plantwide control; Appendix G deals specifically with how to develop a control system design for a new plant.

Most continuous processing plants contain many units, such as reactors, furnaces, heat exchangers, and distillation columns. The goal of process design is to minimize capital costs while operating with optimum utilization of materials and energy. Unfortunately, achieving lower plant capital costs and higher processing efficiencies inevitably makes the individual units interact more with each other and thus makes them harder to control (see Chapter 16). The process control engineer deals with these unit-to-unit interactions by designing a control system that counteracts disturbances before they propagate from their source to other units.

A typical plantwide control system will consist of many single-loop controllers as well as multivariable controllers such as model predictive control (Chapter 20). A key characteristic of many plantwide control systems is the very large number of process variables, involving

1. Thousands of measurements
2. Hundreds to thousands of manipulated variables
3. Hundreds of disturbance variables

Fortunately, a plant with a large number of processing units can be analyzed as smaller clusters of units—for example, a gas treatment plant and a separations plant that interact very little with each other. Then, with even simple steady-state and dynamic process models, it is possible to develop a design using the standard analytical methods we developed in Chapter 16 (RGA, SVA, etc.) for multivariable control problems. In the absence of process models, one must resort to heuristic (rule-of-thumb) approaches. Although these approaches generally are based on prior experience, they also incorporate an understanding of the fundamental physics and chemistry that apply to all plants. In this chapter, several case studies are used to introduce important plantwide concepts. In the final chapter (Appendix G), we present a general strategy for designing plantwide control systems.

F.1 PLANTWIDE CONTROL ISSUES

One of the most basic issues in plantwide control is flow/inventory control. If a train of continuous processing units (reactors, columns, etc.) is considered, where should the production rate be controlled? It can be controlled at the exit of the line (e.g., a series of unit operations (as in Fig. F.1*a*)), at the beginning of the line (Fig. F.1*b*), or at any point in between. In these figures, the sensors/transmitters have been omitted for clarity. It might seem logical to use a feed flow rate into each unit to control the inventory (level) in that unit as illustrated in the *downstream method* of Fig. F.1*a*. However, as discussed below, adjusting each unit's effluent flow rate may be an easier way to control inventories if the flow rates of multiple streams into a unit are ratioed (see the *upstream method* in Fig. F.1*b*).

The objectives for any of these methods are (1) to maintain the production rate of the line (or the

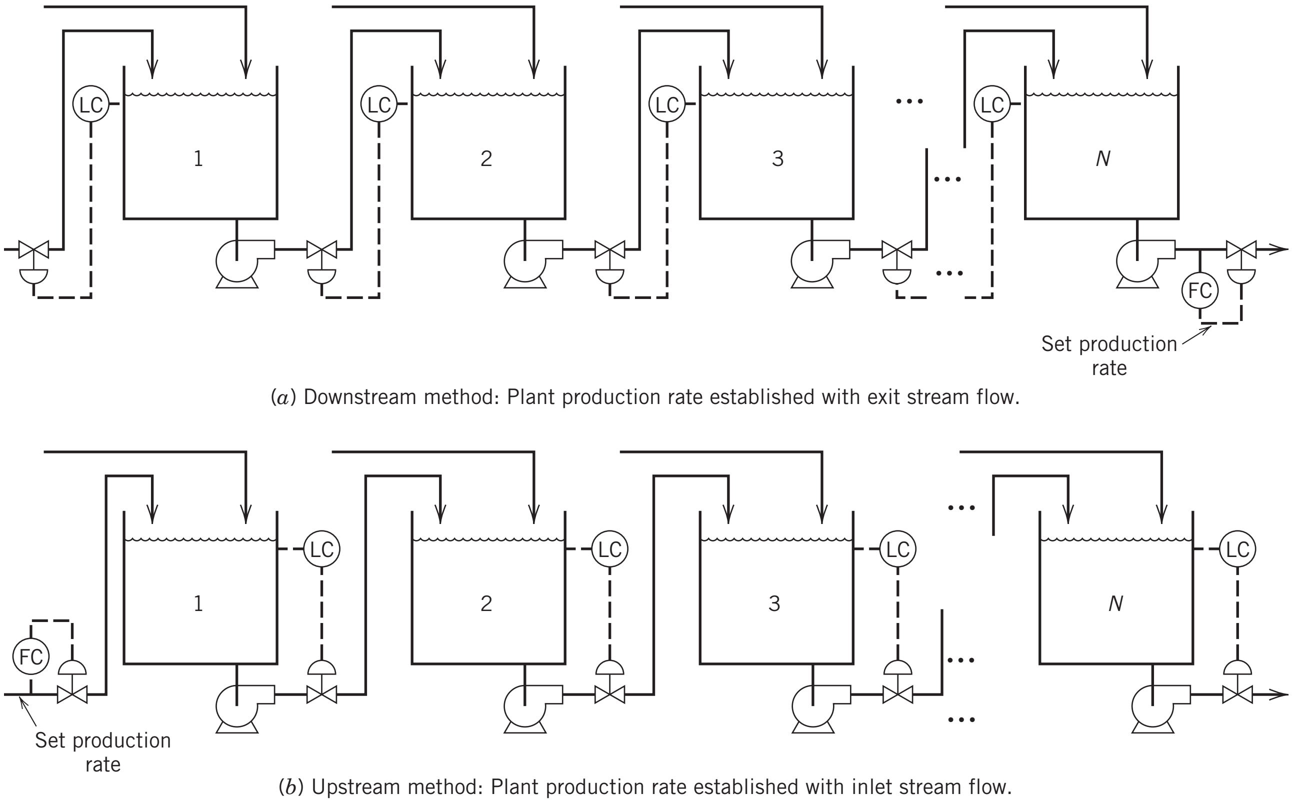


Figure F.1 Train of continuous processing units.

production rates, if there is more than one product), (2) to regulate the liquid level in each unit, and (3) to reduce the effect of disturbances as much as possible. Note that the downstream method has the advantage of fixing the actual product rate, but any flow disturbance to a single unit will tend to propagate successively to upstream units as manipulated flow rates are changed to deal with the disturbance. Also, in this case each additional stream into a unit may have to be regulated at a fixed ratio to one of the streams (the primary manipulated variable) if composition upsets as well as inventory disturbances are to be avoided.

The upstream method has the disadvantage that production rate is established via flow of materials into the first unit. Thus, flow or level disturbances will propagate downstream, eventually affecting the plant's production rate of the desired product, the flow rate from the final unit. Such a situation can be quite undesirable. Consider a bank of extruders or fiber-spinning machines that utilize the product of a continuous polymerization line. An increase in flow rate to the final polymerization unit causes its level to be increased. The resulting increased residence time can lead to increased degradation of the polymer as a result of extended high-temperature processing. In such a situation, excess product may have to be recycled back to an earlier unit and reprocessed, or even sent to "waste." If the flow rate to the final polymerization unit is reduced, one or more extruders/spinning machines may have to be shut down for a period of time to maintain a reasonably constant level in the final unit. Modern processing plants cannot be operated in this manner.

When continuous processing methods first achieved widespread usage in industry, disturbance propagation was reduced by placing surge vessels between key processing units. This arrangement allowed separate control systems to be used for each unit. In Fig. F.2 a reactor and distillation column are separated by a surge vessel. The surge tank prevents flow disturbances from the reactor from upsetting the column, and also prevents short-term production rate changes for the column from propagating back to the reactor. Note that the level in a surge vessel either is not controlled unless it reaches the high or low alarm position. Alternatively, it can be loosely controlled by averaging level control (see Chapter 11). The net effect is to dampen flow disturbances by allowing the level to "float" between low and high limits.

Modern plants are designed to avoid the extra capital and operating costs of surge tanks, related piping, and space in the operations area. Thus, extraneous

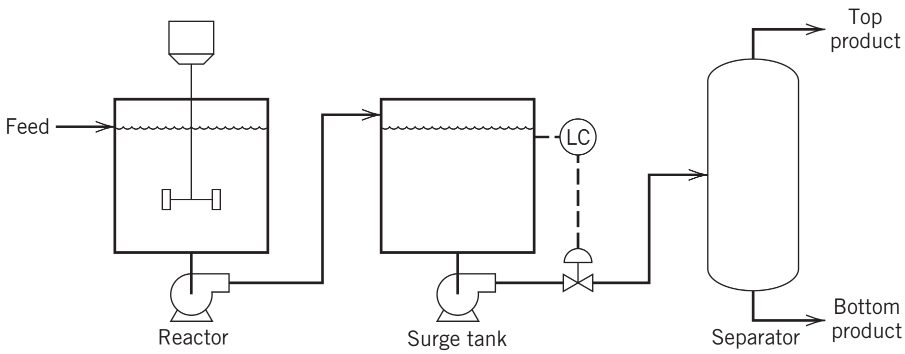


Figure F.2 Use of a surge tank to dampen the propagation of flow disturbances between a reactor and a separator.

vessels, whose only function is to make the plant easier to operate, are normally avoided.

In the continuing search for lower plant operating costs, two other process design techniques often are employed that make plants more difficult to control. One of these techniques is *heat integration*, in which the overhead vapor from one distillation column provides the energy for vaporizing liquid in the reboiler of another column, typically in the same separation train. Recapturing energy in this manner is a major concern in the design of modern processing plants. However, in obtaining the increased energy efficiency available through heat integration, designers must pay close attention to the more complicated plant that results, as noted below and, in more detail, in Section F.3.

Figure F.3 illustrates another commonly employed process design technique, *material recycle*. Here two reactors are connected in series, followed by a flash unit whose vapor product is recycled back to the first reactor. Unreacted reactants concentrated in the vapor stream are recycled to increase the reaction conversion or yield.

Although heat integration and material recycle can significantly reduce plant capital and operating costs, these techniques inevitably increase the amount of interaction among operating units and reduce the control degrees of freedom (see Chapter 12). Nevertheless, appropriate control strategies can deal with such undesirable consequences.

Plantwide control is concerned with designing control systems for large numbers of individual process units that may be highly interacting. Several additional issues arise from these interactions, which further distinguish plantwide control from the control of single units. A hypothetical plant consisting of a reactor and separation unit provides the basis for useful analytical and simulation results that are presented in Section F.3.

F.2 HYPOTHETICAL PLANT FOR PLANTWIDE CONTROL STUDIES

This type of plant has been considered by Papadourakis et al. (1987), and Luyben (1993). Though conceptually simple, the use of recycle considerably complicates steady-state and transient operations.

F.2.1 Reactor/Distillation Column Plant

Figure F.4 illustrates a simple generic plant, an isothermal reactor coupled with a distillation column. A mixture

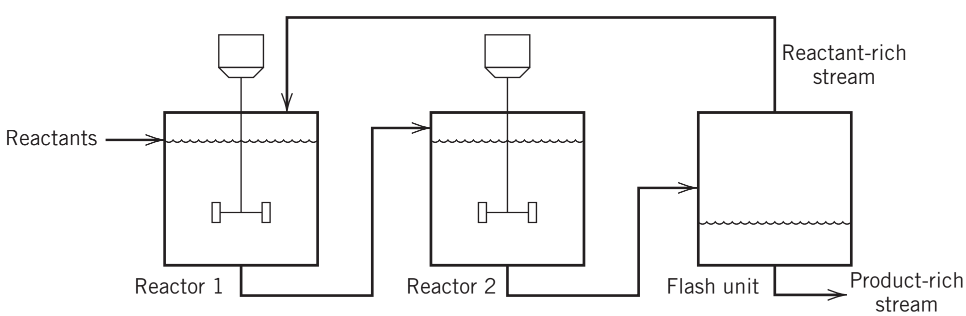


Figure F.3 Use of material recycle to increase reactor yields.

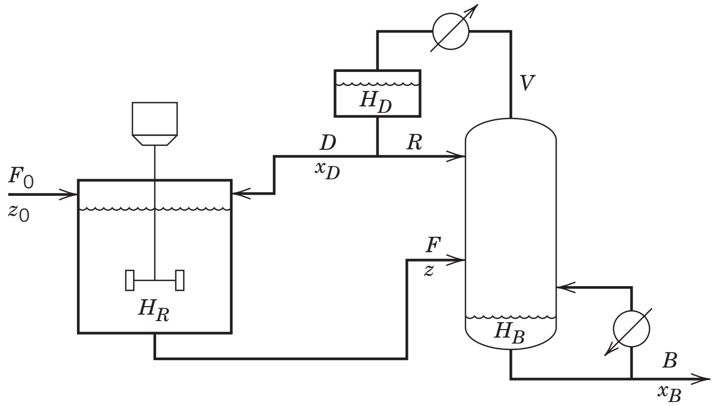


Figure F.4 Reactor/distillation column plant for plantwide control studies.

of two species, mainly A but also some B, is fed to a reactor where the reaction A → B takes place isothermally. The binary distillation column has 20 stages and produces two product streams: an overhead (distillate) stream rich in A and a bottoms stream rich in the desired product B. The A-rich distillate is recycled to the reactor to increase the conversion of A to B.

Table F.1 provides the dynamic model for the two process units. Parameter values for the individual process units and the nominal operating conditions of Luyben (1993) and Wu and Yu (1996) are shown in Table F.2. A number of simplifications are used here:

1. The reaction rate is first-order in A.
2. Reactor operation is isothermal.
3. The column operates with equimolal overflow.
4. Column operation is at atmospheric pressure.
5. Constant relative volatility is used to describe vapor/liquid equilibrium.
6. Each tray represents an equilibrium stage.

The simulation results presented below are based on the 26th-order model in Table F.1 that includes variable liquid holdups in the reactor, the distillate receiver, and the reboiler, but not on the individual equilibrium stages.[1] Thus, the dynamic column model reduces to the steady-state model used in the McCabe-Thiele analysis (Seader and Henley, 1997) if the column accumulation terms are all zero. With a nominal internal reflux ratio (V/D) of 1.455, the column yields a *separation concentration ratio*, $S \triangleq \overline{x}_D/\overline{x}_B$, of 90.5.

The control objective is to maintain the composition of B in the product stream x_B at the nominal value given in Table F.2, despite disturbances in the fresh feed composition z_0 and the feed flow rate F_0. We assume initially that the production rate is established either upstream or downstream of the plant. Later, we discuss ways of accommodating that objective using alternative plant control structures.

F.2.2 Degrees of Freedom Analysis

The 12 process variables in Table F.3 are now considered for control of this plant. A total of six flow rates can be manipulated—three levels and three compositions. As discussed in Chapter 12, the number of control degrees of freedom is usually equal to the number of variables that can be manipulated. Thus, the hypothetical plant has six control degrees of freedom corresponding to the six control valves. They can be used to control a maximum of six measured variables at desired set points (or the levels can be controlled within limits, as discussed in Chapter 20), assuming that no physical or operational constraints are violated.

It is important to recall the dual nature of the flow rates: for example, the fresh feed flow rate F_0 can be used to control reactor level directly (Fig. F.5*a*). Alternatively, if a flow transmitter is placed in this line, F_0 can be controlled to its desired set point (Fig. F.5*b*), or it can be cascaded within a level control loop (Fig. F.5*c*). Recall from Chapter 15 that cascade control does not eliminate a control degree of freedom; the flow rate itself is simply replaced by the set point of the flow controller.

Next, several single-unit control issues for this plant will be considered—for example, whether the reflux flow rate R for the column will be under flow control or used as the manipulated variable to control the reflux drum holdup/level H_D or the distillate composition x_D. Depending on the application, either the bottoms composition x_B can be controlled (Luyben, 1993), or both x_D and x_B can be explicitly controlled to their set points (Luyben, 1994). Several alternative control configurations can be used to accomplish the latter (two-point composition control). In the *material balance configuration*, H_D is controlled by manipulating D, and H_B is controlled by adjusting B. This choice leaves R and V to control, respectively, x_D and x_B. By contrast, in the *energy*

[1]Including variable liquid flow holdup for each stage would increase the model order from 26 to 46. There would be one additional differential equation for each of the 20 stages in the column whose holdup is allowed to vary. Elimination of the very fast liquid flow dynamics can reduce simulation times considerably by eliminating model stiffness, without sacrificing accuracy.

Table F.1 Dynamic Model for Reactor/Distillation Column Plant (Symbol definitions and values provided in Table F.2)

Reactor	
General Information:	Reaction: A $\rightarrow$ B Reaction rate expression is first-order in reactant A. $r_A = -k_R H_R z$
Reactor Model:	$\frac{dH_R}{dt} = F_0 + D - F$ (= 0 for perfect reactor level control) $\frac{d(H_R z)}{dt} = F_0 z_0 + D x_D - Fz + r_A$
Column	
General Information:	Saturated liquid feed is to 12th stage (of 20) numbered from the bottom upward. Equimolal overflow is assumed. A is the more volatile component; assume equilibrium holds for each stage: $y_i = \frac{\alpha x_i}{1 + x_i}$
Column Model:	
Reflux drum:	$\frac{dH_D}{dt} = V - R - D$ (= 0 for perfect level control) $\frac{d(H_D x_D)}{dt} = V y_{20} - R x_D - D x_D$
Stage *i* above feed:	$H_S \frac{dx_i}{dt} = L(x_{i+1} - x_i) + V(y_{i-1} - y_i)$ for $13 \leq i \leq 20$ where $L = R$ and $x_{21} = x_D$ by convention
Feed stage:	$H_S \frac{dx_{12}}{dt} = (L x_{13} - L' x_{12}) + V(y_{11} - y_{12}) + Fz$ where $L' = L + F$
Stage *j* below feed:	$H_S \frac{dx_j}{dt} = L'(x_{j+1} - x_j) + V(y_{j-1} - y_j)$ for $1 \leq j \leq 11$ where $y_0 = y_B$ by convention
Reboiler:	$\frac{dH_B}{dt} = L' - V - B$ (= 0 for perfect reboiler level control) $\frac{d(H_B x_B)}{dt} = L' x_1 - V x_B - B x_B$

balance configuration, the two manipulated variables at the top of the column are switched. Thus, H_D is controlled by R, and x_D is controlled by D. In addition, the control loop pairings at the bottom are switched (Shinskey, 1996).

In order to analyze either column control configuration, we assume for simplicity that the result is perfect control at the desired steady state. In other words, the levels and compositions will be held at the nominal values in Table F.2, while F and z vary. The steady-state material and component balances for the column are

$$\overline{F} = \overline{D} + \overline{B} \tag{F-1}$$

$$\overline{F}\,\overline{z} = \overline{D}\,\overline{x}_D + \overline{B}\,\overline{x}_B \tag{F-2}$$

Equations F-1 and F-2 indicate that fixing the values of x_D and x_B (via perfect control) determines the *steady-state* flow rates $\overline{D}$ and $\overline{B}$ for any values of $\overline{F}$ and $\overline{z}$. Here $\overline{x}_D$ and $\overline{x}_B$ denote the steady-state values of x_D and x_B (0.95 and 0.0105, respectively).

Table F.2 Parameter Values and Steady-State Conditions for the Reactor/Distillation Column Recycle Process (adapted from Wu and Yu (1996))

Reactor	
Fresh feed, F_0	460 lb-mol/h
Fresh feed composition, z_0	0.9 mole fraction A
Reactor holdup, H_R	2400 lb-mol
Recycle flow rate, D	500 lb-mol/h
Recycle composition, x_D	0.95 mole fraction A
Reactor residence time, $H_R/(F_0 + D)$	2.5 h
Specific reaction rate, k_R	0.33 h^{-1}
Distillation Column	
Column feed rate, F	960 lb-mol/h
Column feed composition, z	0.5 mole fraction A
Distillate flow rate, D	500 lb-mol/h
Relux flow rate, R	1100 lb-mol/h
Reflux ratio, R/D	2.20
Bottoms flow rate, B	460 lb-mol/h
Vapor boilup, V	1600 lb-mol/h
Number of equilibrium stages	20
Feed stage	12
Distillate composition, x_D	0.95 mole fraction A
Bottoms composition, x_B	0.0105 mole fraction A
Relative volatility, α	2
Bottoms holdup, H_B	275 lb-mol
Reflux drum holdup, H_D	185 lb-mol
Individual stage holdup, H_S	23.5 lb-mol

Assume that a two-point composition control system has been designed using the material balance configuration. Note that whether a material balance or energy balance column control structure is chosen does not restrict the discussion of plantwide issues below in any way. The column control structure can consist of four single-loop controllers:

Controlled Variable	**Manipulated Variable**
H_D	D
x_D	R
H_B	B
x_B	V

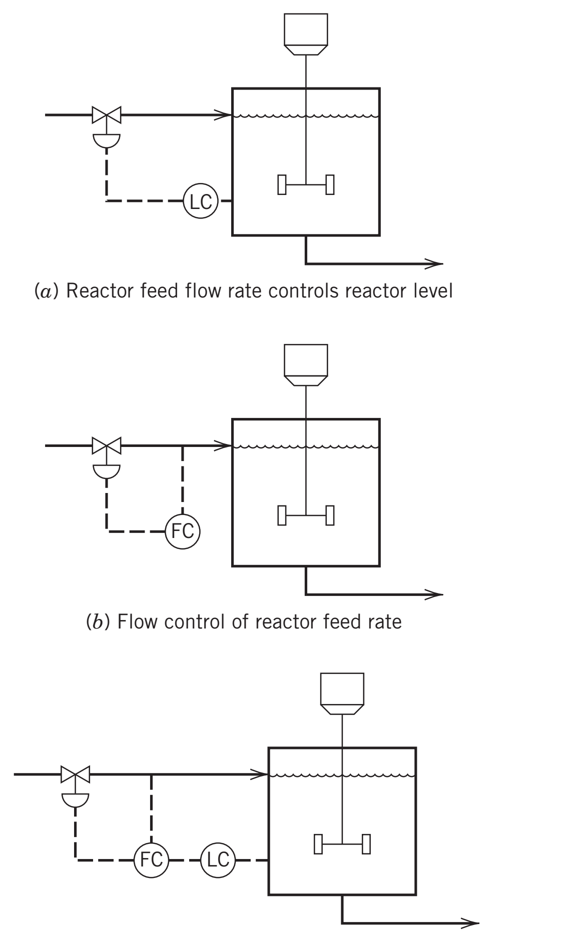
In this analysis, column pressure control has been disregarded, as would be the case, for example, if the column overhead is vented to another vessel at atmospheric pressure. When pressure control must be considered, the flow rate of cooling water to the condenser will be a logical manipulated variable, and an energy balance around the condenser/reflux drum must be added to the model. The number of single-loop controllers would then be five.

If control of reflux drum and bottoms holdups and product compositions is perfect, we can consider the plant as represented in Fig. F.6. Here, D and B can vary, because the two flow rates are manipulated variables; hence, they vary with the column feed flow rate and feed composition whenever the plant is disturbed in order to control x_D and x_B at their set-point values.

Table F.3 Process Variables in the Reactor/Distillation Column Plant Identified as Important for Control

$F_0{}^\dagger$	Reactor feed flow rate
z_0	Reactor feed composition
H_R	Reactor level (proportional to the holdup)
$F^\dagger$	Column feed flow rate (saturated liquid)
z	Column feed composition
H_D	Distillate reflux drum level
$R^\dagger$	Reflux flow rate
D†	Distillate (recycle) flow rate
H_B	Bottoms level
$B^\dagger$	Bottoms (product) flow rate
$V^\dagger$	Reboiler (column) vapor flow rate
x_D	Distillate composition
x_B	Bottoms (product) composition

†Denotes a stream flow rate that can be measured and adjusted by a control valve.



(*a*) Reactor feed flow rate controls reactor level

(*b*) Flow control of reactor feed rate

(*c*) Cascade control of reactor level via secondary controller for feed flow rate

Figure F.5 Multiple uses of a flow variable.

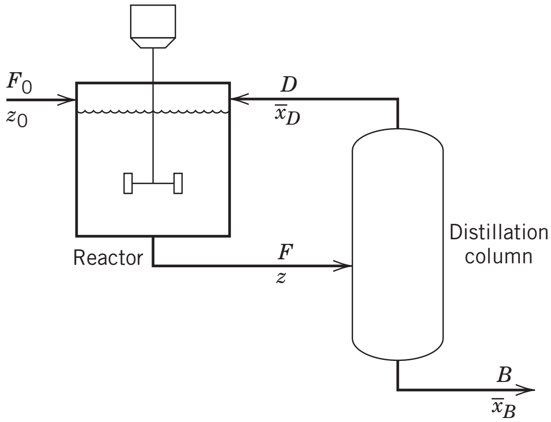


Figure F.6 Schematic diagram of reactor/distillation column plant with perfect control of all three levels and both column product compositions.

In summary, the column controllers have utilized four control degrees of freedom, and eight variables have been removed from the list in Table F.3, leaving

F_0	Fresh feed rate
H_R	Reactor level
F	Column feed rate
z	Column feed composition
z_0	Fresh feed composition

Next, we assume that the primary control objective is to maintain plant operation as much as possible at the set points, despite fresh feed flow rate or composition changes. Thus, F_0 (assuming it is not chosen as a manipulated variable) and z_0 are the disturbance variables. At this point in the analysis, a true plantwide control problem is encountered. Two control degrees of freedom remain, and either F_0 or F could be manipulated to control reactor level. However, it is not obvious whether one choice or the other is better in some sense. In principle, the remaining flow rate (control degree of freedom) could be used to control z or B; for example, if F is available, F_0 is used as a manipulated variable for reactor-level control.

In summary, if we choose to deal with this plantwide control situation by using a multiloop strategy, the consequences of material feedback need to be considered before making any more controller pairing decisions. The reactor/distillation column recycle system, with its two remaining control degrees of freedom, is fairly simple. Nevertheless, it provides several general results about plantwide control strategies.

F.3 INTERNAL FEEDBACK OF MATERIAL AND ENERGY

Processes that include recycle systems have an important design requirement—namely, that there must be an exit path for every chemical species. For example, inert components can be kept at reasonable levels by "bleeding off" a small portion of the recycle stream. However, Luyben (1994) discussed a subtle problem with recycle systems, the *snowball effect*, which is characterized by a large sensitivity of one or more of the variables in a recycle loop to small changes in a disturbance variable. This problem arises from both a small reactor holdup and a particular control structure.

In particular, if changes in fresh feed composition "load the reactor excessively"—that is, beyond its ability to provide the required conversion—then the separator/recycle system will be called on to make up the deficiency. For the case where extra reactor capacity is available through an increase in the reactor level, the particular choice of level/flow control structure within the recycle loop can radically affect plant gains (sensitivities). In the following, we assume that the reactor design is fixed and its level is set at less than full capacity. The question to be considered is how alternative designs of the level and flow loops mitigate the effect of fresh feed flow rate or composition disturbances.

F.3.1 Steady-State Behavior: The Snowball Effect

Because the snowball effect is a steady-state phenomenon, it can be analyzed by considering a steady-state model. We first consider two alternatives for controlling reactor level H_R (Luyben, 1994). For Alternative 1 in Fig. F.7*a*, H_R is controlled by manipulating the column feed rate F (i.e., the reactor effluent rate). For Alternative 2 in Fig. F.7*b*, H_R is allowed to "float" while F is held constant. This strategy is possible because, in theory, the reactor level in this structure is self-regulating (Larsson et al., 2003). For the moment, we assume that the plant production rate is established either upstream or downstream of the plant and analyze these two simple cases to see what insight can be obtained. Later, we consider the implications of setting production rate within the plant.

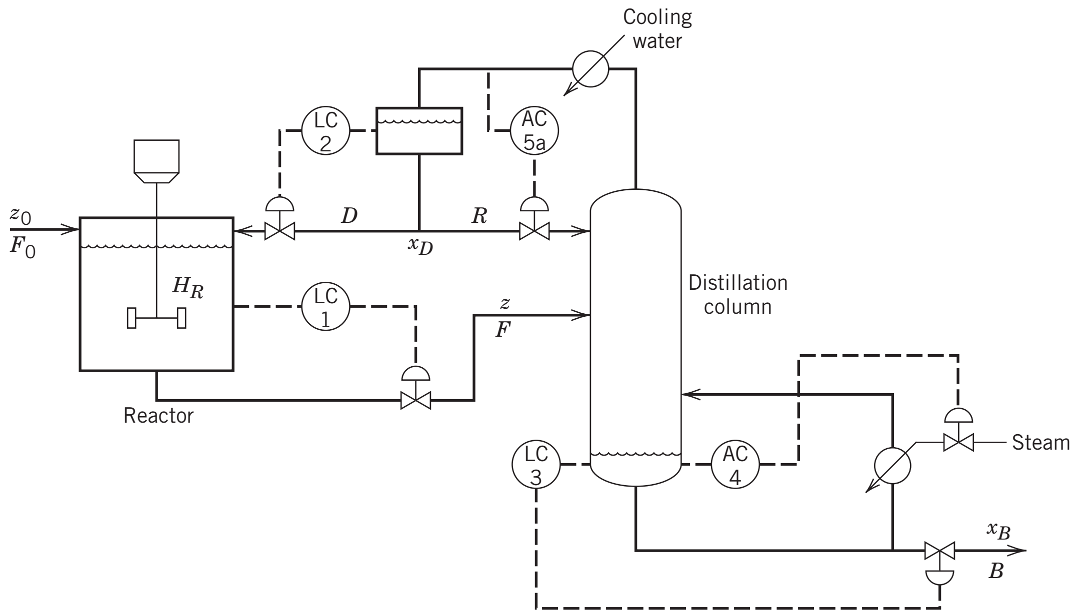
*Alternative 1 (Fig. F.7*a*)*

The key feature in this alternative is that H_R is held constant by manipulating F, with implications for operation of the column. To examine the steady-state sensitivities of key variables within the recycle loop (F, z, and D) with respect to the disturbance variables (F_0 and z_0), consider the steady-state version of the dynamic model in Table F.1.

Reactor

$$\overline{F}_0 + \overline{D} = \overline{F} \quad \text{(F-3)}$$

$$\overline{F}_0 z_0 + \overline{D}\,\overline{x}_D = \overline{F}\,\overline{z} + k_R H_R \overline{z} \quad \text{(F-4)}$$



Alternative 1: H_R is controlled by manipulating F.

Figure F.7a Alternative control structures for the reactor/distillation column plant.

Column. The column equations were developed in the previous section:

$$\overline{F} = \overline{D} + \overline{B} \tag{F-1}$$

$$\overline{F}\,\overline{z} = \overline{D}\,\overline{x}_D + \overline{B}\,\overline{x}_B \tag{F-2}$$

Combining (F-1) and (F-3) (or, equivalently, by writing an overall balance around both units),

$$\overline{B} = \overline{F}_0 \tag{F-5}$$

Similarly, from Eqs. F-2 and F-4,

$$\overline{F}_0\overline{z}_0 = \overline{B}\overline{x}_B + k_R\overline{H}_R\overline{z} \tag{F-6}$$

To simplify the sensitivity analysis, consider the following limiting case:

$$\overline{x}_D \approx 1 \tag{F-7}$$

$$\overline{x}_B \approx 0 \tag{F-8}$$

Now, substitute the approximations of (F-7) and (F-8) into (F-2) to obtain

$$\overline{D} \approx \overline{F}\,\overline{z} \tag{F-9}$$

Similarly, from (F-6),

$$\overline{F}_0\overline{z}_0 \approx k_R\overline{H}_R\overline{z} \tag{F-10}$$

Finally, by manipulating Eqs. F-1, F-3, F-5, F-9, and F-10, the desired expressions for $\overline{z}$, $\overline{D}$, and $\overline{F}$ can be obtained in terms of the reactor fresh feed variables, $\overline{F}_0$ and $\overline{z}_0$.

$$\overline{z} = \frac{\overline{F}_0\overline{z}_0}{k_R\overline{H}_R} \tag{F-11}$$

$$\overline{D} = \frac{(\overline{F}_0)^2\overline{z}_0}{k_R\overline{H}_R - \overline{F}_0\overline{z}_0} \tag{F-12}$$

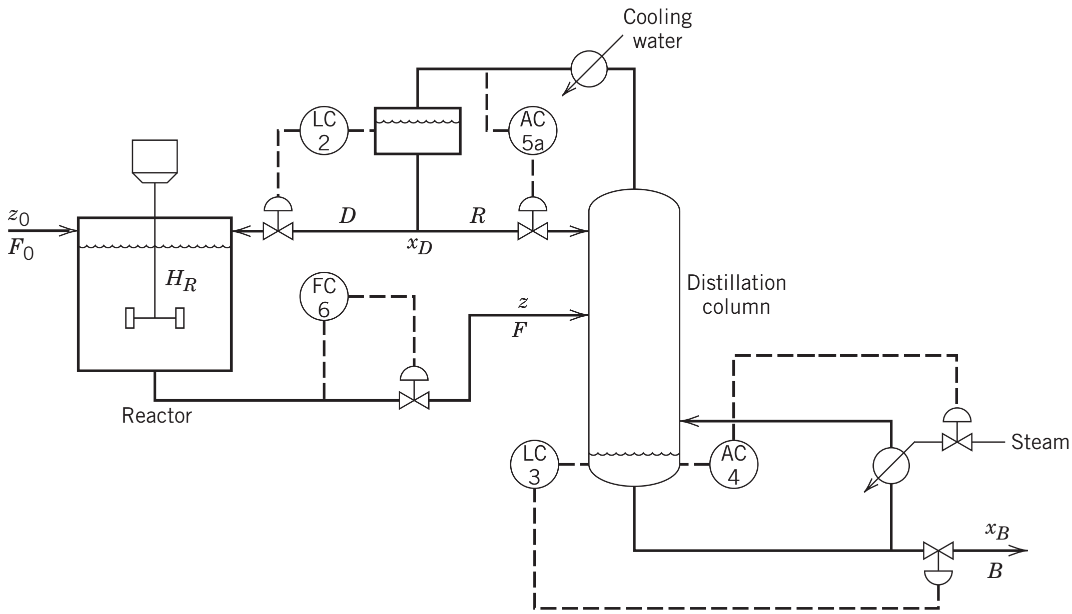
$$\overline{F} = \frac{\overline{F}_0\,k_R\overline{H}_R}{k_R\overline{H}_R - \overline{F}_0\overline{z}_0} \tag{F-13}$$

Equation F-12 indicates that any change in $\overline{F}_0$ or $\overline{z}_0$ will be considerably amplified in $\overline{D}$ because of the presence of the difference between two terms in the denominator, which is significantly increased or decreased by small changes in feed conditions. In a similar fashion, $\overline{D}$ and $\overline{F}$ are sensitive to changes in $\overline{F}_0$ and $\overline{z}_0$. High sensitivity to a disturbance is termed the *snowball effect* by analogy to a snowball, which grows larger as it rolls downhill.

An important point should be emphasized here—namely, that the snowball effect in D and F, while resulting from a particular control structure, is a steady-state phenomenon. In that sense, it is similar to the RGA, which is also a measure of steady-state sensitivities. Luyben (1994) suggested an alternative control method that was intended to reduce the snowball effect in D and F. We investigate a variation of his proposed method next.

Alternative 2 (Fig. F.7b)

In this alternative, F is held constant via a flow controller while H_R is allowed to float. Note that allowing the reactor level (holdup) to vary as disturbance variables F_0 and z_0 change still allows z to change. Luyben (1994) originally proposed controlling H_R with F_0. Larsson et al. (2003) recognized this structure to be self-regulating because H_R adjusts as required to match changes in F. Thus, there is no need to manipulate F_0.



Alternative 2: F is held constant; H_R is allowed to float.

Figure F.7*b* Alternative control structures for the reactor/distillation column plant.

Because F is held constant instead of H_R, as in Alternative 1, we derive approximate expressions for the key recycle loop variables at steady-state ($\overline{D}$, $\overline{z}$, and $\overline{H}_R$) in terms of the disturbances ($\overline{F}_0$ and $\overline{z}_0$). Rearranging Eq. F-3 yields

$$\overline{D} = \overline{F} - \overline{F}_0 \tag{F-14}$$

From (F-9) and (F-14)

$$\overline{z} = \frac{\overline{F} - \overline{F}_0}{\overline{F}} \tag{F-15}$$

Substituting Eq. F-15 into (F-10) yields

$$\overline{H}_R = \frac{\overline{F}\,\overline{F}_0\overline{z}_0}{k_R\,(\overline{F} - \overline{F}_0)} \tag{F-16}$$

Rearrangement of (F-16) yields

$$\overline{H}_R = \frac{\overline{z}_0}{k_R\left(\dfrac{1}{\overline{F}_0} - \dfrac{1}{\overline{F}}\right)} \tag{F-17}$$

Equation F-14 shows clearly that Alternative 2 does not produce a snowball effect in distillate flow rate, because $\overline{D}$ is simply a linear function of $\overline{F}_0$. However, $\overline{H}_R$ now changes in a manner that is proportional to $\overline{z}_0$ and, as is shown below in the examples, is even more strongly related to $\overline{F}_0$. Larsson et al. (2003) showed that the reactor level is intrinsically self-regulating for Alternative 2, a feature that is evaluated in Exercise F.4. In considering Alternative 2, note that a level controller may be incorporated for safety reasons, even if not specifically required—for example, to prevent tank overflow.

Using the equations derived above, we can evaluate and compare quantitatively the sensitivities of key recycle loop variables to sustained changes in either input, z_0 or F_0, for each of the two alternative control structures. Recall that the sensitivity, or gain on a fractional basis, of any output variable y_i *at a specified steady state* $(\overline{x}, \overline{y})$ to a sustained change in an input variable x_j is given by the expression:

$$\left.\frac{\partial(y_i/\overline{y})}{\partial(x_j/\overline{x})}\right|_S = \left.\frac{\partial y_i}{\partial x_j}\right|_S \left(\frac{\overline{x}}{\overline{y}}\right) \tag{F-18}$$

where subscript S indicates that the partial derivatives are evaluated at steady state $(\overline{x}, \overline{y})$.

EXAMPLE F.1

Calculate the sensitivities of the plant recycle flow rate D to changes in both F_0 and z_0 for Alternative 1 and the operating conditions given in Table F.2.

SOLUTION

The sensitivities can be calculated from (F-12). First, the overbars are omitted from these variables, and then the sensitivites are calculated according to Eq. F-18.

$$\left.\frac{\partial(D/\overline{D})}{\partial(F_0/\overline{F}_0)}\right|_S = \left(\frac{\overline{F}_0}{\overline{D}}\right)\left[\frac{(k_R\overline{H}_R - \overline{F}_0\overline{z}_0)(2\overline{F}_0\overline{z}_0) - (\overline{F}_0)^2(\overline{z}_0)(-\overline{z}_0)}{(k_R\overline{H}_R - \overline{F}_0\overline{z}_0)^2}\right] = 2.86 \tag{F-19}$$

and

$$\left.\frac{\partial(D/\overline{D})}{\partial(z_0/\overline{z}_0)}\right|_S = \left(\frac{\overline{z}_0}{\overline{D}}\right)\left[\frac{(k_R\overline{H}_R - \overline{F}_0\overline{z}_0)(\overline{F}_0)^2 - (\overline{F}_0)^2(\overline{z}_0)(-\overline{F}_0)}{(k_R\overline{H}_R - \overline{F}_0\overline{z}_0)^2}\right] = 1.92 \tag{F-20}$$

where subscript S indicates that the partial derivatives are evaluated at the nominal steady-state conditions of Table F.2.

Equation F-19 indicates that the percentage change in D is nearly three times as large as the percentage change in F_0. This is quite a high sensitivity. The second expression indicates that the recycle flow rate is also sensitive to changes in feedstock composition.

EXAMPLE F.2

Repeat Example F.1, analyzing sensitivities for Alternative 2.

SOLUTION

The relative sensitivities for D are obtained from Eq. F-14 using the method in Example F.1:

$$\left.\frac{\partial(D/\overline{D})}{\partial(F_0/\overline{F}_0)}\right|_S = -\frac{\overline{F}_0}{\overline{D}} = -0.92 \tag{F-21}$$

and

$$\left.\frac{\partial(D/\overline{D})}{\partial(z_0/\overline{z}_0)}\right|_S = 0 \tag{F-22}$$

In the latter case, the sensitivity is zero, because D in Eq. F-14 is not a function of z_0.

Thus, the sensitivity of D to F_0 is less than one-third of the value for Alternative 1. With respect to sensitivity to z_0, the Alternative 2 control system completely eliminates the dependence of D on z_0.

Because H_R is allowed to float, we should analyze its sensitivities to F_0 and z_0. From Eq. F-16,

$$\left.\frac{\partial(H_R/\overline{H}_R)}{\partial(F_0/\overline{F}_0)}\right|_S = \frac{\overline{F}_0}{\overline{H}_R}\,\frac{k_R(\overline{F}-\overline{F}_0) - \overline{F}\,\overline{F}_0\overline{z}_0(-k_R)}{[k_R(\overline{F}-\overline{F}_0)]^2} = 4.8 \tag{F-23}$$

and

$$\left.\frac{\partial(H_R/\overline{H}_R)}{\partial(z_0/\overline{z}_0)}\right|_S = \frac{\overline{z}_0}{\overline{H}_R}\,\frac{\overline{F}\,\overline{F}_0}{k_R(\overline{F}-\overline{F}_0)} = 1 \tag{F-24}$$

Although the sensitivity of H_R to changes in z_0 is satisfactory, it is quite large with respect to F_0. Thus, Alternative 2 does not eliminate the snowball effect; it simply shifts it from D and F to H_R. In typical industrial practice, changing the reactor level over a relatively wide range would be undesirable; it normally is held reasonably constant.

Luyben (1994) has investigated these and similar relations for more complex reaction kinetics over a wide range of the disturbances (F_0 and z_0). The snowball effect is not an artifact of the simplifying assumptions employed (e.g., perfect composition control in the column). It appears to be a general effect in recycle systems that can arise from inadequate reactor holdup or a particular choice of the plant inventory/flow control structure. However, before attempting to generalize, we look at two other control structures and their sensitivity characteristics.

Other Level/Flow Control Structures

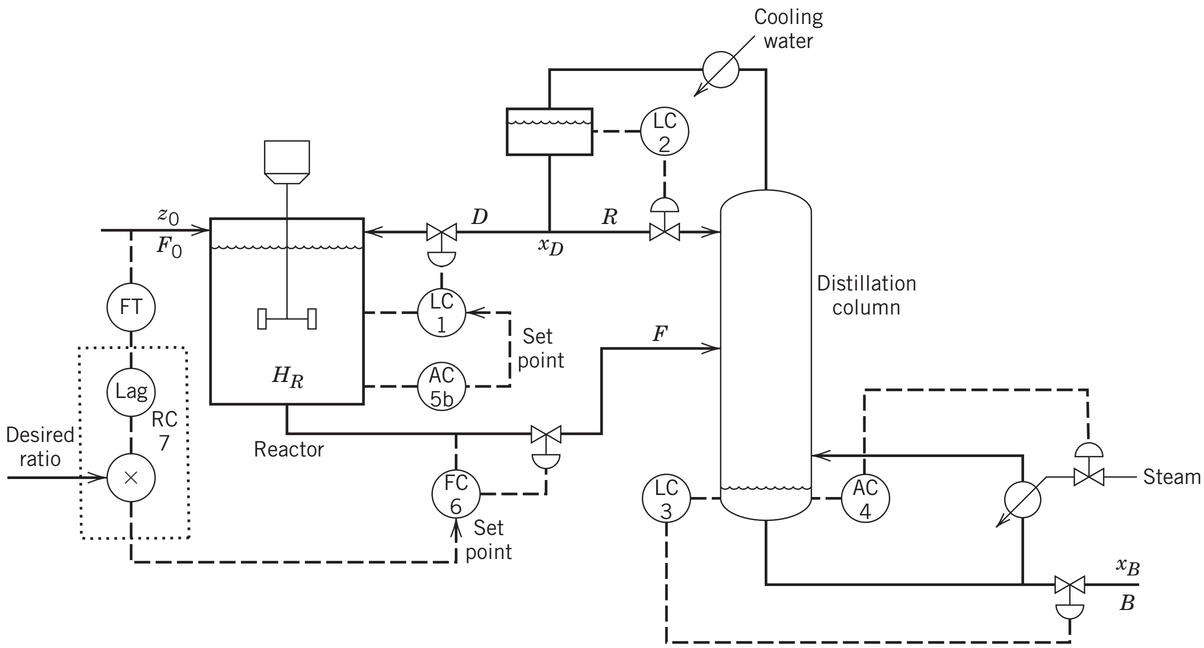
Wu and Yu (1996) identified the major disadvantage associated with the Alternative 2 control structure discussed in Example F.2—namely, that it eliminates snowballing in D but introduces the same effect in H_R, which becomes sensitive to z_0 and F_0. With this objection in mind, they proposed two control structures which they referred to as "balanced" in the sense that feed disturbances are intended to be distributed to both units to smooth out the effects on any particular unit. Their configurations (designated here as Alternatives 3 and 4) include the following features:

Alternative 3 (Fig. F.8*a*). H_R is controlled by manipulating D; however, the set point of the H_R controller is manipulated to control reactor composition z (cascade control). Thus, H_R floats, but only as required to control z.

Alternative 4 (Fig. F.8*b*). H_R is controlled by manipulating D, but the H_R set point is manipulated to control distillate composition x_D. Again, H_R floats, but only as required to control x_D.

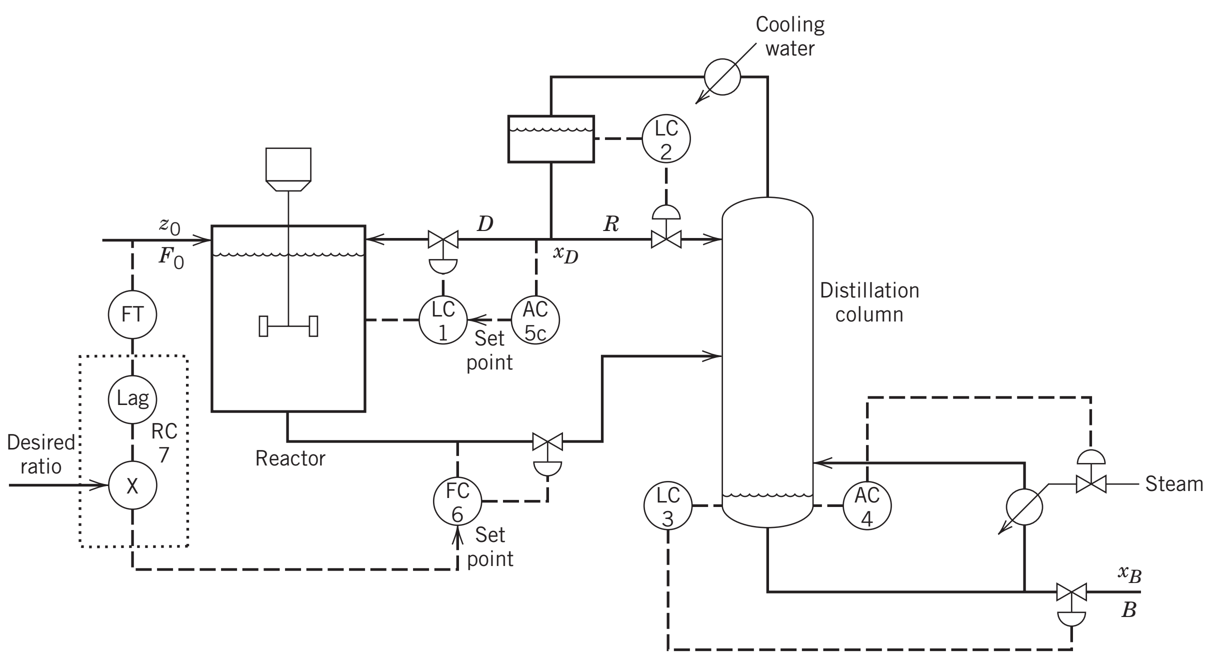
In both of these alternatives, Wu and Yu (1996) proposed that the ratio of F/F_0 be maintained constant by ratio control, a type of feedforward control. In summary, the key features are as follows:

1. D is used to control reactor level in the secondary loop of a cascade controller. H_R, in turn, floats to control one of the recycle loop compositions (z in Alternative 3 or x_D in Alternative 4) by adjusting the set point of the level controller.
2. Unlike Alternatives 1 and 2, disturbance F_0 is now measured and used as input to a ratio controller regulating F; thus, variations in F_0 are reflected directly in changes in F, loading the column somewhat.
3. Variations in F_0 and z_0 will cause changes in both H_R and D if a composition somewhere within the recycle loop is forced to remain constant. Luyben (1994) had suggested that a flow rate be specified



Alternative 3
- H_R is controlled by manipulating D
- z is controlled by manipulating the set point of the H_R controller
- F/F_0 is maintained constant by means of a ratio controller

Figure F.8*a* Additional control structure for the reactor/distillation column plant.



Alternative 4
- H_R is controlled by manipulating D
- x_D is controlled by manipulating the set point of the H_R controller
- F/F_0 is maintained constant by means of a ratio controller

Figure F.8*b* Additional control structure for the reactor/distillation column plant.

Table F.4 A Comparison of Alternative Control Strategies for the Reactor for the Reactor/Distillation Column Plant

Loop Number	Controller Type	Purpose of Control Loop	Controlled Variable	Manipulated Variable Alternatives 1	2	3	4
1	Feedback	Reactor holdup	H_R	F	*Floating*	D*	D*
2	Feedback	Distillate holdup	H_D	D	D	R	R
3	Feedback	Bottoms holdup	H_B	B	B	B	B
4	Feedback	Bottoms composition	x_B	V	V	V	V
5a	Feedback	Distillate composition	x_D	R	R		
5b	Cascade** Primary	Reactor composition	z			$H_{R,sp}$ (Loop 1)	
5c	Cascade** Primary	Distillate composition	x_D				$H_{R,sp}$ (Loop 1)
6	Feedback	Dist. column feed rate	F		$F^{\dagger}$	$F^{\dagger}$	$F^{\dagger}$
7	Ratio	Dist. column feed rate	F			F set point (Loop 6)	F set point (Loop 6)

†Denotes a flow stream adjusted by a flow controller
*Variable controlled in secondary loop of cascade controller (Alternatives 3 and 4 only)
**Primary loop of cascade controller (Alternatives 3 and 4 only)

(fixed) somewhere within each recycle loop. The more complicated Wu and Yu proposal is to specify a composition within the loop while fixing the ratio of the recycle loop flow rate to the reactor feed rate.

Figure F.8 illustrates the control configurations for Alternatives 3 and 4. Both utilize R to control H_D. Table F.4 provides a detailed comparison of all four alternative control configurations. Note that the only features common to all alternatives are the two loops controlling H_B and x_B.

Wu and Yu (1996) performed an extensive steady-state analysis of these control structures using a 2×2 RGA analysis (see Chapter 16). For each structure, one controlled variable is selected from x_D, z and one manipulated variable from R, $H_{R,sp}$. The relative gains are

Case	**Structure**	**Relative Gain (λ)**
Alternative 1	$x_D - R/x_B - V$	2.8
Alternative 2	$x_D - R/x_B - V$	12.2
Alternative 3	$z - H_{R,sp}/x_B - V$	0.78
Alternative 4	$x_D - H_{R,sp}/x_B - V$	0.59

The 2×2 control structure for Alternative 2 is the most interacting. From the results for Alternatives 3 and 4, one might conclude that Alternative 3 is the preferred control structure, because the calculated value for Alternative 4 ($\lambda = 0.59$) is very close to 0.5, where the two pairings would be indistinguishable. Interestingly, dynamic simulation of these four control configurations led to the recommendation of Alternative 4 by Wu and Yu (1996). It exhibited the best closed-loop responses—that is, less interaction between the x_D/H_R primary control loop and other loops—than did Alternative 3 with its corresponding z/H_R loop. When compared with Alternatives 1 and 2, Alternative 4 provided better control of key product composition output x_B and handled larger disturbance changes without violating process constraints. This last point is important: transferring disturbances to more than one plant unit reduces the possibility of intermediate variables' violating a constraint, with the accompanying loss of controllability.

Figures F.9*a* and F.9*b* compare Alternatives 1, 2, and 4, showing the response of several intermediate plant variables (F, D, and H_R) to step changes in F_0 and z_0, respectively. Note that the responses of Alternative 4 in Fig. F.9*a* for feed flow changes lie between Alternatives 1 and 2 as expected; however, the Alternative 4 responses to feed composition changes in Fig. F.9*b* closely resemble those of Alternative 2.

The following generalizations can be made from this case study:

1. Wu and Yu (1996) recommend controlling one composition measurement somewhere in each recycle loop (x_D or z) to accomplish the desired balancing; however, how to couple that composition to key variables in other units, such as H_R, is not clear. Also, they ratio F to F_0 in order to hold the recycle loop flows at constant ratios to F_0. With these changes, Alternative 4 mitigates flow rate changes to reactor level much better than Alternative 2, as expected, but that is not the case for feed composition changes.
2. In general, the best choice of how to "allocate" anticipated disturbances to one or more units in a plant is an unresolved problem.
3. Although all design tools (both steady-state and dynamic) can be important in deciding among alternative control structures, determining the

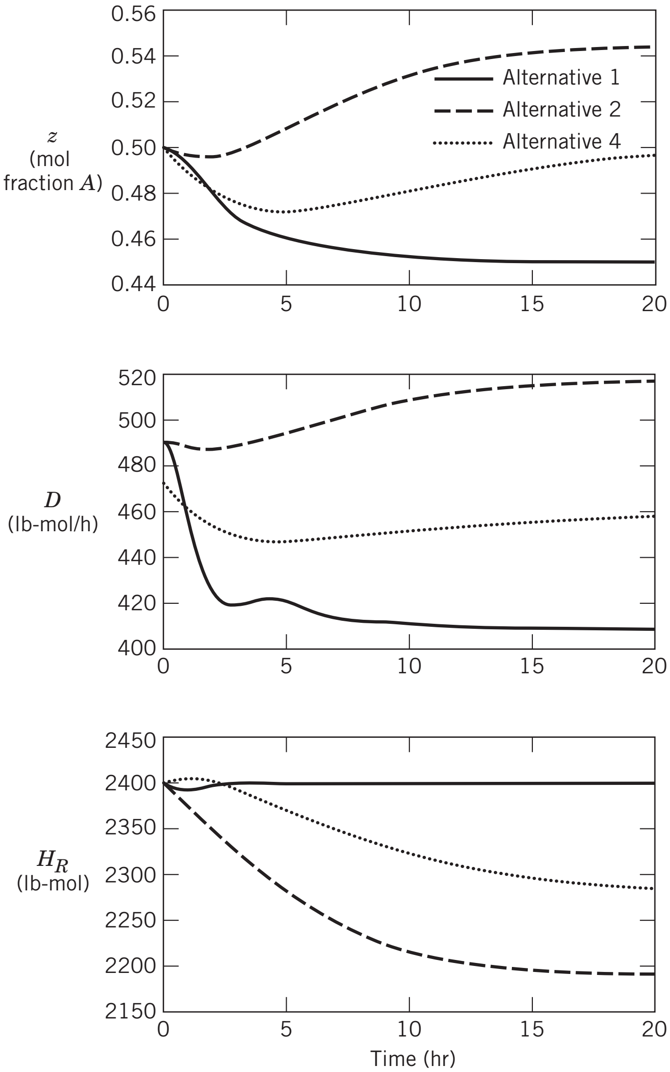


Figure F.9*a* Disturbance response of the reactor/distillation column plant using three alternative flow/level control structures (-10% change in F_0).

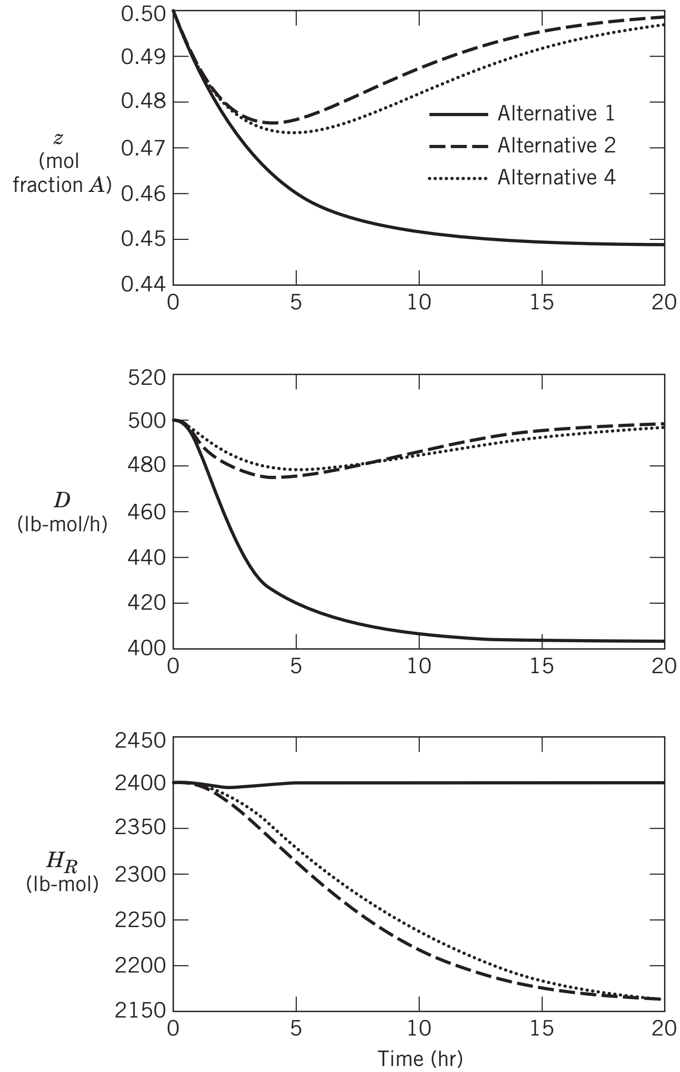


Figure F.9*b* Disturbance response of the reactor/distillation column plant using three alternative flow/level control structures (-10% change in z_0).

"best" structure should involve a rigorous dynamic simulation of the entire plant without using the sort of simplifying assumptions made in this chapter. Of course, final evaluation of the chosen method should be based on plant tests.

This last point, the need to consider process dynamics, is well illustrated by a discussion of how recycling material within a plant can drastically affect its overall dynamics. This topic is considered in the next section.

F.3.2 Transient Behavior: The Slowdown in Overall System Dynamics

A second characteristic of using material recycle and/or heat integration is that the plant may respond to disturbances much more slowly than would be anticipated based on the time constants of individual units.

Consider a simple dynamic system, the reactor/column plant described in Table F.1, and assume that the column dynamics are fast compared to the reactor dynamics. Table F.3 indicates that the holdups in these two units are $H_R = 2{,}400$ lb-moles and $H_B + 20\ H_S + H_D = 930$ lb-moles. Because each unit has the same flow rate F, the mean residence times for the two units are in the ratio of 2,400/930, or approximately 2.5. The effect of chemical reaction normally is to make the reactor time constant somewhat smaller than its mean residence time (see Eq. 3-89); however, the portion of column holdup located directly in the recycle loop, that is, the reflux drum plus the stripping stages, is only about one-half the total column holdup. Thus, the actual ratio of the basic time constants for the two units is probably still on the order of 2.5. As an approximation for illustrative purposes, it is reasonable to treat the column as if it operates much faster than the reactor, so that it essentially is in *quasi-steady-state operation*. In summary, the column operations can be approximated by steady-state equations when the column holdups are small compared to the reactor holdup.

The following analysis assumes that all flows and levels within the plant are constant (perfect level control). The column compositions are not controlled. If they were controlled, as in the previous analysis, the assumption of constant flows would not be valid. The only plant disturbance is the feed composition z_0. With the assumption of quasi-steady-state operation for the column, a simple linear dynamic model for compositions can be developed.

A steady-state model of the column can be derived using the two balances around the entire column (Eqs. F-1 and F-2) obtained earlier. Substitution of Eq. F-5 yields

$$\overline{F} = \overline{D} + \overline{F}_0 \tag{F-25}$$

$$\overline{F}\,\overline{z} = \overline{D}\overline{x}_D + \overline{F}_0\overline{x}_B \tag{F-26}$$

A third equation is obtained from the definition of the steady-state separation concentration ratio (see Section F.2.1) at the nominal operating conditions:

$$S = \frac{\overline{x}_D}{\overline{x}_B} \tag{F-27}$$

For this example, $S = 0.95/0.0105 = 90.5$.

From Eqs. F-25 through F-27, the quasi-steady-state relations relating x_D and x_B to z are

$$x_B \approx \frac{\overline{D} + \overline{F}_0}{\overline{D}S + \overline{F}_0}\, z \tag{F-28}$$

$$x_D \approx Sx_B = \frac{\overline{D} + \overline{F}_0}{\overline{D}S + \overline{F}_0}\, Sz = Kz \tag{F-29}$$

where K is defined as

$$K \triangleq \frac{\overline{D} + \overline{F}_0}{\overline{D}S + \overline{F}_0}\, S \tag{F-30}$$

Note that $K = 1.90$ for the column at nominal operating conditions.

For the case of constant holdup and flow rates, the reactor can be described by an unsteady-state component balance:

$$\overline{H}_R \frac{dz}{dt} = \overline{F}_0 z_0 + \overline{D}x_D - k_R\overline{H}_R z - \overline{F}z \tag{F-31}$$

Substituting Eq. F-29 gives

$$\overline{H}_R \frac{dz}{dt} = \overline{F}_0 z_0 + \overline{D}Kz - k_R\overline{H}_R z - \overline{F}z \tag{F-32}$$

Because (F-32) is an ordinary differential equation with constant coefficients, we can derive the transfer function that relates changes in z to changes in z_0

$$\frac{Z'(s)}{Z_0'(s)} = \frac{K_{Pl}}{\tau_{Pl}s + 1} \tag{F-33}$$

where the subscript Pl denotes "plant." Thus, gain K_{Pl} and time constant τ_{Pl} represent the entire plant (reactor, column, and recycle) and are defined as follows:

$$K_{Pl} \triangleq \frac{\overline{F}_0}{\overline{F} + k_R\overline{H}_R - \overline{D}K} \tag{F-34}$$

$$\tau_{Pl} \triangleq \frac{\overline{H}_R}{\overline{F} + k_R\overline{H}_R - \overline{D}K} \tag{F-35}$$

Substituting (F-25) into (F-35) yields

$$\tau_{Pl} = \frac{\overline{H}_R}{\overline{F}_0 + k_R\overline{H}_R + \overline{D}(1-K)} \tag{F-36}$$

Note that the plant time constant without recycle ($\overline{D} = 0$) reduces to the reactor time constant

$$\tau_{pl}^0 = \frac{\overline{H}_R}{\overline{F}_0 + k_R\overline{H}_R} \tag{F-37}$$

This result is obtained if there is no separation of A and B in the column ($K = 1$), regardless of the magnitude of the recycle flow rate! The effect of having a recycle stream that is richer in reactant than the product stream ($K > 1$) is to slow down the operation of the two-unit plant, because

$$\tau_{Pl} \geq \tau_{pl}^0 \tag{F-38}$$

as a result of the negative $\overline{D}(1-K)$ term in the denominator of (F-36).

This analysis can be performed for the same assumptions that were used in investigating the snowball effect; that is, x_D and x_B are perfectly controlled. In this case, the slowdown effect is even more pronounced, but a simple expression for the plant time constant is not obtained. This exercise is left for the reader.

EXAMPLE F.3

Find the time constant of the reactor/steady-state column model for the operating conditions given in Table F.2 with $K = 1.90$. Determine how much the plant dynamics are slowed by the effect of material recycle.

SOLUTION

From (F-36),

$$\tau_{Pl} = \frac{2400}{460 + (0.33)(2400) + (500)(1.0 - 1.9)} = 3.0 \text{ h} \tag{F-39}$$

Without recycle, from (F-37),

$$\tau_{pl}^{0} = \frac{2400}{460 + (0.33)(2400)} = 1.92 \text{ h} \tag{F-40}$$

Thus, the approximate effect of recycle on this plant is to increase the time constant by

$$\frac{\tau_{Pl}}{\tau_{pl}^{0}} - 1 = \frac{3.0}{1.92} - 1 = 0.56 \text{ or } 56\% \tag{F-41}$$

This result means that a change in z_0 will take 56% longer to work its way through the system with recycle than it would without recycle. Kapoor and McAvoy (1987) provide a more general analysis of how internal recycle affects the time constants of a distillation column.

It is interesting to generalize the results of this example:

1. Any multi-unit plant with a recycle stream from a separation unit is likely to exhibit slower dynamics. Just as negative feedback normally speeds up the process response, the positive feedback of material in this recycle stream slows down the response.
2. The process response becomes slower as either the degree of separation or the recycle flow rate increases.

F.3.3 Propagation and Recirculation of Disturbances

A third major effect often encountered with recycle and heat integration is the propagation of disturbances from unit to unit, and their recirculation around internal process flow paths. To understand why this plantwide control issue differs so substantially from single-unit issues, it is worthwhile to review briefly the objectives of single-unit regulation.

In Chapter 12 we indicated that one desirable effect of using feedback control to attenuate disturbances in a process unit is to transfer these variations to a utility stream. For example, if a reactor temperature is disturbed, the cooling water flow rate will be changed by the temperature controller so as to reduce the effect of the disturbance. Even so, some variation in reactor temperature inevitably will remain, and this will propagate to downstream units as a disturbance.

A minor side effect of these actions is that the supply header temperature itself will change slightly as cooling water demand is raised/lowered by actions of a reactor temperature controller. Although utility supply systems are built with their own internal controllers, and these are designed to attempt to regulate the utility outputs in the face of process disturbances, it is not possible to attenuate utility disturbances entirely. These propagate directly throughout the plant.

In older plants, surge tanks were used to damp flow variations between units. Material holdup in a surge tank can also serve as a thermal capacitance and thus reduce effluent temperature variations; only reduced flow and temperature variations propagate to downstream units. In today's more highly integrated plants, containing material recycle and/or heat integration but little surge capacity, unattenuated disturbances propagate directly to downstream units, even to adjacent (coupled) units and to upstream units.

F.4 INTERACTION OF PLANT DESIGN AND CONTROL SYSTEM DESIGN

In the past, when continuous processing plants were designed to be much less interacting than now, it was possible to complete the plant design before considering control system design. After the proposed plant's flowsheet and equipment specifications were completed, process control engineers were responsible for specifying instruments and controllers. By providing an excess of measurements (instruments) and control valves, plus a feedback controller for every important process variable, the control system designer was reasonably sure that the new plant could be started up and controlled. Continuous processing plants designed or retrofitted today no longer can utilize a sequential design process in which plant design is followed by control system design (Keller and Bryan, 2000), nor can designers specify redundant equipment, except for safety purposes.

Without careful attention to design, highly integrated plants may have too few control degrees of freedom, which makes them difficult to start up and operate safely. For example, in designing the heat exchanger and related equipment for heat integration, the heating and cooling loads first must be approximately balanced by the process designer. Then the designer must establish whether the approach temperatures are satisfactory to meet the heat transfer requirements with a reasonably sized heat exchanger; in this step, adjustment of column operating pressures may be required (Seider et al., 2003). Because the energy supply capability in one unit usually will not balance the demand in another unit exactly, a "trim exchanger" (small heat exchanger sized to make up the difference in heating/cooling capability) generally has to be provided in the steady-state design.

Note that introducing a heat integration scheme also causes two control degrees of freedom to be "lost": the cooling water flow rate control valve that would have been located in the Column 1 condenser, plus the steam control valve that would have been used in the Column 2 reboiler. If process control engineers are not involved in the plant design process from the beginning, the critical process dynamic and control evaluations may be omitted that would provide such information and an opportunity to resolve any problems (Keller and Bryan, 2000). In short, a suitably sized trim unit must be available to make up for any steady-state heating/cooling deficiency plus lost control degrees of freedom necessary for normal operations. It also can assist in start-up and shutdown operations.

The control system designer must determine whether a proposed plant design will be controllable and operable (Fisher et al., 1988b; Downs and Ogunnaike, 1995). For example, highly integrated distillation columns can cause problems in a number of ways:

1. One or both column products cannot be controlled at the desired set point(s).
2. Disturbances in the Column 1 overhead cannot be prevented from propagating to Column 2.
3. The "lost degrees of freedom" from plant integration need to be restored by the addition of one or two trim heat exchangers operated and controlled using plant utility supplies.
4. The plant cannot be started up easily because of the need to have Column 1 "hot" before Column 2 can be brought into service.

This chapter has provided a brief overview of process integration issues and possible solutions. For a much broader discussion of the topic of heat integration, the reader should consult Douglas (1988) or Seider et al. (2008). For a more extensive development of control system design issues, including a number of simulation case studies, Luyben (2002) should be consulted.

SUMMARY

For new process designs, the control system designer may have little precise knowledge of how to control the proposed plant. Because the plant design may never be replicated, there will be little incentive to spend thousands of hours designing and optimizing the control system structure, as would be done, for example, for a new airplane design. Whether or not the final control structure will be successful depends to a large extent on the knowledge, skill, and intuition of the control system design team. The plant initially can be considered as a collection of reasonably well-understood processing units, but it can operate quite differently than would be expected from knowledge only of its individual units considered separately.

In this chapter, we have introduced some basic plantwide issues that are fundamentally based on multi-unit interactions. These topics have included steady-state issues (sensitivities), dynamic issues (settling times of integrated plants vs. individual units), and the propagation of disturbances from unit to unit in highly integrated plants that involve recycle of material and of energy.

In the next Appendix (Appendix G), we present a systematic procedure that can guide the development of plantwide control system designs. The goal is to design a viable control system structure for a new processing plant that has a high probability of working satisfactorily when actually installed.

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EXERCISES

F.1 Figure EF.1 illustrates two CSTRs in a chemical manufacturing plant. Reactants A and B must be fed to the first stirred-tank at a constant molar ratio. Reactant C is introduced to the second stirred-tank at a constant molar ratio to reactant A. Five control valves are available for purposes of controlling the plant production rate and concentrations. Flow rates shown in the figure are in mass units. Reactor volumes are constant.

The assumed reaction kinetics are:

$$A + B \rightarrow D$$
$$C + D \rightarrow E$$

If each reaction goes to completion in its respective reactor, how can you control the plant production rate of the desired product E using each of the five valves? Specify how you would use ratio controllers to maintain the desired stoichiometry in each case, and explain the advantages and disadvantages relative to the other possible locations.

Note: *In the following exercises, a Simulink model is used to approximate the reactor and distillation column units discussed in this chapter. Information is given in Appendix H.1.*

F.2 In this exercise, you will evaluate the individual units at the nominal steady state for purposes of understanding how the plant would operate *without recycle*. Use Simulink to simulate the full differential equation model given in Table F.1. Then, for purposes of this problem only, "tear" the recycle stream to the reactor—that is, disconnect the distillate line and replace it with a constant stream to the reactor that is set at the recycle stream's nominal conditions of flow rate and concentration.

(a) Using a material balance control configuration and any of the techniques discussed in Chapters 11 or 12, find P or PI controller settings that will regulate the liquid levels in the reboiler and the reflux drum with little overshoot.

(b) Complete the column control structure by finding PI controllers that will satisfactorily maintain the distillate and bottoms composition, again with little overshoot. Test your column level/composition control system by making small step changes in the column feed flow rate and composition.

(c) In a similar manner, develop a P or PI controller for reactor level using F as the manipulated variable. (Note that level controller settings obtained using F_0 for the manipulated variable will be identical to those using F.) Again, test your reactor level control system by making small step changes in the feed flow rate and composition.

For each of the following exercises, either work Exercise F.2 first or use controller settings similar to those provided with the parameters and Simulink model of the two-component plus recycle process in Appendix H.1.

F.3 Starting with a Simulink model of the recycle process, implement a reactor level controller using F as the manipulated variable. Confirm via simulation that control scheme Alternative 1 works effectively for a step change in F_0. If necessary, detune any of the controllers to keep oscillations to a minimum.

F.4 Starting with a Simulink model of the recycle process,

(a) Place a tightly tuned flow controller on F. Confirm, via a step change in F_0, that Alternative 2 is self-regulating; that is, that the level in the reactor automatically seeks a suitable steady-state value if the reactor feed flow rate is subjected to a sustained change.

(b) Show that this level is identically equal to the value given by Eq. F-16.

F.5 Luyben's original proposal (1994) for the Alternative 2 control structure incorporated a reactor level controller using F_0 as the manipulated variable. However, the level controller prevents specifying the plant's production rate by a flow controller on F_0, as can be done with Alternative 1. To deal with the problem that arises when F_0 is allocated for level control, Luyben proposed that the steady-state relation given by Eq. F-17 be rewritten to provide a type of

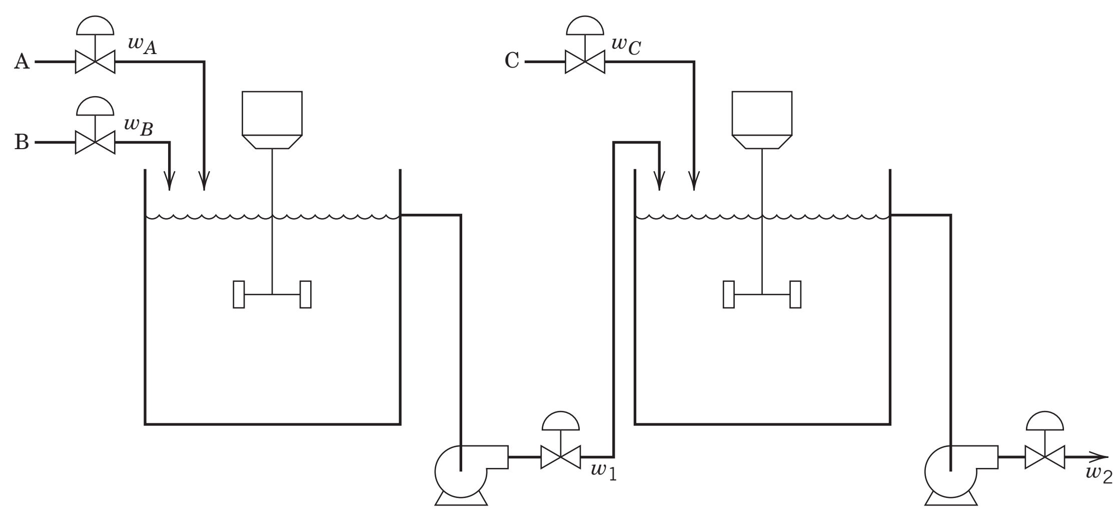

Figure EF.1

feedforward control based on measurements of z_0 and F_0. F_{sp}, the set point of the reactor effluent flow controller, can be used to approximate F. The set point for the reactor level controller would then be

$$H_{R,sp}(t) = \frac{z_0(t)}{k_R\left(\dfrac{1}{F_0(t)} - \dfrac{1}{F_{sp}}\right)}$$

In the following steps, you are to evaluate Luyben's proposed alternative. If necessary, detune any of the controllers to keep oscillations to a minimum.

(a) Determine how Luyben's proposed alternative structure responds to a step change in z_0.

(b) Implement Luyben's proposed feedforward controller—that is the equation above—and implement a similar step change in z_0.

(c) What can you say about the speed of response of this controlled plant with and without the feedforward controller?

(d) What are the advantages and disadvantages of Luyben's proposal?

F.6 Implement Alternatives 3 and 4 using the Simulink model. This will require tuning a cascade loop to control composition in each case.

(a) Evaluate the response of these controlled plants for a step change in F_0 without using the ratio controller proposed by Wu and Yu.

(b) Repeat (a) with a controller used to ratio the column feed flow rate to the reactor feed flow rate. How do the responses in (a) and (b) compare?

F.7 Evaluate any two of the four alternatives we have looked at in this chapter, (i.e., Exercises F.3, F.4*a*, and/or F.6).

(a) Compare the responses of each control structure to step changes in reactor feed flow rate.

(b) Compare the sensitivities of each alternative to changes in this variable.

F.8 The recycle plant discussed in this chapter utilizes a composition-only model; that is, thermal effects are neglected. Appendix H contains equations and parameters that can be used to model temperature effects in the reactor. Implement the cooling coil equations for the reactor and, using the cooling water flow rate as manipulated variable, design a PI controller that will control reactor temperature. Test your plant's response to a step change in reactor feed flow rate using any of the control structure alternatives discussed in this chapter. What can you conclude regarding the effect of a well-tuned controller for reactor temperature on the responses of the other system variables such as F and D?