Appendix G

Plantwide Control System Design

In this chapter, we describe a hierarchical design procedure that can be used to develop multiloop and multivariable measurement and control strategies for plantwide control systems. The procedure assists the engineer in determining how to choose the best controlled, manipulated, and measured variables in the plant, when to use advanced control techniques such as MPC, and how to select appropriate multiloop control structures with minimum interactions among the coupled processes in the plant. The proposed design procedure is based on the hierarchy of process control activities described in Chapter 1, the control system design guidelines discussed in Chapter 12, RGA and SVA multivariable methods presented in Chapter 16, the model predictive control approach of Chapter 20, plantwide control concepts of Appendix F, and designers' experience. It is important to realize that the design of plantwide control systems is an art as well as a science. Typically, more than one design will be satisfactory; thus, there is no single solution to the design problem. Furthermore, a design procedure generally involves iteration of individual steps until a satisfactory design results. Thus, the application of a systematic design procedure, such as the hierarchical approach of this chapter, produces preliminary designs that are subject to further exploration and refinement. Simulation methods should be employed to examine alternative control configurations while exploring the effect of controller tuning on the response of key process variables. The hierarchical procedure recommended in this chapter is illustrated by a case study.

The goal is a plantwide control system design that is no more complicated or expensive than necessary and that, when built, can be operated easily by typical plant operators. Ultimately, the only definitive way of validating a selected plantwide control system design is by plant tests and by the operating plant's performance.

G.1 PROCEDURES FOR THE DESIGN OF PLANTWIDE CONTROL SYSTEMS

The design of a plantwide control system consists of four major steps:

1. The overall specifications for the plant and its control system are stated.
2. The control system structure is developed. This step includes selecting controlled, measured, and manipulated variables; choosing multiloop or multivariable control; deciding how to control production rate, product quality, and inventories; and handling operating constraints. Decomposition of the plantwide control problem into smaller problems for the purpose of analysis may also be employed here.
3. Design is followed by a detailed specification of all instrumentation/hardware and software, cost estimation, evaluation of alternatives, and the ordering and installation of equipment.
4. Following design and construction of the plant, plant tests, including startups, operation at design conditions, and shutdowns, are carried out prior to commissioning of the plant.

This chapter is concerned with the first two steps, beginning with the plant control system design specifications.

In principle, a comprehensive *top-down* formulation could be used to develop the required plantwide control design. We assume that general requirements for the plant, such as product specifications and production rates, have been established at the plant, division, or corporate level. The specifications for plant operating conditions have been developed by the plant design group working in collaboration with product development and process control specialists. Starting with the above specifications plus knowledge of the potential measured, manipulated, and controlled variables, optimization methods could be employed to develop the control system design based on a comprehensive dynamic model of the plant. Unfortunately, such an

approach is impractical because of the large number of process variables involved in modern processing plants. On the other hand, some aspects of a top-down design approach may be quite useful as part of a realistic design procedure.

The traditional design procedure used for industrial control systems has been a *bottom-up*, unit-by-unit approach. Even though it incorporates systematic methods to develop the control structure, this approach also relies on heuristic design methods and rules of thumb developed from previous designs and the experience of both the process and control system design groups.

Control system design has been organized into a logical, sequential, process-oriented methodology by numerous control researchers—for example, Buckley (1964), Downs (1992), Luyben et al. (1998), Larsson and Skogestad (2000), and Stephanopoulos and Ng (2000). However, it would be naive to assume that any proposed design procedure can generate a suitable plantwide control system design after one pass through a set of sequential steps. At virtually every step of a given procedure, alternative choices will present themselves, each leading to a different final design alternative. The knowledge, skill, intuition, and persistence of the plantwide control design team will always be key elements in the design process. In addition, the control design specifications and models used for design will exhibit uncertainty, which means that multiple designs may need to be developed (Skogestad, 2002).

An effective way to make the large number of decisions is to organize the procedures in a generally hierarchical manner. Thus, detailed studies should not be undertaken until important general questions have been answered. Skogestad (2002) has developed a design procedure based on the intrinsically hierarchical nature of plantwide control systems while incorporating the best aspects of top-down and bottom-up design approaches. As shown in Fig. G.1, the most critical control tasks deal with the safety system (Chapter 9) and with regulating the integrating response modes usually associated with liquid levels (holdups in the vessels). Thus, the basic objective at Levels 1, 2, and 3 is to provide safe, stable control of the plant. Level 4 is concerned with economic optimization of plant operating conditions, and this step is usually decoupled from the control system operation.

Missing from many control system design methodologies, even hierarchical ones, is the important role that *decomposition* and *decentralization* play in a plantwide design approach. Procedures that lead to decomposition of the overall design into smaller subproblems can be advantageous. Even highly integrated plants do not require a multivariable approach linking all of the controlled variables with all of the manipulated variables.
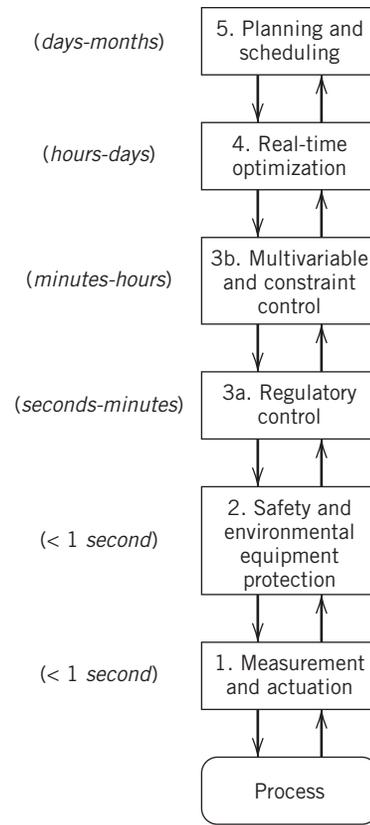
Figure G.1 Hierarchy of process control activities.

The extent to which a plantwide control system can be decentralized into smaller control systems designed to work at the process unit level invariably determines how easily the control system can be designed, tuned, and understood by plant operators. Decentralized control system designs generally are more robust when operating conditions change and are more tolerant to individual component failures.

G.2   A SYSTEMATIC PROCEDURE FOR PLANTWIDE CONTROL SYSTEM DESIGN

Table G.1 provides the key steps in a systematic procedure recommended here for design of plantwide control structures. It is based on the combined top-down/bottom-up approach of Larsson and Skogestad (2000) and Skogestad (2002) and the hierarchical organization that generally matches Fig. G.1. The proposed systematic plantwide control design approach consists of the four major steps shown in Table G.1.

G.2.1   Control System Design Objectives

Plant operating/control objectives must be established at the outset of the design process. Two categories of

Table G.1 Recommended Procedure to Design a Plantwide Control System

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  I. Specify the control system design objectives.
      A. State the plant production, economic, and control objectives, including composition and production rates of all products.
      B. Identify process constraints that must be satisfied, including safety, environmental, and quality restrictions.
 II. Perform a top-down analysis.
      A. Identify the process variables, control degrees of freedom, control structure, and options for decomposition.
      B. Establish the overall control structure (in conceptual form).
III. Develop a bottom-up design.
      A. Develop a strategy for regulatory control.
      B. Examine the potential of applying advanced control strategies.
      C. Evaluate the economic benefits of real-time optimization.
 IV. Validate the proposed control structure.
      A. Perform a final control degrees of freedom analysis. Check the allocation of the N_{FC} degrees of freedom.
      B. Check control of individual process units.
      C. Check the effect of constraints and disturbances on manipulated and controlled variables.
      D. Simulate control system performance for a wide range of conditions.

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information must be provided: (1) plant production and control objectives and (2) process constraints (Step I).

Step I. Specify the control system design objectives.

A. State the plant production, economic, and control objectives, including composition and production rates of all products.

B. Identify process constraints that must be satisfied, including safety, environmental, and quality restrictions.

In this chapter we use box outlines to summarize the tasks in each step. A full case study and references to related work are provided to clarify the detailed procedures.

G.2.2 Top-Down Analysis

The top-down analysis identifies both the scope and complexity of a plantwide control design project and its

control structure. (See Step II for an outline of individual tasks.) Among the conceptual issues considered at this point in the design are where to control the key production and quality measurements, how the overall plant might be divided into smaller subsystems (*decomposition*) to simplify control system design, and where variable coupling or constraint handling may justify, or even require, the use of multivariable control. For example, it is important to identify certain subsystems whose control system designs cannot be developed separately because the processes are so closely coupled, such as in heat integration.

Step II. Perform a top-down analysis.

A. Identify the process variables, control degrees of freedom, control structure, and options for decomposition.
  1. Identify the potential controlled variables.
  2. Determine how the CVs can be measured or inferred, and identify other process variables to be measured.
  3. Select the potential manipulated variables.
  4. Perform a preliminary control degrees of freedom analysis (compare the numbers of potential manipulated and controlled variables).
  5. Identify the source and nature of the significant disturbances that must be mitigated.
  6. Perform a structural analysis based on a steady-state model, select the final controlled and manipulated variables, and evaluate the possibilities for decomposition of the control problem.

B. Establish the overall control structure (in conceptual form).
  1. Identify where the production rate of each product will be measured and controlled.
  2. Identify how quality will be measured for each product, and how quality will be controlled.
  3. Determine how each recycle loop throughput/composition will be controlled.
  4. Specify how the constraints will be satisfied.
  5. Determine how major disturbances will be handled.
  6. Analyze the energy management scheme, and indicate conceptually how it will be controlled.

Note that the number of control degrees of freedom can be influenced by constraints imposed during the control system design process. Once the scope of the

design problem has been determined and guidance is available to begin the control system design task, it is much easier to develop a preliminary (conceptual) control structure. An important goal at this level is to utilize structural analysis techniques (SVA, RGA) subject to the availability of a steady-state and/or a dynamic model of the plant. As part of the conceptual design of the plant, one should attempt to identify the most effective measured and manipulated variables, and identify any highly decoupled or highly interacting process units that will need special attention.

Normally, careful consideration of the process design itself will indicate how the control system of an entire facility (for example, a refinery) might be decomposed to control systems for its individual sections—for example, the gas treatment section or the separations section. In addition, a top-down analysis generally will provide further clues as to how the overall control problem can be reduced to a set of smaller problems. From this discussion, it should be clear why recycle, heat integration, and constraint handling systems are best dealt with conceptually before decomposition decisions are made.

After completing the top-down design step, the designer should have an excellent overview of the plantwide control system design task in terms of subsystems of processes rather than as many single-unit control systems. For example, a train of distillation columns coupled via heat integration is probably best considered as a single subsystem for purposes of control. Of course, several single-loop controllers may have already been identified in the top-down sequence to deal with production rate and quality variables, and a structural analysis may point to certain specific pairings of the CVs and MVs that will be worth considering. However, a detailed design of the control system is properly treated in the bottom-up procedure considered next.

G.2.3 Bottom-Up Design

Once a conceptual control structure has been developed and the plant has been decomposed into subsystems, the control design procedure reverts to a traditional bottom-up approach. However, there are good reasons to treat the different control activities in a multilevel hierarchy, as shown in Fig. G.1. The first task in Step III is to identify the essential controllers, those that are absolutely required. The safety and regulatory levels in Fig. G.1 enable safe and stable operation of the plant. The advanced control functions are handled at Level 3 and keep the controlled variables close to their optimum set points through standard methods such as cascade, ratio, feedforward, and multivariable control. Level 4 in Fig. G.1

considers the real-time optimization of the process operations. The purpose of control at this level is to choose operating conditions that meet overall objectives in an economically optimum fashion.

Step III. Develop a bottom-up design.

A. Develop a strategy for regulatory control.
  1. Specify how the control system will respond to unsafe or abnormal operating conditions and deal with constraints.
  2. Identify control loops to regulate production rates and inventories.
  3. Identify control loops that will mitigate major disturbances.

B. Examine the potential of applying advanced control strategies.
  1. Evaluate the use of enhanced single-loop control strategies, including feedforward, ratio, cascade, and selective control schemes.
  2. Employ MIMO control for highly interactive processes.

C. Evaluate the economic benefits of real-time optimization.

G.2.4 Validation of the Plantwide System Design

Finally, the design of the plantwide control system needs to be checked carefully and validated. At this point, a series of checks should be performed to ensure that the plantwide control structure is complete, is internally consistent, and functions appropriately, as shown in Step IV.

Step IV. Validate the proposed control structure.

A. Perform a final degrees of freedom analysis. Check the allocation of the N_{FC} degrees of freedom.
B. Check control of individual process units.
C. Check the effect of constraints and disturbances on manipulated and controlled variables.
D. Simulate control system performance for a wide range of conditions.

After Steps I–IV are completed, a number of other tasks must be finished to complete the control system design. They include detailed specification and costing of instrumentation and control equipment, purchase, installation, and checkout. Then the control system must be evaluated during actual plant operation. The

final step is to certify that the plant and control system meet safety, environmental, production, and quality requirements (the *commissioning* step).

G.3   CASE STUDY: THE REACTOR/FLASH UNIT PLANT

We now apply the principles from the previous two sections to a specific case study—a reactor/flash unit plant with recycle similar to the plants discussed in Section G.2 and by Robinson et al. (2001). The plant consists of a reactor, flash unit, and recycle (surge) tank as shown in Fig. G.2. The reactor produces a product C from two feed streams consisting of pure A and B, which contains a small amount of *D*. The reaction is

$$A + B \rightarrow C$$

A single-stage flash unit separates unreacted A and product C (liquid phase) from reactant B and an impurity D (vapor phase). A small portion of the vapor stream is purged to keep the composition of D from building up to a point where the reaction would be reduced significantly. Figure G.2 indicates that the recycle tank is intended to operate at a high enough pressure to recondense B and D for introduction back into the

CSTR in the liquid phase; a condensor in the recycle line is used for this purpose. It is assumed that a compressor is not required.

The reactor is fitted with a cooling coil for temperature control. A heat exchanger (preheater) is provided to heat the feed stream to the flash unit to ensure that the feed enthalpy is sufficient to provide a complete separation of B and D (vapor) from A and C (liquid). Several dynamic models of the primary process units in this plant are presented in Appendix H.2. For simplicity, the flash unit is modeled as a splitter rather than by a more complex flash model.

We now discuss in detail each step in the design procedure presented in the previous section. Table G.2 lists the controlled and manipulated variables.

G.3.1   Step I: Specify the Control System Design Objectives

A. State the plant production and control objectives, including composition and production rates of all products plus economic objectives.

We assume that plant management and the design group have already developed product quality and production rate specifications, nominal operating conditions, and operating constraints for the plant. The
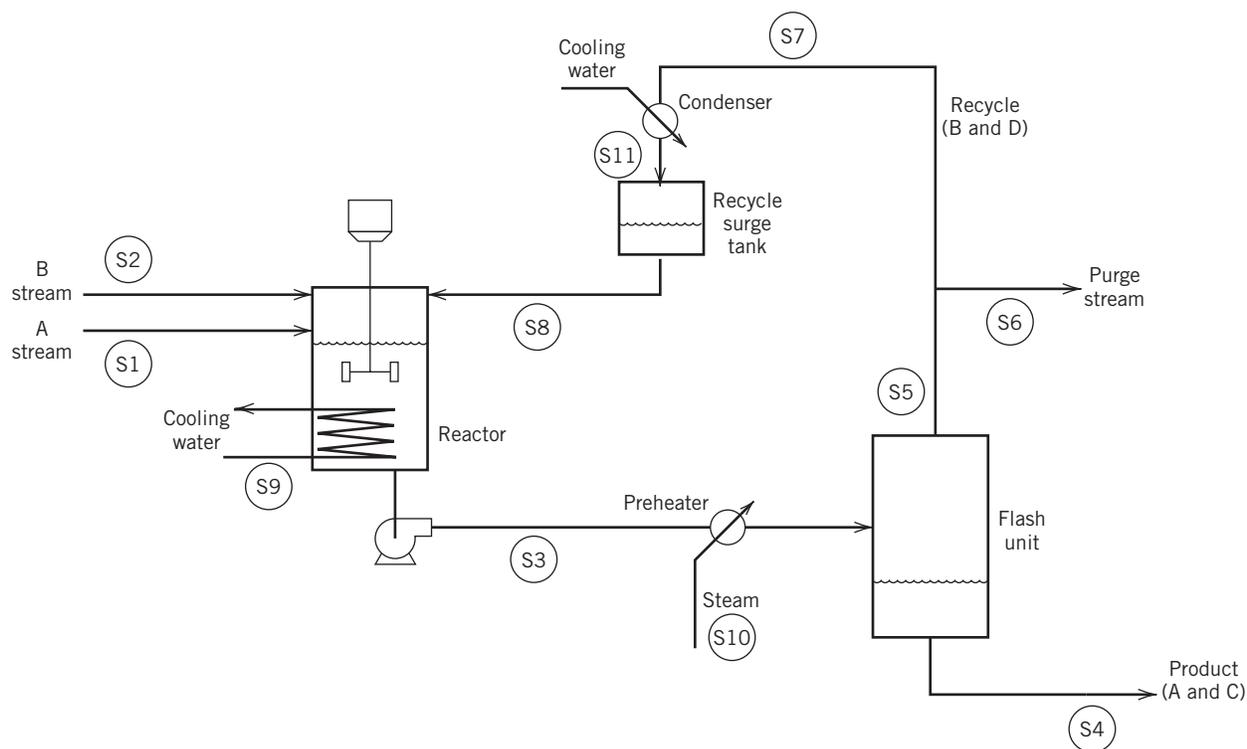
Figure G.2 Schematic diagram for the reactor/flash unit plant showing stream numbers (circles).

Table G.2 Potential Controlled and Measured Variables for the Reactor/Flash Unit Plant[a]

Controlled and Measured Variables	Location/Symbol
Composition, CV	Product stream, x_{4A}
Composition	Reactor effluent, x_{3A}
Composition, CV	Recycle stream, x_{8D}
Flow rate, CV	A feed stream to reactor, w_1
Flow rate, CV	B feed stream to reactor, w_2
Flow rate, CV	Product stream, w_4
Temperature, CV	Reactor, T_R
Temperature, CV	Flash unit feed stream, T_{FF}
Temperature, CV	Recycle tank (condenser exit temperature), T_C
Liquid level, CV	Reactor, H_R
Liquid level, CV	Flash unit, H_F
Liquid level, CV	Recycle tank, H_T
Pressure, CV	Flash unit, P_F
Flow rate	Reactor effluent stream, w_3
Flow rate	Recycle vapor stream, w_7
Flow rate	Purge stream, w_6
Flow rate	Recycle liquid stream, w_8
Temperature	Flash unit, T_F
Pressure	Reactor, P_R
Pressure	Recycle Tank, P_T

[a]Controlled variables are designated by CV and are also measured.

Notes:
i. Compositions in A and B feed streams cannot be measured.
ii. Compositions x_{4A} and x_{8D}, pressures P_R, P_F, and P_T, and temperature T_R must satisfy specific constraints.
iii. Production rate w_4 has to be established via direct flow measurement (not inferred).
iv. Flow rates w_1 and w_2 should be measured and considered for flow control in order to isolate the reactor from upstream pressure disturbances.
v. All vessel inventories, H_R, H_F, and H_T, must be measured and eventually controlled. However, only H_R must be controlled to a set point.
vi. Temperature T_R must be controlled.
vii. Temperature T_F is included to be conservative. Normally, P and T are closely related in an adiabatic flash unit. (For a binary mixture, one measurement is equivalent to the other; also approximately true for a pseudobinary such as this one consisting of four components.)

control objectives are determined so as to meet customer requirements and anticipated sales figures, to reflect plant raw material and operating costs, and to satisfy materials of construction and environmental limitations:

1. The product should contain approximately 99% C; the remaining impurity is A.
2. The desired production rate w_4 to the downstream unit should meet the following specifications:

Nominal value $\pm$ 1% on long-term basis (days); Nominal value $\pm$ 3% on short-term basis (hours)

3. The reactor should be operated with approximately constant conversion as production rate varies within expected limits. Because a suitable value of conversion will depend on the production rate, no specific requirement can be provided. The nominal reactor temperature T_R is specified.

B. Identify process constraints that must be satisfied, including safety, environmental, and quality restrictions.

1. Mass fraction of A in the product stream, x_{4A}, should be less than 0.011 (1.1%), a quality constraint.
2. Mass fraction of D in the recycle liquid stream, x_{8D}, is 0.1 (10%), a value determined by steady-state economic optimization.
3. $P_F^L \leq P_F \leq P_F^H$ (low-level constraint to yield smooth operation: high-level constraint required to meet materials limits).
4. $T_R \leq T_R^H$ and $P_R \leq P_R^H$ (high-level constraints on reactor temperature and pressure imposed by materials limits).
5. All vessel levels (H_R, H_F, and H_S) maintained between high and low limits.

Note: The product must be sent to waste if $x_{4A} >$ 1.1%. On the other hand, maximizing x_{4A} while satisfying the constraint is the optimum economic strategy. Depending on how tightly x_{4A} can be controlled, some nominal value, such as 1%, should be used as the set point.

G.3.2   Step II: Perform a Top-Down Analysis

This step is intended to develop a conceptual design of the plantwide control system. Step II.A is concerned primarily with analysis; in Step II.B, the overall control structure is established in a conceptual form.

A. Identify the process variables, control degrees of freedom, control structure, and options for decomposition.

A.1. Identify the potential controlled variables. The schematic flow diagram in Fig. G.3 shows the most important measurement locations and the process variables. The operating objectives clearly require that two key variables be controlled, x_{4A} and w_4. Composition
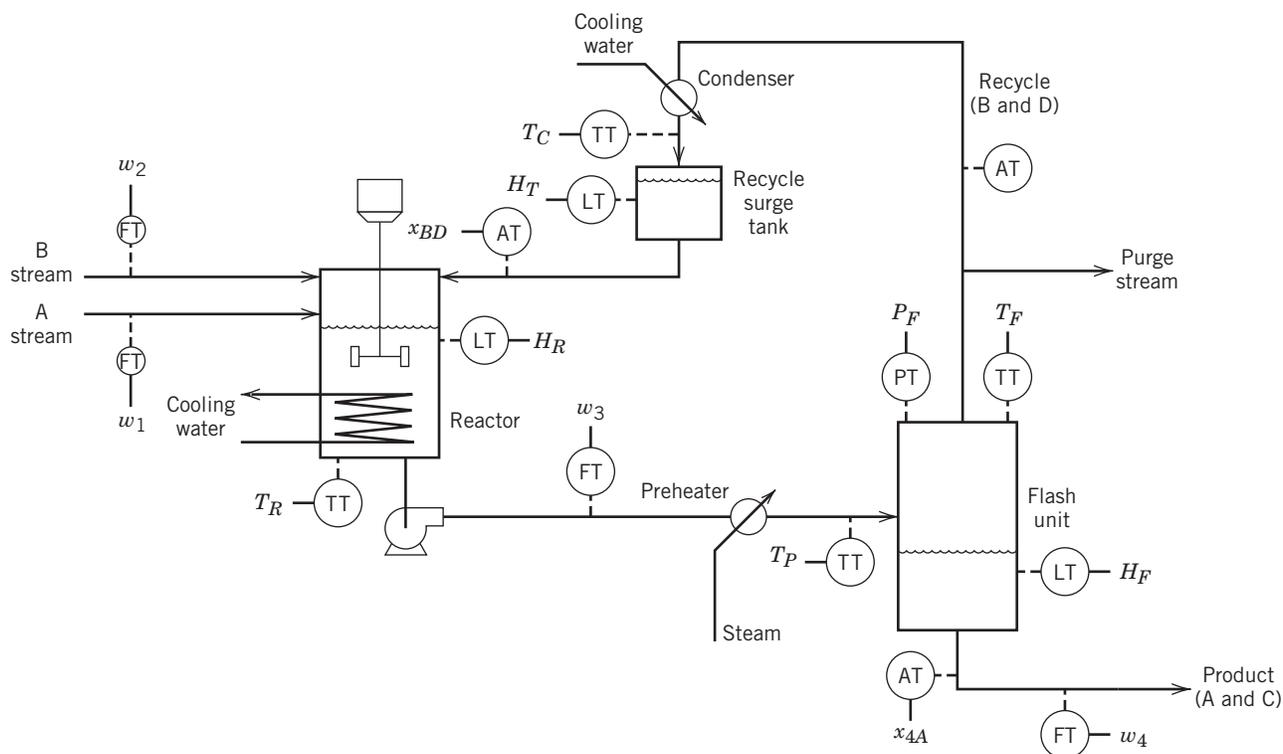
Figure G.3 Instrumentation for the reactor/flash unit plant.

x_{4A} has to be tightly controlled, but w_4 only needs to be tightly controlled on a long-term basis. Whether composition x_{3A} must be controlled explicitly is unclear at this point. However, we assume that it can be maintained approximately constant without using a specific composition controller. Furthermore, the composition of D in the recycle stream, x_{8D} (and consequently x_{8B}), also will likely need to be controlled. If x_{8D} is controlled, operation of the reactor/flash plant probably will not be unduly subject to disturbances (snowball effect). However, these issues need to be checked carefully at the end of the design using simulation.

Table G.2 lists the controlled variables for the plant developed using the specific arguments given above and the general guidelines given in Section 12.2. At this point, each process measurement could potentially be used as a control variable.

A.2. Determine how the controlled variables can be measured or inferred and identify other process variables to be measured. The schematic flow diagram in Fig. G.3 also shows the locations of the most important sensors/transmitters. In addition to measurements for the controlled variables, actual plants are routinely provided with many additional, but less important, measurements. Measurements such as cooling water inlet and outlet temperatures on the reactor cooling coil and heat exchanger steam supply pressure are required to

give the operators a clear picture of the process behavior and its environment. Such information is particularly important during plant start-up, shutdown, and periods when the plant is upset.

The measured variables listed in Table G.2 are based on the problem description, the control system specifications (Step I.A), and the general guidelines of Section 12.2. Included are several potentially useful measurements that are not needed in the control system design.

A.3. Select the potential manipulated variables. Unless a stream is "wild" and cannot be manipulated (such as an exit stream from an upstream unit) or cannot be manipulated independently, its flow rate will be adjusted via a control valve. An example of the latter restriction would be a valve in Stream 5 (Fig. G.2), which cannot be used to manipulate flow rate independently if control valves are installed in both Streams 6 and 7.

General guidelines for selecting manipulated variables are given in Section G.2. All of the manipulated variables in the case study are adjusted by control valves. In general, we try to select manipulated variables that have the most direct influence on the controlled variables—that is, largest sensitivity (gain) and fastest dynamic effects. The primary requirement is to enable pairings in which there is a large, direct influence (high process gain) and, do not exacerbate loop interactions. Structural analysis (RGA or SVA) can provide specific guidance for sensitivity and
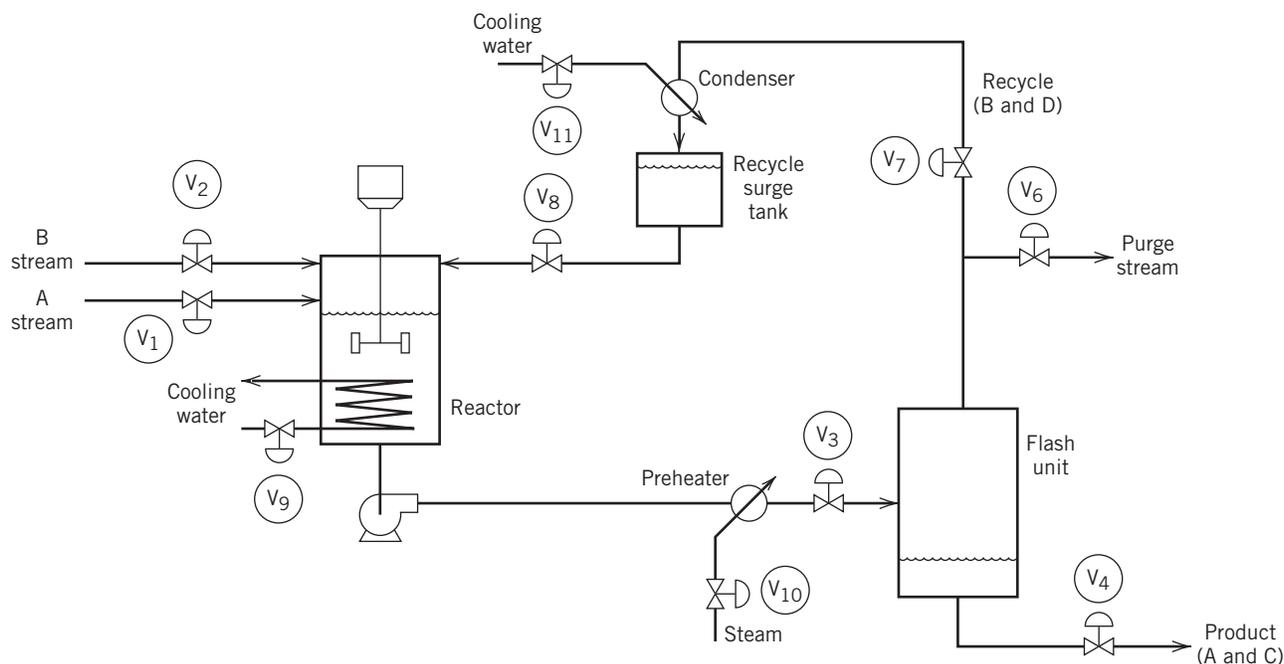
Figure G.4 Control valves for reactor/flash unit plant.

process interactions. A secondary heuristic is to select manipulated variables that are physically close to the controlled variables to take advantage of potentially fast dynamics. Figure G.4 and Table G.3 indicate locations selected for the reactor/flash plant control valves. Again, Stream 5 contains no valve, because its flow rate cannot be manipulated independently if V_6 and V_7 are installed.

Table G.3 Manipulated Variables (and Associated Valves) of the Reactor/Flash Unit Plant

Stream Number/MV	Valve
1. Reactor A feed, w_1	V_1
2. Reactor B feed, w_2	V_2
3. Reactor effluent, w_3	V_3
4. Flash unit liquid product, w_4	V_4
6. Purge, w_6	V_6
7. Recycle vapor, w_7	V_7
8. Recycle liquid, w_8	V_8
9. Reactor cooling water supply, w_9	V_9
10. Flash unit preheater steam supply, w_{10}	V_{10}
11. Condenser cooling water supply, w_{11}	V_{11}

Notes

i. Both feed streams (1 and 2) and the plant product stream (4) are provided with control valves.

ii. Stream 5 contains no valve.

iii. The reactor, flash unit (liquid), and recycle tank effluents, and the purge stream are available for inventory control.

iv. Control valves are required to manipulate cooling water flow rate in the reactor and condenser coils, and steam pressure in the flash unit preheater.

We assume that the feed flow rates can be manipulated, because the specifications do not indicate that these variables are considered to be disturbances.

A.4 Perform a preliminary control degrees of freedom analysis (compare the numbers of manipulated and controlled variables). The number of control degrees of freedom is the number of manipulated variables (10). Recall that a control degree of freedom is allocated each time a manipulated variable is utilized in a control loop, except in cascade control or in other applications where a set point is manipulated instead of a control valve.

From Table G.3, we obtain

Control degrees of freedom = 10

corresponding to the 10 control valves shown in Fig. G.4. Note that N_{FC} (10) is less than the number of controlled variables (12) shown in Table G.2. It might appear to be necessary at this point to identify additional manipulated variables or to omit some of the controlled variables. However, if certain variables do not have to be independently controlled, it is possible to handle this situation by using advanced control methods (cascade control) or partial control (Kothare et al., 2000). This feature will be illustrated in the bottom-up design.

A.5. Identify the source and nature of the significant disturbances that must be mitigated. There are four primary sources of disturbances, three from within the plant itself or its immediate environment: composition

variations in the feed streams, and temperature or pressure variations in the cooling water and steam utility streams. The fourth disturbance is caused by planned changes in production rate:

1. x_D stream feed; random variation.

2. T_W (temperature of cooling water supply to reactor and to recycle condenser); diurnal (24-hour) cycle.

3. P_S (pressure of steam supply to flash unit pre-heater); relatively slowly varying supply pressure as other units load the steam supply header. We discuss in Step III.A.3 what to do in the event this disturbance turns out to be more difficult to handle.

4. Operator-implemented changes in desired production rate w_4.

A.6. Perform structural analysis based on a steady-state model and evaluate the possibilities for decomposition of the control problem.

To simply this step, we assume that the pressure and temperature control loops are essentially decoupled from the plant holdups (integrating modes), the compositions, and the liquid flows. If this assumption is approximately valid, we can analyze a core plant model ("core model") that comprises the reactor, flash unit, and recycle tank—all assumed to operate isothermally and isobarically (see Fig. G.5). Thus, the approximate plant model consists only of material balances but includes the key flows, levels, and compositions. This type of approach, in which temperatures and pressures are assumed to remain constant at their nominal values, was employed by Robinson et al. (2001) in their analysis of a similar plant.

The resulting core model (see Appendix H.2) contains six controlled variables:

Production rate, w_4
Composition of A in the product stream, x_{4A}
Reactor holdup, H_R
Flash unit holdup, H_F
Recycle tank holdup, H_T
Composition of D in the recycle stream, x_{8D}
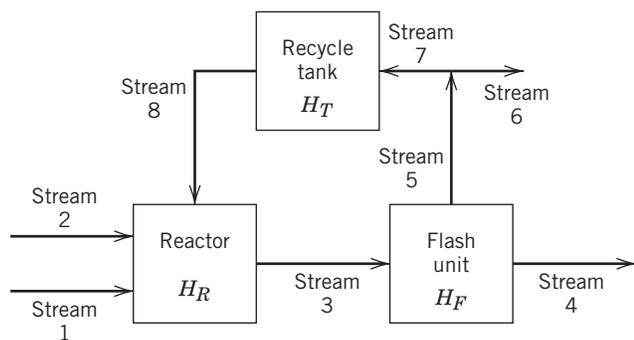
Figure G.5 Process flow diagram for the core model of the plant: the core model consists of reactor, flash unit, and recycle tank, all operated isothermally and isobarically.

where the flow rates and holdups are in mass units and the compositions are mass fractions. Six manipulated variables (all flow rates established by control valves shown in parentheses) are available:

A feed flow rate w_1 (V_1)
B feed flow rate w_2 (V_2)
Reactor exit flow rate w_3 (V_3)
Flash unit liquid flow rate w_4 (V_4)
Purge flow rate w_6 (V_6)
Recycle flow rate w_8 (V_8)

At this point, one could develop a 6×6 RGA that would provide guidance on how the plant might be decomposed for multivariable control and how variables might be paired in a subsequent bottom-up (detailed) design. First, we recognize that the most direct way of controlling the plant production rate w_4 is to use V_4. However, making that choice leads to a problem discussed in Appendix F regarding the design of flow/level controllers for vessels in series. If V_4 is used to control w_4, then only V_3 can be used to control flash unit holdup H_F. Furthermore, there is no easy way to control the reactor holdup H_R, because use of any reactor inlet valve (V_1, V_2, or V_8) to adjust the reactor level can change the molar ratios of reactants. Of course, that problem could be mitigated by ratioing all three valves, but normally this approach is undesirable. Thus, we conclude that it is better to control H_R by V3 and H_F by V4, and to control the production rate in an indirect manner.

These preliminary decisions leave four controlled variables and four manipulated variables that can be analyzed using a 4×4 relative gain array. For the core plant model (mass balance equations only) and values of the operating parameters given in Appendix H, the steady-state gain matrix is

Steady-State Gain Matrix

		w_1	w_2	w_6	w_8
	w_4	1.93	2.34×10^{-2}	0	6.29×10^{-3}
$K=$	x_{8D}	8.46×10^{-4}	-7.97×10^{-4}	0	5.72×10^{-6}
	x_{4A}	2.51×10^{-5}	-1.18×10^{-5}	0	-3.17×10^{-6}
	H_T	-0.93	0.977	-1	-6.29×10^{-3}

$$(G\text{-}1)$$

Note that w_6 only affects H_T. The elements in the H_T row (K_{4j}) consist of rate-of-change coefficients instead of gains, because it is an integrating variable. Woolverton (1980) and Arkun and Downs (1990) showed that, in order to calculate the RGA, the rate-of-change coefficients for an integrating variable can be treated just as if they were gains. Using their approach, we can

obtain the RGA:

Relative Gain Array

$$
\Lambda = \begin{array}{c} \\ w_4 \\ x_{8D} \\ x_{4A} \\ H_T \end{array}
\begin{array}{c} \overset{w_1 \quad\quad w_2 \quad\quad w_6 \quad\quad w_8}{} \\ \left[\begin{array}{cccc} 0.975 & 0.013 & 0 & 0.012 \\ 0 & 0.974 & 0 & 0.026 \\ 0.025 & 0.013 & 0 & 0.962 \\ 0 & 0 & 1 & 0 \end{array} \right] \end{array} \qquad \text{(G-2)}
$$

From the RGA, it is clear that the core plant model is not very interacting; however, it gives little insight into potential decomposition of the full plant. Thus, a control system design developed with a multiloop approach based on a simplified model should be tested using simulation, and eventually with the actual plant, to see how well the simplifying assumptions hold. Note that the RGA and similar analytical methods are intended to be used for initial screening.

Because the degree of interaction is low, there appears to be no compelling reason to employ a multivariable control methodology such as MPC in dealing with the core plant. Thus, it is possible to decompose down to the individual unit and, except for cascade and ratio control applications discussed below, even to the single-loop level.

Those potential pairings exhibiting relative gain elements approximately equal to one serve to guide the detailed bottom-up design that follows. For this simple plant model, with its straightforward reaction kinetics and separator modeled by a splitter rather than a flash model, enough information is already provided at this point to design the control system structure. However, we continue with application of the recommended design procedure to illustrate its application.

B. Establish the overall control structure in conceptual form.

B.1. Identify where the production rate of each product will be measured and controlled.
The production rate w_4 is measured by means of a flow transmitter placed directly in the product line, rather than inferred from a measurement elsewhere in the plant, as is sometimes required. Because the desired variability of the production rate is small ($\pm 1\%$), measuring a related flow rate further upstream (e.g., w_3) could introduce too much variability if the flash unit level controller manipulates product stream flow rate w_4. However, with only two units in the downstream path of this plant (reactor and flash unit), we have already discussed why it is reasonable to manipulate the production rate at an upstream location using a variable that directly influences this flow rate. In principle, either w_1 or w_2 could be manipulated for this purpose, because both reactants are required to make product C. However,

that is true only as long as A and B compositions in the reactor are near the stoichiometric ratio. We know that A is the limiting reactant. Thus, that is why the RGA indicates that only flow rate w_1 has a meaningful effect. Valve V_1 is allocated for this purpose.

Initially, we assume that flow rate w_1 will be maintained using a flow controller whose set point is adjusted manually to hold w_4 within the desired limits. However, what type of control loop to use or how its set point is to be adjusted is uncertain until we develop the detailed bottom-up design. Following the introduction of several additional considerations in the bottom-up design phase, these details can be developed.

> **Result:** Valve V_1 is allocated for control of production rate.
>
> Remaining control degrees of freedom $= 10 - 1 = 9$

B.2. Identify how quality will be measured for each product and how quality will be controlled.
Composition x_{4A} is a key quality variable because it is strictly limited to less than 1.1%. Because the RGA recommendations are unambiguous (Eq. G-2), the recycle stream valve V_8 (flow rate w_8) is chosen as the manipulated variable.

> **Result:** Valve V_8 is used to control x_A.
>
> Remaining control degrees of freedom $= 9 - 1 = 8$

Although one of the secondary control objectives is to keep the reactor exit composition x_{3A} reasonably constant, control of this intermediate variable does not appear difficult enough to require a separate feedback controller.

B.3. Determine how each recycle loop throughput/composition will be controlled.
Because this plant does not appear to be sensitive to disturbances leading to effects such as snowballing, controlling x_{8D} in the recycle loop appears to be sufficient.

B.4. Specify how the constraints will be satisfied.
All of the operating constraints can be addressed by selectors and overrides (Chapters 9 and 15). These include

1. x_{4A}
2. x_{8D}
3. P_F^H and P_F^L
4. T_R^H and P_R^H
5. High and low levels in all three vessels.

Note that Constraint 2 on x_{8D} has been specified by plant designers in advance. If x_{8D} should be changed in response to operating and economic conditions, it could be determined on-line via real-time optimization (Chapter 19).

There is one implicit condition, not part of the design specifications, that needs to be considered in the bottom-up design. Because the flash unit separates a pseudobinary mixture (A/C and B/D), its temperature and pressure cannot be independently specified. Thus, either T_F or P_F, but not both, can be controlled.

B.5. Determine how major disturbances will be handled. The effect of temperature changes in the cooling water supply on reactor temperature or condenser exit temperature presumably can be attenuated by their temperature control loops without upsetting the composition and flow loops. The same is true for the effect of pressure changes in the steam supply header on flash unit temperature (or pressure). All three of these loops will have to be tested carefully after design is complete, through simulations and/or plant tests, to determine whether the assumptions underlying the decision to design the actual plant using a reduced model (the core model) are correct.

Variations in the remaining environmental disturbance x_{2D} require an explicit feedback controller to maintain x_{8D} near its desired value. The purge stream w_6 (the only path by which D leaves the plant) could be manipulated to control x_{8D}. However, RGA results clearly indicate that approach will only work if *all* control loops are closed (see Exercise G.4); but an effective manipulated variable (w_2) is available to deal with this disturbance without imposing such a severe restriction.

B.6. Analyze the energy management scheme and indicate conceptually how it will be controlled. A potential source of energy savings for this process appears to be the energy liberated in the exothermic reaction. That "lost energy" could provide some of the energy required to heat the flash unit feed to flash conditions. However, it is impossible to control the reactor temperature at a desired value and simultaneously heat up its exit stream, the flash unit feed stream, to a higher value. In other words, this energy is *thermodynamically unavailable* (Sandler, 2006).

The flash unit vapor condenser might supply some part of the flash preheater energy requirements, but recovering it with the small approach temperature differences that are available would require too large a heat exchanger. Thus, it appears that the potential for energy savings through heat integration is small, and plant heating/cooling utilities will have to be used.

G.3.3   Step III: Develop a Bottom-Up Design

A. Develop a strategy for regulatory control.

The primary tool for regulatory control is the SISO control loop. However, many applications benefit from the use of (and may even require) advanced methods such as selective control techniques and multivariable control (Chapters 15 and 16).

A.1. Specify how the control system will treat unsafe or abnormal operating conditions and deal with constraints. First identify those variables that potentially pose a safety or environmental hazard (cf. Chapter 12):

P_R subject to high-pressure limit P_R^H: Pressure-operated overrides (see Chapter 15) should be placed on the reactor level controller and on any flow controllers that feed or empty the reactor. To ensure that this hard constraint is never violated, a quick opening valve and rupture disk should be installed in a separate line leading to a flare or blow-down tank.

T_R subject to high-temperature limit T_R^H: A temperature-operated override should be placed on the reactor temperature controller and a fail-open valve used to manipulate the cooling water flow rate.

P_F subject to high-pressure limit P_F^H: Pressure-operated overrides should be placed on the flash unit level controller and on the recycle and purge stream controllers. Again, the contents should be released to a flare or blow-down tank if the hard constraint is reached.

Then identify those variables whose constraints might compromise satisfactory operation of the plant:

P_F subject to a low limit P_F^L: Place an override on the purge stream flow controller to make sure that the recycle tank is not upset by a low-flow situation in the recycle line.

H_R, H_F, H_T subject to high- and low-level limits: Place an override on the level controller for each vessel to open or close the pertinent control valve in the event that an "overflow" or "running dry" state is approached.

A.2. Identify control loops to regulate production rates and inventories. First, we specify control loops for the holdups in the major vessels. This step deals with the plant's integrating modes discussed in the top-down analysis (Step II.A.6). The RGA results (Eq. G-2) indicate that the recycle tank level is controlled best by manipulating the purge stream valve rather than the recycle stream valve:

> **Results:** Reactor exit stream valve (V_3) is used to control H_R.
> Liquid product stream valve (V_4) is used to control H_F.
> Purge stream valve (V_6) is used to control H_T.
>
> Remaining control degrees of freedom $= 8 - 3 = 5$

Note that *averaging control* can be used for H_F and H_T where tight level control is not required to smooth out the effect of disturbances, but not for H_R.

A.3 Identify control loops that will mitigate major disturbances. Variations in x_{2D} will produce deviations in x_{8D} from its desired nominal operating value. The RGA analysis has identified the B feed stream valve (V_2) as the most effective actuator.

Recall that plant designers included the purge stream in order to remove D from the plant. However, the RGA results in Eq. G-2 indicate that manipulating V_6 is not an effective way to control x_{8D}. Instead, V_2 is chosen, based on the RGA analysis:

Result: V_2 is used to control x_{8D}.

Remaining control degrees of freedom $= 5 - 1 = 4$

Upstream pressure variations in the two reactor feed streams (w_1 and w_2) can be attenuated by using a flow controller in each line. However, x_{8D} only needs to be controlled approximately at the desired value of 10%; thus, a flow controller for w_2 appears to be an unnecessary complication. Using a flow controller on w_1 implies that its set point will be adjusted to maintain production rate w_4. Note that an additional control degree of freedom is not required, because control valve V_1 was already allocated in the top-down analysis (Step II.B.1) to adjust production rate.

Results: Flow controller manipulates V_1 to control w_1.
Set point of w_1 controller is adjusted to set production rate w_4.

Remaining control degrees of freedom are unchanged $= 4 - 0 = 4$

Because energy management is simple for this plant, disturbances in energy balances presumably can be handled by single-loop controllers. It is assumed that disturbances to reactor temperature (caused by a varying reaction rate or cooling water temperature changes) can be mitigated by using a reactor temperature control loop. The same is true for the effects of pressure changes at the steam supply header on flash unit pressure or temperature. Because there are explicit constraints on pressure but not on temperature, pressure is chosen. This item is considered more fully in Step IV.D, where the accommodation of constraints is discussed in detail.

Similarly, the effects of temperature variations in the cooling water supply on operation of the condenser can be mitigated by use of a temperature control loop. If disturbances are particularly large in a utility supply, a cascade secondary controller can be employed to control the temperature or pressure of the utility stream at the point it leaves the process, with a primary controller used to maintain the process temperature (Chapter 15). Cascade control, which is applied in Step III.B, is not used here for reasons of simplicity.

Results: V_9 is used to control T_R.
V_{10} is used to control flash unit feed temperature T_{FF}.
V_{11} is used to control condenser exit temperature T_C.

Remaining control degrees of freedom $= 4 - 3 = 1$

One major disturbance remains: the variation in P_F caused by changes in w_6. P_F can be controlled by manipulating valve V_7 in the recycle vapor line.

Result: V_7 is used to control P_F.

Remaining control degrees of freedom $= 1 - 1 = 0$

Note that some designers would choose to operate V_7 fully open and let P_F "float" in order to save pumping costs associated with the pump in the reactor effluent line. We assume here that pressure control is necessary to maintain flash unit pressure constant. Disturbance sensitivity is assumed not to be an issue for this plant, so any need to control a flow rate or composition variable within the recycle loop will be satisfied by controlling x_{8D}.

At this point in the analysis, there are no remaining control degrees of freedom. However, the design of the plantwide control system is by no means complete: advanced control methods that adjust the set points of already specified feedback controllers can be used to make the plant operate better. The objective here is to structure the control system in ways that avoid the need for operator intervention except when absolutely necessary.

B. Examine the potential of applying advanced control strategies.

Advanced control is intended to provide improved performance over traditional single-loop control. It includes such techniques as multivariable, cascade, feedforward (including ratio), and inferential control. As noted already, overrides (Chapter 15) can be particularly helpful in dealing with variable constraints.

Cascaded flow controllers can reduce the effect of any upstream pressure variations or changes in control valve characteristics resulting from nonlinearities or from fouling, as noted in Step II.A.3. Ratio control between w_1 and w_2 can maintain the desired stoichiometric ratio of reactants approximately constant, despite changes in production rate or feed composition. Finally, cascade control can help deal with disturbances introduced by intentional changes in production rate w_4, as is discussed next.

B.1. Evaluate the use of advanced single-loop control strategies, including feedforward, ratio, cascade, and selective control schemes. In reviewing the plant processing and control objectives, a variable that needs

further attention is the production rate w_4. One way of automating w_4 is to measure it and use a cascade controller to adjust the set point of the w_1 controller. An additional degree of freedom is not required to implement the cascade (master) controller, because the set point of the w_1 controller is available. A slow change in the w_{1sp} should meet the long-term production rate requirements and not interfere on a short-term basis with the H_F control loop.

> **Results:** A cascade controller for w_4 is employed to adjust the set point of the w_1 flow controller. Its set point, w_{4sp}, is used to set the desired production rate.
>
> Remaining control degrees of freedom $= 0$ (unchanged)

It is also desirable to speed up the adjustment of w_2 so that the ratio of B to A remains approximately at its correct (stoichiometric) value. A ratio controller (Section 14.2) whose internal ratio is adjusted by a primary composition controller is used to control x_{8D}.

> **Results:** Controller for x_{8D} adjusts the B to A ratio controller set point. The ratio controller manipulates V_2.
>
> Remaining control degrees of freedom $= 0$ (unchanged)

B.2. Employ multivariable control for highly interactive processes. So far we have assumed that a multiloop control approach will be sufficient and that multivariable control will not be necessary. One way to help ensure that this assumption will eventually be validated is to design the individual control loops so they interact as little as possible by careful selection of controlled variables and their pairing with manipulated variables. For example, adjusting the value of w_1 and the ratio of w_2/w_1 (in order to control w_4 and x_{8D}, respectively), instead of directly controlling the two flow rates individually, is one way of physically decoupling the two control loops (see Chapter 16).

Only a dynamic simulation of the controlled plant can determine whether the multiloop control strategy works satisfactorily. If the proposed loops interact too much or fail to achieve the desired control objectives, a more powerful multivariable approach such as MPC may be required.

C. Evaluate the economic benefit of real-time optimization.

The major process variables that are candidates for real-time optimization are T_R and H_R (to optimize operation of the reaction process), P_F (to optimize the

separation process), and x_{8D} (here assumed to be constant at 10%). A steady-state process model must be available to carry out such calculations; see Chapter 19 for more details.

G.3.4 Step IV: Validate the Proposed Control Structure

A. Perform a final control degrees of freedom analysis. Check the allocation of the N_{FC} degrees of freedom.

Because we have kept track of the control degrees of freedom as individual control loops have been proposed (control valves have been allocated), it is clear that we have not attempted to use too many degrees of freedom. Nor have any possibilities been neglected to obtain better control through the use of additional control loops that utilize already allocated degrees of freedom.

B. Check control of individual process units.

The next step is to make sure that no physically unrealizable control schemes have been proposed—for example, to attempt to control all of the component concentrations in a stream plus its total flow rate. Even with controllers in place, there must be some way for each species to leave the plant. For example, when the purge line is closed, there is no way for component D to leave the recycle path; thus, constraint handling methods associated with the control system cannot close V_6 for a significant period of time.

Finally, if steady-state simulation software is available, this is a good place to check anticipated concentrations and flows throughout the plant with the controllers implemented. Failure of the simulator to converge to the design operating conditions may be an indication that something is fundamentally wrong.

In designing the plantwide control system, we have essentially dealt with control of individual process units, with one exception. P_F must be maintained high enough to provide a sufficient pressure drop across V_7. If not, the recycle vapor stream valve may have to be operated fully open, and P_F controlled by adjusting V_{11}. If V_7 is always open, one control degree of freedom will be lost.

C. Check the effect of constraints and disturbances on manipulated and controlled variables.

All of the design constraints have been addressed by the proposed feedback control loops:

1. x_{4A} *constraint*: Manipulate w_1.
2. x_{8D} *constraint*: Manipulate w_2 by means of the $w_2 : w_1$ ratio.

3. P_F^H *and* P_F^L: Manipulate w_7, with an override probably required on the flash unit feed temperature to handle the situation where V_7 either opens or closes fully.

4. T_R^H: Manipulate w_9.

5. T_R^H: Provide an override on w_6.

In the top-down design phase, we identified an additional, implicit constraint to be addressed that is not part of the design specifications. Specifically, the flash unit processes a pseudobinary mixture (A/C and B/D), and thus, its temperature and pressure cannot be specified independently. This design issue has been handled by controlling the flash unit feed temperature (flash unit preheater exit temperature) T_{FF} rather than T_F itself.

D. Simulate control system performance for a wide range of conditions.

If a dynamic process simulator is available, it should be used to evaluate the proposed plantwide control strategy and to determine recommended initial controller settings. It also should be used to evaluate the assumptions behind the core model analysis—namely, that the pressure and thermal control loops can be considered to be substantially decoupled from the flow/level/composition loops.

We present closed-loop simulation results for the core model (the reduced holdup form) and controller settings given in Appendix H.2, showing its responses to two important process disturbances. For simplicity, the reactor holdup (level) is assumed to be controlled perfectly. In addition, holdup in the flash unit is assumed to be zero, because it normally will be quite small compared to the reactor holdup. The base case control

structure consists of 4 of the 12 control loops given in Table G.4: composition loop 7 ($x_{4A} - w_8$), modified composition loop 12 ($x_{8D} - w_2$—that is, manipulating V_2 directly), level loop 10 ($H_T - w_6$), and flow loop 11 ($w_4 - w_1$, primary loop only). Note that for the core model, flow rates are manipulated directly so the secondary controller for w_1 (flow loop 1) is not required.

Figure G.6 shows how three key controlled variables, w_4, x_{4A}, and x_{8D}, react to a $+0.03$ change in x_{2D}. In this simulation, the B-to-A ratio controller (ratio loop 2) was not implemented, because it does not affect the responses when there is no change in w_1. Note that all three controlled variables exhibit only small deviations from their set points as a result of the tight controller tuning that can be used in these loops. Recall that these controller pairings were chosen because of the direct influence of the manipulated variable in each loop and the relatively low degree of process interactions indicated by the RGA in Eq. G-2.

Figure G.7 shows the responses for x_{4A} and x_{8D} to a production rate change in w_4. A set-point change of $+100$ kg/h was made to the w_4 controller, first without ratio control of w_2. When ratio control is not used, larger deviations in x_{8D} occur as a result of the induced changes in A feed flow rate (w_1), with no corresponding immediate change in B feed flow rate w_2.

A form of ratio control was implemented in the second test by including ratio loop 2 ($w_2 - w_1$) with an initial desired ratio of 1.09, while retaining the four base case controllers (loops 7, 10, 11, and 12 in Table G.4). The addition of ratio control results in essentially no deviation in x_{8D}, and the beneficial effect of maintaining the B to A ratio during production rate changes is seen to affect only an unimportant recycle stream variable. Control of production rate w_4 and product quality x_{4A} is not significantly improved.

Table G.4 Proposed Control System Structure (Control Loops) for the Reactor/Flash Unit Plant

Loop Number	Controller Type	Controlled Variable	Manipulated Variable/Valve
1	Cascade (Secondary)	A stream flow rate, w_1	A feed stream, V_1
2	Ratio	B stream flow rate, w_2	B feed stream, V_2
3	Feedback	Reactor temperature, T_R	Cooling water, V_9
4	Feedback	Reactor level, H_R	Reactor effluent, V_3
5	Feedback	Flash unit feed temperature, T_{FF}	Steam supply, V_{10}
6	Feedback	Flash unit liquid level, H_F	Plant product, V_4
7	Feedback	A composition in product, x_{4A}	Recycle liquid stream V_8
8	Feedback	Flash unit pressure, P_F	Recycle vapor stream, V_7
9	Feedback	Condenser exit temperature, T_C	Cooling water, V_{11}
10	Feedback	Recycle surge tank level, H_T	Purge line, V_6
11	Cascade (Primary)	Plant production rate, w_4	Set point for w_4 (FC 1)
12	Feedback	D composition in recycle, x_{8D}	Ratio $w_2 : w_1$ (RC 2)
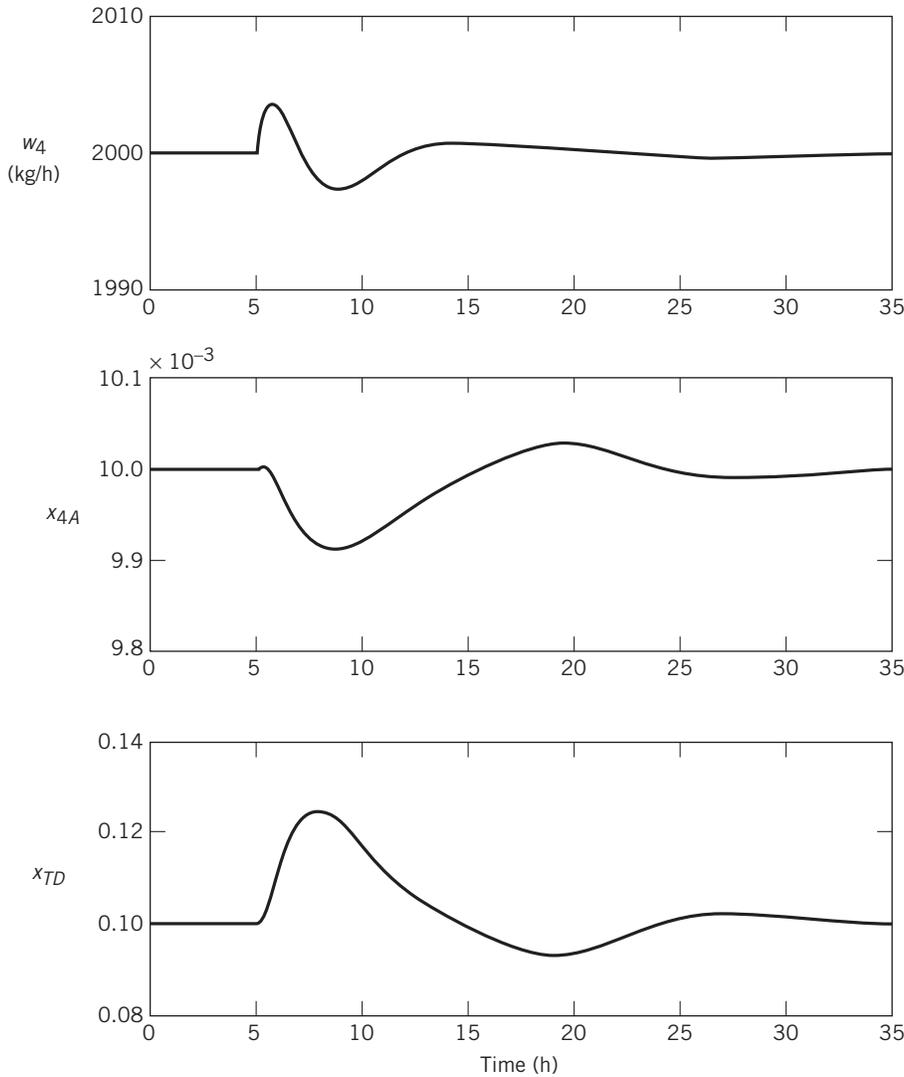
Figure G.6 Closed-loop responses of the reactor/flash unit core model: $+0.03$ change in x_{2D}, without ratio control. (Controller settings are in Table H.7.)

G.3.5   Summary and Interpretation of Control Structure for the Reactor/Flash Unit Plant

The proposed plantwide control system determined with the guidance of the 4×4 RGA-recommended pairings is summarized in Fig. G.8 and Table G.4. This case study represents one hypothetical plant and may give a misleading picture as to how the recommended control system design procedures lead to a particular structure. In general, design procedures are iterative, and thus they can lead to many alternative designs.

An example can be given of just how much the plantwide control system design changes if a slightly different set of assumptions is made. What if the short-term operating constraint on production rate was tighter than specified above ($\pm 1\%$ instead of $\pm 3\%$), or the plant involved more than just three process units? In such a situation, the decision to control the production rate by a cascade loop that extends back to the A feed stream flow rate (w_1) may not be practical. The intervening dynamic lags within the master loop might then preclude its holding the required long-term tolerances. In this case, one alternative would be to control the production rate directly (via a flow controller on w_4) and to employ "upstream" control of H_R and H_F with the related complexities.

A control structure obtained using the hierarchical procedures in the previous section can normally be expected to work reasonably well. However, the only valid test of that conjecture is actually to perform simulations or plant tests after individual controllers have been tuned. In that way, one can determine just how well the controlled system deals with disturbances, production rate changes, and so on. For our purposes, we have focused initially on the core process units in the plant (reactor, flash unit, and recycle tank) to determine how well a design likely would work if it were developed using heuristics, strongly guided by simplified structural analysis. Other credible alternatives are possible. Which of the many alternatives are acceptable
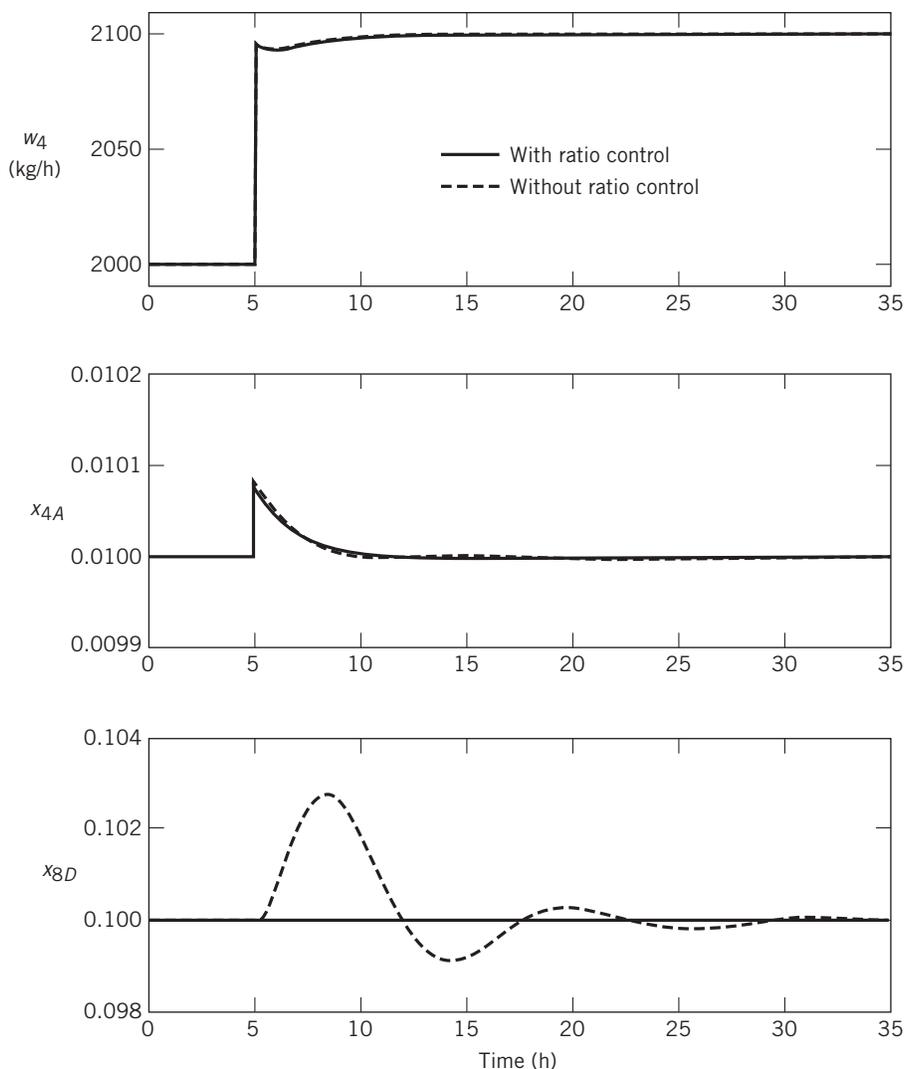
Figure G.7 Closed-loop responses of the reactor/flash unit core model: +100 kg/h step change in w_{4sp}, with and without ratio control. (Controller settings are in Table H.8.)

and which one is "best" in some sense can only be explored via simulation of the full plant model. An extended design including simulation of the full plant is left for the reader.

In making these comparisons, we developed the core model, a level/flow/composition model that neglects the effect of thermal (temperature) and pressure dynamics. For this plant, with only one recycle stream and no heat integration, the assumption is that the temperature and pressure control loops are largely isolated and noninteracting. This assumption has to be tested for accuracy via simulation.

A number of plant-scale control studies have been published. Luyben (2002) presented a series of case studies using dynamic simulation. Downs and Vogel (1993) documented a Tennessee Eastman Company challenge problem that has received considerable subsequent attention from control researchers. Larsson and Skogestad (2000) cited many efforts to deal with the Tennessee Eastman problem and supported the conclusions that alternative control system designs, while sharing certain common features, were almost always highly idiosyncratic. Different plant control engineers or researchers will propose very different control system structures depending on the specific background they bring to the task, the specific design methodology employed, and the simplifying assumptions made.

G.4   EFFECT OF CONTROL STRUCTURE ON CLOSED-LOOP PERFORMANCE

In developing the control structure of the reactor/flash plant, an RGA analysis of the 4×4 core plant was used for guidance. In the final design steps of the case study, we introduced ratio control to maintain the feed stream flow rates at a B:A ratio, R, that is adjusted by the D-composition controller.
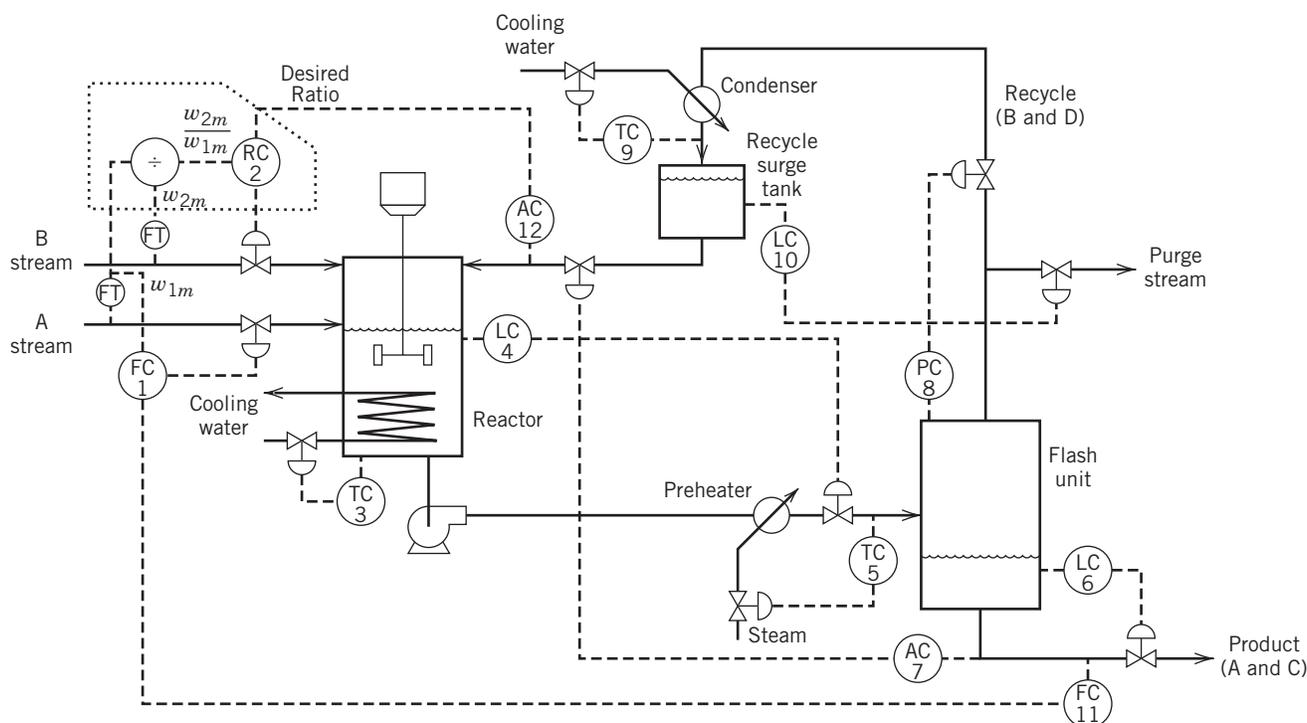
Figure G.8 Proposed decentralized (multiple-loop) control system specification for the full reactor/flash unit plant.

The RGA analysis was performed again, first replacing w_2 by R, the adjusted ratio. Linearizing the core model leads to slightly improved RGA values; in particular, the interaction measure relating w_4 to changes in w_1 becomes slightly higher (0.988 vs. 0.975). Improvement was seen in the dynamic responses of x_{8D}, whose oscillations were seen to be eliminated in Fig. G.7 with ratio control, compared to the responses without ratio control and to those in Fig. G.6.

Based on these results, it appears that the reactor may operate better if the ratio of combined B (feed plus recycle) to the A feed were controlled instead of just the feed stream ratio. To analyze this case, we let R^* be the ratio of combined B-stream flow rate ($w_2 + w_8$) to the A-stream flow rate (w_1). Recalculating the relative gain array yields a surprising result: two of the recommended control loops now become highly interacting (λ values of ~ 3.6) as given in Eq. G-3. This undesired result can be confirmed by simulation.

Relative Gain Array Using a "Combined B" Ratio

$$\Lambda = \begin{array}{c} \\ w_4 \\ x_{8D} \\ x_{4A} \\ H_T \end{array} \begin{array}{|cccc} w_1 & R^* & w_6 & w_8 \\ \hline 0.998 & 0.035 & 0 & -0.034 \\ 0 & -2.66 & 0 & 3.66 \\ 0.002 & 3.62 & 0 & -2.63 \\ 0 & 0 & 1 & 0 \end{array} \qquad \text{(G-3)}$$

Apparently, an important point has been overlooked—namely, that the recycle stream consists of the same flow of B + D that leaves the reactor. Because this material simply recirculates and does not participate in the reaction process under steady-state operating conditions, it should be ignored in applying feedforward control of the ratio. Often, a recombination of variables can lead to a less coupled system, as shown in Chapter 16, and, again, with the addition of simple ratio control between the B and A feed streams in the case study. That is not the situation for the combined B flow rate.

This type of control structure issue can also arise quite naturally through a particular physical feature of the steady-state plant design. Suppose that the reactor is not piped as shown before (Fig. G.2, with a separate recycle inlet port in the reactor) but with the two B streams piped together and entering the reactor through a common port. This design might be chosen to reduce reactor fabrication costs by eliminating an unneeded port, a potentially significant savings for some materials of construction. Faced with this steady-state plant design feature, the control system designer might fall into the trap of believing that a flow transmitter placed to measure the combined B flow rate would be suitable. In fact, as was shown above, the flow transmitter should be placed before the recycle stream inlet so as to measure only w_2.

SUMMARY

It is exciting and yet intimidating for an engineer to be given a steady-state design proposal and some general ideas about how the new plant is to operate, and then be asked to specify the complex system of controllers, safety interlocks, operator interfaces, hardware, and software that comprise a modern control system. In this chapter, we have presented a general procedure for designing plantwide control systems. The steps in the proposed hierarchical design procedure provided here are by no means unique or complete. However, it is important to use an organized approach in the design of a plantwide control strategy, regardless of which design procedure is chosen. Choosing controlled and manipulated variables and pairing them in an ad hoc fashion without a coherent design procedure can lead to serious problems.

Throughout this book, we have emphasized that both qualitative and quantitative process information should be utilized in designing and evaluating control systems. Intelligent use of process models and simulation tools is required to develop a successful design. In addition, difficulties faced by the control system designer often can be mitigated or even eliminated by timely communication with the process design group.

REFERENCES

Arkun, Y., and J. J. Downs, A General Method to Calculate Input-Output Gains and the Relative Gain Array for Integrating Processes, *Computers Chem. Engrg.*, **14**, 1101 (1990).

Buckley, P. S., *Techniques of Process Control*, Wiley, New York, 1964.

Downs, J. J., Distillation Control Design in a Plantwide Control Environment, in *Practical Distillation Column Control*, W. L. Luyben (Ed.), Wiley, New York, 1992, Chapter 20.

Downs, J. J., and E. F. Vogel, A Plantwide Industrial Process Control Problem, *Compute. Chem. Engrg.*, **17**, 245 (1993).

Kothare, M. V., R. Shinnar, I. Rinard, and M. Morari, On Defining the Partial Control Problem: Concepts and Examples, *AIChE J.*, **46**, 2456 (2000).

Larsson, T., and S. Skogestad, Plantwide Control: A Review and a New Design Procedure, *Model. Ident. and Control*, **21**, 209 (2000).

Luyben, W. L., *Plantwide Dynamic Simulators in Chemical Processing and Control*, Marcel Dekker, New York, 2002.

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

Robinson, D., R. Chen, T. J. McAvoy, and P. D. Schnelle, An Optimal Control Based Approach to Designing Plantwide Control System Architectures, *J. Process Control*, **11**, 223 (2001).

Sandler, S. I., *Chemical and Engineering Thermodynamics*, 4th ed., Wiley, New York, 2006.

Skogestad, S., Plantwide Control: Toward a Systematic Procedure, ESCAPE 12 Symposium, The Hague, Netherlands (June 2002).

Stephanopoulos, G., and C. Ng, Perspectives on the Synthesis of Plantwide Control Structures, *J. Process Control*, **10**, 97 (2000).

Woolverton, P. F., How to Use Relative Gain Analysis in Systems with Integrating Variables, *InTech*, **27**(9), 63 (1980).

EXERCISES

G.1 In Appendix H.2 three versions of the core model of the reactor/flash unit plant are developed. One is a "full-composition model" (Eqs. H-9 through H-31) that provides the relations needed to calculate every stream variable and every vessel holdup in the plant design. The second model (Eqs. H-33 through H-40) is a reduced-composition model, obtained from the full model by elimination of all variables and equations not needed to implement the control loops in this chapter. Thus only the necessary manipulated and disturbance variables, the dependent variables in the differential equations (predominantly reactor and recycle tank compositions), and the controlled (output) variables remain in the second model. The third model (Eqs. H-47 through H-55) is a reduced version of the original model equations in which component mass holdups have been used instead of vessel concentrations as the dependent variables.

Note that all the numerical and simulation results in this chapter were obtained using the third model.

(a) Provide a degrees of freedom analysis for each of the three models. Identify all variables, and list all of the required equations by number. Specify the parameters required for each model.

(b) What are the advantages and disadvantages of each model form?

(c) Implement the full concentration model and either of the reduced models using Simulink. Investigate the dynamic nature of the recycle plant and compare the responses of the two uncontrolled plants using changes in one or more disturbances.

G.2 The gain matrix in Eq. G-1 for the plant in this chapter was obtained using analytical methods (Mathematica: see www.mathematica.com) with the reduced holdup model. An alternative way to evaluate the plant interactions is to find the gain matrix and the relative gain array (Chapter 16) using a Simulink model by making small step changes in each input, and then determining the steady-state output changes in order to estimate the gains. Evaluate the relative gain array of this plant using the alternative approach along with a simulation of the full concentration model. Compare your results with the analytical results given in Eq. G-2.

G.3 Using the control loops in Table H.7 and H.8 with the full-composition model created in Exercise G.1, evaluate the use of ratio control in Loop 2 of Table G.4. In particular, indicate why ratio control should improve

the plant performance for production rate (w_4 set point) changes, but not for disturbance (x_{2D}) changes. Your responses should be similar to those in Figures G.6 and G.7.

G.4 A purge stream often is included in recycle plants, such as the reactor/flash plant discussed in this chapter, to keep the concentration of a contaminant from building up within the plant. Thus, one might conclude that the best way to control the concentration of the contaminant below some acceptable level would be to manipulate the purge stream flow rate—i.e., w_6. Examine whether such an approach will work well here, using the following approach:

(a) Modify the Simulink program of the reactor/flash unit core model so that it includes an x_{8D}-w_6 loop and an H_T-R loop along with the remaining two loops summarized in Table H.8. Tune each of the new controllers so as to obtain a low level of interaction with the other control loops.

(b) What features in the RGA analysis given in Eq. G-2 or in the discussion in the first two paragraphs of Section G.4 explain the response results for your alternative control structure, when compared to those shown in Figure G.7?

(c) Can you give a physical explanation for the observed process gain between x_{8D} and w_6?

(d) How would you expect the plant to respond if one of the control loops was inoperable—e.g., as a result of a sensor failure? Remove the x_{4A}-w_8 control loop and compare the structure presented in this problem with the RGA suggested pairing for a +5% step change in production rate.

(e) How well do these two control structures (with all loops closed) handle a larger set-point change (10%) in the production rate, w_4? In particular, what modifications (if any) must be made to handle a set-point change of this magnitude? Can a 20% change in w_4 be accommodated? Why, or why not?

G.5 Using a Simulink representation of the reactor/flash plant, add an additional feedback control loop for reactor level H_R (assumed in this chapter to be perfectly controlled) by the following steps:

(a) Modify the gain matrix (Eq. G-1) to incorporate both a new manipulated variable w_3 and a new controlled variable H_R. What does the RGA indicate about the pairing of controlled and manipulated variables for this situation?

(b) Use RGA-recommended pairings, or any others that are appropriate, to control the plant (including H_R). Can you achieve essentially the same responses as shown in Figure G.7 while controlling reactor level with a PI controller? Using P-only control of level? Explain your results, and discuss whether it is important to control reactor level exactly at the set point.

G.6 If the flash unit in the example plant operates as an ideal splitter but with a non-negligible liquid holdup (e.g., 1,000 kg), what would be the effect on the response of composition loop 12 in Table H.8 for a change in x_{2D}? On the response of composition loop 7? Simulate the modified plant and give logical arguments why one would, or would not, expect a difference.

G.7 Design an MPC controller for the reactor/flash unit plant and test it using a simulation of the linearized model of this plant. For purposes of this exercise, first design and implement a PI controller for the reactor level using the reactor outflow rate w_3 as manipulated variable (see Exercise G.5). Also, include a flash unit holdup of 500 kg and implement a PI controller for the liquid level with the flash unit outflow rate w_4 as manipulated variable (see Exercise G.6).

The following manipulated and controlled variables are to be used in the 4 × 4 MPC:

Manipulated Variables	Controlled Variables
A feed stream flow (w_1)	Production rate (w_4)
B feed stream flow (w_2)	Composition of A in the product stream (w_{4A})
Purge stream flow (w_6)	Recycle tank holdup (H_T)
Recycle line flow (w_8)	Composition of D in recycle stream (x_{8D})

Your design must meet the following control objectives:

(i) The product should contain approximately 99% C; the remaining impurity is A.

(ii) The desired production rate of product w_4 to the downstream unit should meet the following specifications: nominal value ±1% on long-term basis (days), nominal value ±3% on short-term basis (hours).

(iii) The reactor should be operated with approximately constant conversion (unspecified) as production rate varies within expected limits. The nominal reactor temperature T_R is fixed.

(iv) Important quality constraint: mass fraction of A in the product stream (x_{4A}) must be less than 0.011.

(v) For the purposes of this design, the only manipulated variable constraint is that flows are required to be positive.

Once the controller has been designed, evaluate its performance for the following set-point and disturbance sequences (each one separately):

(a) Disturbance response to a +0.03 change in x_{2D}

(b) Set-point response to a 5% change in the production rate (w_4)

(c) Same as (a), but with the rate constant k increased by 20%

(d) Set-point response to a 20% change in the production rate (w_4)

Hints: Use the MATLAB MPC Toolbox, if desired, for this exercise. Two commands are used to produce a linear model of the plant in the representation needed for controller design. First, the *dlinmod* command obtains a state-space representation (**A, B, C, D**). To use this command, be sure that the Simulink diagram is drawn so that the process manipulated inputs and disturbances correspond to "in ports" on the top level of the Simulink flow sheet; similarly, the outputs must correspond to "out ports." Then the *ss2mod* command produces a model in MPC *mod* format, specifying inputs that are manipulated variables, measured disturbances, and unmeasured disturbances. The *scmpc* command simulates control of the linearized plant with the MPC controller.