Process Dynamics and Control

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Chapter 1

Introduction to Process Control

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In recent years the performance requirements for process plants have become increasingly difficult to satisfy. Stronger competition, tougher environmental and safety regulations, and rapidly changing economic conditions have been key factors in tightening product quality specifications. A further complication is that modern plants have become more difficult to operate because of the trend toward complex and highly integrated processes. For such plants, it is difficult to prevent disturbances from propagating from one unit to other interconnected units.

In view of the increased emphasis placed on safe, efficient plant operation, it is only natural that the subject of *process control* has become increasingly important in recent years. Without computer-based process control systems it would be impossible to operate modern plants safely and profitably while satisfying product quality and environmental requirements. Thus, it is important for chemical engineers to have an understanding of both the theory and practice of process control.

The two main subjects of this book are *process dynamics* and *process control.* The term *process dynamics* refers to unsteady-state (or transient) process behavior. By contrast, most of the chemical engineering curricula emphasize steady-state and equilibrium conditions in such courses as material and energy balances, thermodynamics, and transport phenomena. But process dynamics are also very important. Transient operation occurs during important situations such as start-ups and shutdowns, unusual process disturbances, and planned transitions from one product grade to another. Consequently, the first part of this book is concerned with process dynamics.

The primary objective of process control is to maintain a process at the desired operating conditions, safely and efficiently, while satisfying environmental and product quality requirements. The subject of process control is concerned with how to achieve these goals. In large-scale, integrated processing plants such as oil refineries or ethylene plants, thousands of process variables such as compositions, temperatures, and pressures are measured and must be controlled. Fortunately, large numbers of process variables (mainly flow rates) can usually be manipulated for this purpose. Feedback control systems compare measurements with their desired values and then adjust the manipulated variables accordingly.

As an introduction to the subject, we consider representative process control problems in several industries.

1.1 REPRESENTATIVE PROCESS CONTROL PROBLEMS

The foundation of process control is *process understanding*. Thus, we begin this section with a basic question–What is a process? For our purposes, a brief definition is appropriate:

> ***Process:*** *The conversion of feed materials to products using chemical and physical operations. In practice, the term* process *tends to be used for both the processing operation and the processing equipment.*

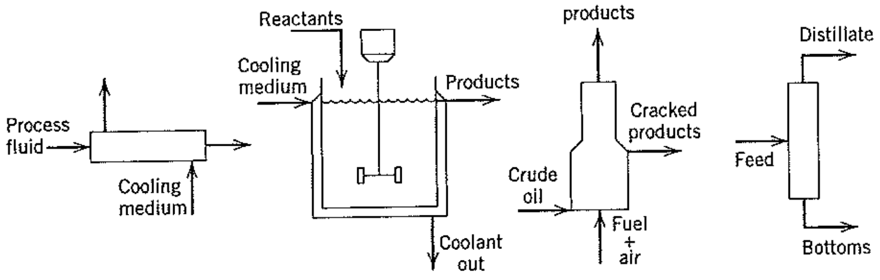
Note that this definition applies to three types of common processes: continuous, batch, and semibatch. Next, we consider representative processes and briefly summarize key control issues.

1.1.1 Continuous Processes

Four continuous processes are shown schematically in Fig. 1.1:

(a) ***Tubular heat exchanger.*** A process fluid on the tube side is cooled by cooling water on the shell side. Typically, the exit temperature of the process fluid is controlled by manipulating the cooling water flow rate. Variations in the inlet temperatures and the process fluid flow rate affect the heat exchanger operation. Consequently, these variables are considered to be disturbance variables.

(b) ***Continuous stirred-tank reactor (CSTR).*** If the reaction is highly exothermic, it is necessary to control the reactor temperature by manipulating the flow rate of coolant in a jacket or cooling coil. The feed conditions (composition, flow rate, and temperature) can be manipulated variables or disturbance variables.



(*a*) Heat exchanger
(*b*) Chemical reactor
(*c*) Cracking furnace
(*d*) Distillation column

Figure 1.1 Some typical continuous processes.

(c) ***Thermal cracking furnace.*** Crude oil is broken down ("cracked") into a number of lighter petroleum fractions by the heat transferred from a burning fuel/air mixture. The furnace temperature and amount of excess air in the flue gas can be controlled by manipulating the fuel flow rate and the fuel/air ratio. The crude oil composition and the heating quality of the fuel are common disturbance variables.

(d) ***Multicomponent distillation column.*** Many different control objectives can be formulated for distillation columns. For example, the distillate composition can be controlled by adjusting the reflux flow rate or the distillate flow rate. If the composition cannot be measured on-line, a tray temperature near the top of the column can be controlled instead. If the feed stream is supplied by an upstream process, the feed conditions will be disturbance variables.

For each of these four examples, the process control problem has been characterized by identifying three important types of process variables.

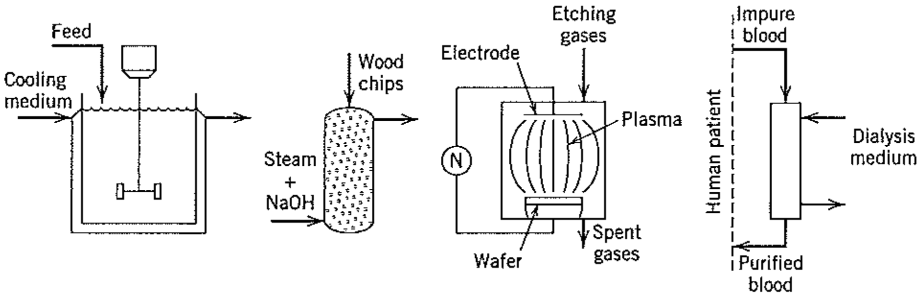
- *Controlled variables (CVs):* The process variables that are controlled. The desired value of a controlled variable is referred to as its *set point.*
- *Manipulated variables (MVs):* The process variables that can be adjusted in order to keep the controlled variables at or near their set points. Typically, the manipulated variables are flow rates.
- *Disturbance variables (DVs):* Process variables that affect the controlled variables but cannot be manipulated. Disturbances generally are related to changes in the operating environment of the process, for example, its feed conditions or ambient temperature. Some disturbance variables can be measured on-line, but many cannot such as the crude oil composition for Process (c), a thermal cracking furnace.

The specification of CVs, MVs, and DVs is a critical step in developing a control system. The selections should be based on process knowledge, experience, and control objectives.

1.1.2 Batch and Semi-Batch Processes

Batch and semi-batch processes are used in many process industries, including microelectronics, pharmaceuticals, specialty chemicals, and fermentation. Batch and semi-batch processes provide needed flexibility for multiproduct plants, especially when products change frequently and production quantities are small. Figure 1.2 shows four representative batch and semi-batch processes:

(e) ***Batch or semi-batch reactor.*** An initial charge of reactants is brought up to reaction conditions, and the reactions are allowed to proceed for a specified period of time or until a specified conversion is obtained. Batch and semi-batch reactors are used routinely in specialty chemical plants, polymerization plants (where a reaction byproduct typically is removed during the reaction), and in pharmaceutical and other bioprocessing facilities (where a feed stream, e.g., glucose, is fed into the reactor during a portion of the cycle to feed a living organism, such as a yeast or protein). Typically, the reactor temperature is controlled by manipulating a coolant



(*e*) Semi-batch reactor (*f*) Wood chip digester (*g*) Plasma etcher (*h*) Kidney dialysis unit

Figure 1.2 Some typical processes whose operation is noncontinuous.

flow rate. The end-point (final) concentration of the batch can be controlled by adjusting the desired temperature, the flow of reactants (for semi-batch operation), or the cycle time.

(f) ***Batch digester in a pulp mill.*** Both continuous and semi-batch digesters are used in paper manufacturing to break down wood chips in order to extract the cellulosic fibers. The end point of the chemical reaction is indicated by the kappa number, a measure of lignin content. It is controlled to a desired value by adjusting the digester temperature, pressure, and/or cycle time.

(g) ***Plasma etcher in a semiconductor processing.*** A single wafer containing hundreds of printed circuits is subjected to a mixture of etching gases under conditions suitable to establish and maintain a plasma (a high voltage applied at high temperature and extremely low pressure). The unwanted material on a layer of a microelectronics circuit is selectively removed by chemical reactions. The temperature, pressure, and flow rates of etching gases to the reactor are controlled by adjusting electrical heaters and control valves.

(h) ***Kidney dialysis unit.*** This medical equipment is used to remove waste products from the blood of human patients whose own kidneys are failing or have failed. The blood flow rate is maintained by a pump, and "ambient conditions," such as temperature in the unit, are controlled by adjusting a flow rate. The dialysis is continued long enough to reduce waste concentrations to acceptable levels.

Next, we consider an illustrative example in more detail.

1.2 ILLUSTRATIVE EXAMPLE—A BLENDING PROCESS

A simple blending process is used to introduce some important issues in control system design. Blending operations are commonly used in many industries to ensure that final products meet customer specifications.

A continuous, stirred-tank blending system is shown in Fig. 1.3. The control objective is to blend the two inlet streams to produce an outlet stream that has the desired composition. Stream 1 is a mixture of two chemical species, A and B. We assume that its mass flow rate w_1 is constant, but the mass fraction of A, x_1, varies with time. Stream 2 consists of pure A and thus $x_2 = 1$. The mass flow rate of Stream 2, w_2, can be manipulated using a control valve. The mass fraction of A in the exit stream is denoted by x and the desired value (set point) by x_{sp}. Thus for this control problem, the controlled variable is x, the manipulated variable is w_2, and the disturbance variable is x_1.

Next we consider two questions.

Design Question. *If the nominal value of* x_1 *is* $\bar{x}_1$, *what nominal flow rate* $\bar{w}_2$ *is required to produce the desired outlet concentration,* x_{sp}?

To answer this question, we consider the steady-state material balances:

Overall balance:

$$0 = \bar{w}_1 + \bar{w}_2 - \bar{w} \tag{1-1}$$

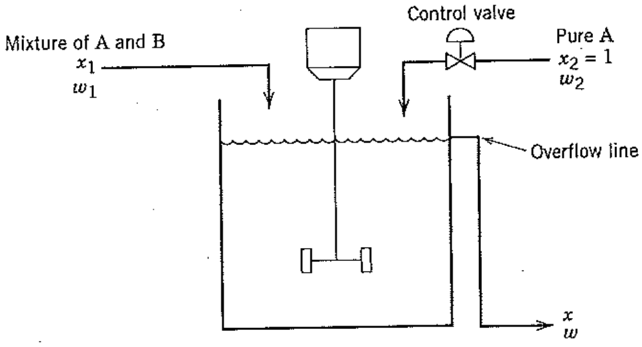


Figure 1.3 Stirred-tank blending system.

Component A balance:

$$0 = \overline{w}_1\overline{x}_1 + \overline{w}_2\overline{x}_2 - \overline{w}\,\overline{x} \tag{1-2}$$

The overbar over a symbol denotes its nominal steady-state value, for example, the value used in the process design. According to the process description, $\overline{x}_2 = 1$ and $\overline{x} = x_{sp}$. Solving Eq. 1-1 for $\overline{w}$, substituting these values into Eq. 1-2, and rearranging gives:

$$\overline{w}_2 = \overline{w}_1 \frac{x_{sp} - \overline{x}_1}{1 - x_{sp}} \tag{1-3}$$

Equation 1-3 is the design equation for the blending system. If our assumptions are correct and if $x_1 = \overline{x}_1$, then this value of w_2 will produce the desired result, $x = x_{sp}$. But what happens if conditions change?

> ***Control Question.*** *Suppose that inlet concentration x_1 varies with time. How can we ensure that the outlet composition x remains at or near its desired value, x_{sp}?*

As a specific example, assume that x_1 increases to a constant value that is larger than its nominal value, $\overline{x}_1$. It is clear that the outlet composition will also increase due to the increase in inlet composition. Consequently, at this new steady state, $x > x_{sp}$.

Next we consider several strategies for reducing the effects of x_1 disturbances on x.

Method 1. *Measure x and adjust w_2.* It is reasonable to measure controlled variable x and then adjust w_2 accordingly. For example, if x is too high, w_2 should be reduced; if x is too low, w_2 should be increased. This control strategy could be implemented by a person (*manual control*). However, it would normally be more convenient and economical to automate this simple task (*automatic control*).

Method 1 can be implemented as a simple *control algorithm* (or *control law*),

$$w_2(t) = \overline{w}_2 + K_c[x_{sp} - x(t)] \tag{1-4}$$

where K_c is a constant called the *controller gain*. The symbols, $w_2(t)$ and $x(t)$, indicate that w_2 and x change with time. Equation 1-4 is an example of *proportional control* because the change in the flow rate, $w_2(t) - \overline{w}_2$, is proportional to the deviation from the set point, $x_{sp} - x(t)$. Consequently, a large deviation from set point produces a large corrective action, while a small deviation results in a small corrective action. Note that we require K_c to be positive because w_2 must increase when x decreases, and vice versa. However, in other control applications negative values of K_c are appropriate, as discussed in Chapter 8.

A schematic diagram of Method 1 is shown in Fig. 1.4. The outlet concentration is measured and transmitted to the controller as an electrical signal. (Electrical signals are shown as dashed lines in Fig. 1.4.) The controller executes the control law and sends the calculated value of w_2 to the control valve as an electrical signal. The control valve opens or closes accordingly. In Chapters 8 and 9 we consider process instrumentation and control hardware in more detail.

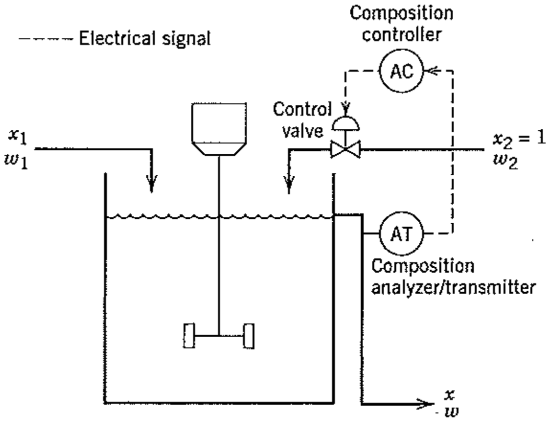


Figure 1.4 Blending system and Control Method 1.

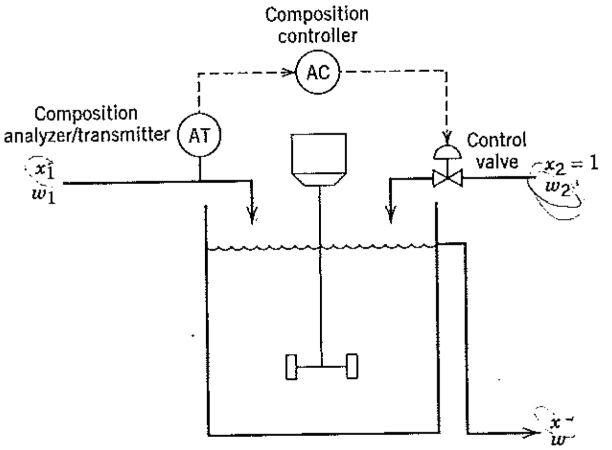


Figure 1.5 Blending system and Control Method 2.

Method 2. Measure x_1, adjust w_2. As an alternative to Method 1, we could measure disturbance variable x_1 and adjust w_2 accordingly. Thus, if $x_1 > \bar{x}_1$, we would decrease w_2 so that $w_2 < \bar{w}_2$. If $x_1 < \bar{x}_1$, we would increase w_2. A control law based on Method 2 can be derived from Eq. 1-3 by replacing $\bar{x}_1$ with $x_1(t)$ and $\bar{w}_2$ with $w_2(t)$:

$$w_2(t) = \bar{w}_1 \frac{x_{sp} - x_1(t)}{1 - x_{sp}} \tag{1-5}$$

The schematic diagram for Method 2 is shown in Fig. 1.5. Because Eq. 1-3 is valid only for steady-state conditions, it is not clear just how effective Method 2 will be during the transient conditions that occur after an x_1 disturbance.

Method 3. Measure x_1 and x, adjust w_2. This approach is a combination of Methods 1 and 2.

Method 4. Use a larger tank. If a larger tank is used, fluctuations in x_1 will tend to be damped out as a result of the larger volume of liquid. However, increasing tank size is an expensive solution due to the increased capital cost.

1.3 CLASSIFICATION OF PROCESS CONTROL STRATEGIES

Next, we will classify the four blending control strategies of the previous section and discuss their relative advantages and disadvantages. Method 1 is an example of a *feedback control* strategy. The distinguishing feature of feedback control is that the controlled variable is measured and the measurement is used to adjust the manipulated variable. For feedback control, the disturbance variable is *not* measured.

It is important to make a distinction between *negative feedback* and *positive feedback*. In the engineering literature, negative feedback refers to the desirable situation where the corrective action taken by the controller forces the controlled variable toward the set point. On the other hand, when positive feedback occurs, the controller makes things worse by forcing the controlled variable farther away from the set point. For example, in the blending control problem, positive feedback takes place if $K_c < 0$ because w_2 will increase when x increases.[1] Clearly, it is of paramount importance to ensure that a feedback control system incorporate negative feedback rather than positive feedback.

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[1] Note that social scientists use the terms, negative feedback and positive feedback, in a very different way. For example, they would say that teachers provide "positive feedback" when they compliment students who correctly do assignments. Criticism of a poor performance would be an example of "negative feedback."

Table 1.1 Concentration Control Strategies for the Blending System

Method	Measured Variable	Manipulated Variable	Category
1	x	w_2	FB
2	x_1	w_2	FF
3	x_1 and x	w_2	FF/FB
4	—	—	Design change

FB = feedback control; FF = feedforward control; FF/FB = feedforward control and feedback control.

An important advantage of feedback control is that corrective action occurs regardless of the source of the disturbance. For example, in the blending process, the feedback control law in (1-4) can accommodate disturbances in w_1, as well as x_1. Its ability to handle disturbances of unknown origin is a major reason why feedback control is the dominant process control strategy. Another important advantage is that feedback control reduces the sensitivity of the controlled variable to unmeasured disturbances and process changes. However, feedback control does have a fundamental limitation: no corrective action is taken until after the disturbance has upset the process, that is, until after the controlled variable deviates from the set point. This shortcoming is evident from the control law of (1-4).

Method 2 is an example of a *feedforward control strategy*. The distinguishing feature of feedforward control is that the disturbance variable is measured, but the controlled variable is not. The important advantage of feedforward control is that corrective action is taken *before* the controlled variable deviates from the set point. Ideally, the corrective action will cancel the effects of the disturbance so that the controlled variable is not affected by the disturbance. Although ideal cancellation is generally not possible, feedforward control can significantly reduce the effects of measured disturbances, as discussed in Chapter 15.

Feedforward control has three significant disadvantages: (i) the disturbance variable must be measured (or accurately estimated), (ii) no corrective action is taken for unmeasured disturbances, and (iii) a process model is required. For example, the feedforward control strategy for the blending system (Method 2) does not take any corrective action for unmeasured w_1 disturbances. In principle, we could deal with this situation by measuring both x_1 and w_1 and then adjusting w_2 accordingly. However, in industrial applications it is generally uneconomical to attempt to measure all potential disturbances. A more practical approach is to use a combined feedforward-feedback control system, where feedback control provides corrective action for unmeasured disturbances, while feedforward control reacts to eliminate measured disturbances before the controlled variable is upset. Consequently, in industrial applications feedforward control is normally used in combination with feedback control. This approach is illustrated by Method 3, a combined feedforward-feedback control strategy because both x and x_1 are measured.

Finally, Method 4 consists of a process design change and thus is not really a control strategy. The four strategies for the stirred-tank blending system are summarized in Table 1.1.

1.4 A MORE COMPLICATED EXAMPLE—A DISTILLATION COLUMN

The blending control system in the previous section is quite simple because there is only one controlled variable and one manipulated variable. For most practical applications, there are multiple controlled variables and multiple manipulated variables. As a representative example, we consider the distillation column in Fig. 1.6 that has five controlled variables and five manipulated variables. The controlled variables are product compositions, x_D and x_B, column pressure, P, and the liquid levels in the reflux drum and column base, h_D and h_B. The five manipulated variables are product flow rates, D and B, reflux flow, R, and the heat duties for the condenser and reboiler, Q_D and Q_B. The heat duties are adjusted via the control valves on the coolant and heating medium lines. The feed stream is assumed to come from an upstream unit. Thus, the feed flow rate cannot be manipulated, but it can be measured and used for feedforward control.

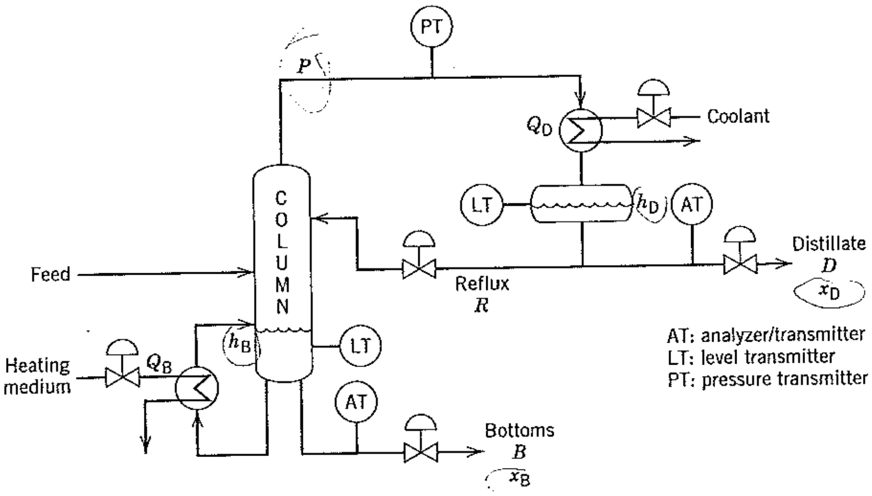


Figure 1.6 Controlled and manipulated variables for a typical distillation column.

A conventional *multiloop control* strategy for this distillation column would consist of five feedback control loops. Each control loop uses a single manipulated variable to control a single controlled variable. But how should the controlled and manipulated variables be paired? The total number of different multiloop control configurations that could be considered is 5! or 120. Many of these control configurations are impractical or unworkable such as any configuration that attempts to control the base level h_B by manipulating distillate flow D or condenser heat duty Q_D. However, even after the infeasible control configurations are eliminated, there are still many reasonable configurations left. Thus, there is a need for systematic techniques that can identify the most promising configurations. Fortunately, such tools are available and are discussed in Chapter 18.

For control applications, where conventional multiloop control systems are not satisfactory, an alternative approach, *multivariable control*, can be advantageous. In multivariable control, each manipulated variable is adjusted based on the measurements of all the controlled variables rather than only a single controlled variable, as in multiloop control. The adjustments are based on a dynamic model of the process that indicates how the manipulated variables affect the controlled variables. Consequently, the performance of multivariable control, or any model-based control technique, will depend heavily on the accuracy of the process model. A specific type of multivariable control, *model predictive control*, has had a major impact on industrial practice, as discussed in Chapter 20.

1.5 THE HIERARCHY OF PROCESS CONTROL ACTIVITIES

As mentioned earlier, the chief objective of process control is to maintain a process at the desired operating conditions, safely and efficiently, while satisfying environmental and product quality requirements. So far, we have emphasized one process control activity, keeping controlled variables at specified set points. But there are other important activities that we will now briefly describe.

In Fig. 1.7 the process control activities are organized in the form of a hierarchy with required functions at the lower levels and desirable, but optional, functions at the higher levels. The time scale for each activity is shown on the left side of Fig. 1.7. Note that the frequency of execution is much lower for the higher-level functions.

Measurement and Actuation (Level 1)

Measurement devices (sensors and transmitters) and actuation equipment (for example, control valves) are used to measure process variables and implement the calculated control actions. These devices are interfaced to the control system, usually digital control equipment such as a digital

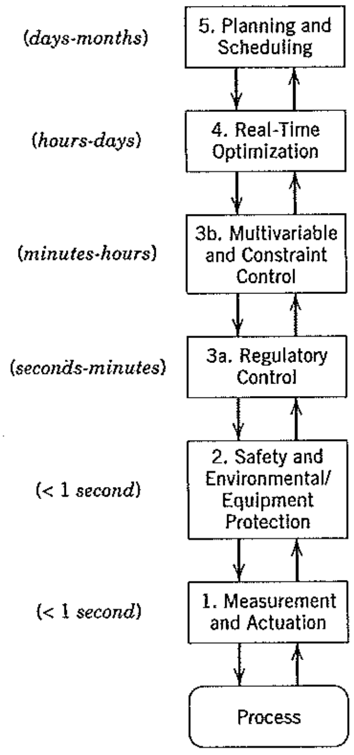


Figure 1.7 Hierarchy of process control activities.

computer. Clearly, the measurement and actuation functions are an indispensable part of any control system.

Safety and Environmental/Equipment Protection (Level 2)

The Level 2 functions play a critical role by ensuring that the process is operating safely and satisfies environmental regulations. As discussed in Chapter 10, process safety relies on the principle of *multiple protection layers* that involve groupings of equipment and human actions. One layer includes process control functions, such as alarm management during abnormal situations, and *safety instrumented systems* for emergency shutdowns. The safety equipment (including sensors and control valves) operates independently of the regular instrumentation used for regulatory control in Level 3a. Sensor validation techniques can be employed to confirm that the sensors are functioning properly.

Regulatory Control (Level 3a)

As mentioned earlier, successful operation of a process requires that key process variables such as flow rates, temperatures, pressures, and compositions be operated at, or close to, their set points. This Level 3a activity, *regulatory control,* is achieved by applying standard feedback and feedforward control techniques (Chapters 11–15). If the standard control techniques are not satisfactory, a variety of advanced control techniques are available (Chapters 16–18). In recent years, there has been increased interest in monitoring control system performance (Chapter 21).

Multivariable and Constraint Control (Level 3b)

Many difficult process control problems have two distinguishing characteristics: (i) significant interactions occur among key process variables, and (ii) inequality constraints exist for manipulated and controlled variables. The inequality constraints include upper and lower limits. For example, each manipulated flow rate has an upper limit determined by the pump and control valve characteristics. The lower limit may be zero or a small positive value based on safety considerations. Limits on controlled variables reflect equipment constraints (for example, metallurgical limits) and the operating objectives for the process. For example, a reactor temperature may have an upper limit to avoid undesired side reactions or catalyst degradation, and a lower limit to ensure that the reaction(s) proceed.

The ability to operate a process close to a limiting constraint is an important objective for advanced process control. For many industrial processes, the optimum operating condition occurs at a constraint limit, for example, the maximum allowed impurity level in a product stream. For these situations, the set point should not be the constraint value because a process disturbance could force the controlled variable beyond the limit. Thus, the set point should be set conservatively, based on the ability of the control system to reduce the effects of disturbances. This situation is illustrated in Fig. 1.8. For (*a*), the variability of the controlled variable is quite high, and consequently, the set point must be specified well below the limit. For (*b*), the improved control strategy has reduced the variability; consequently, the set point can be moved closer to the limit, and the process can be operated closer to the optimum operating condition.

The standard process control techniques of Level 3a may not be adequate for difficult control problems that have serious process interactions and inequality constraints. For these situations, the advanced control techniques of Level 3b, *multivariable control* and *constraint control*, should be considered. In particular, the *model predictive control* (*MPC*) strategy was developed to deal with both process interactions and inequality constraints. MPC is the subject of Chapter 20.

Real-time Optimization (Level 4)

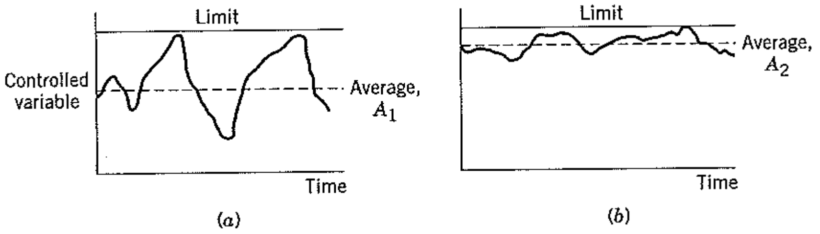
The optimum operating conditions for a plant are determined as part of the process design. But during plant operations, the optimum conditions can change frequently owing to changes in equipment availability, process disturbances, and economic conditions (for example, raw material costs and product prices). Consequently, it can be very profitable to recalculate the optimum operating conditions on a regular basis. This Level 4 activity, *real-time optimization* (*RTO*), is the subject of Chapter 19. The new optimum conditions are then implemented as set points for controlled variables.

The RTO calculations are based on a steady-state model of the plant and economic data such as costs and product values. A typical objective for the optimization is to minimize operating cost or maximize the operating profit. The RTO calculations can be performed for a single process unit and/or on a plantwide basis.

The Level 4 activities also include data analysis to ensure that the process model used in the RTO calculations is accurate for the current conditions. Thus, *data reconciliation* techniques can be used to ensure that steady-state mass and energy balances are satisfied. Also, the process model can be updated using parameter estimation techniques and recent plant data (Chapter 7).

Planning and Scheduling (Level 5)

The highest level of the process control hierarchy is concerned with planning and scheduling operations for the entire plant. For continuous processes, the production rates of all products and intermediates must be planned and coordinated, based on equipment constraints, storage capacity, sales



(*a*) (*b*)

Figure 1.8 Process variability over time: (a) before improved process control; (b) after.

projections, and the operation of other plants, sometimes on a global basis. For the intermittent operation of batch and semi-batch processes, the production control problem becomes a batch scheduling problem based on similar considerations. Thus, planning and scheduling activities pose difficult optimization problems that are based on both engineering considerations and business projections.

Summary of the Process Control Hierarchy

The activities of Levels 1, 2 and 3a in Fig. 1.7 are required for all manufacturing plants, while the activities in Levels 3b–Level 5 are optional but can be very profitable. The decision to implement one or more of these higher-level activities depends very much on the application and the company. The decision hinges strongly on economic considerations (for example, a cost/benefit analysis), and company priorities for their limited resources, both human and financial. The immediacy of the activity decreases from Level 1 to Level 5 in the hierarchy. However, the amount of analysis and the computational requirements increase from the lowest level to the highest level. The process control activities at different levels should be carefully coordinated and require information transfer from one level to the next. The successful implementation of these process control activities is a critical factor in making plant operation as profitable as possible.

1.6 AN OVERVIEW OF CONTROL SYSTEM DESIGN

In this section, we introduce some important aspects of control system design. However, it is appropriate first to describe the relationship between process design and process control.

Traditionally, process design and control system design have been separate engineering activities. Thus, in the traditional approach, control system design is not initiated until after plant design is well underway and major pieces of equipment may even have been ordered. This approach has serious limitations because the plant design determines the process dynamics as well as the operability of the plant. In extreme situations, the process may be uncontrollable, even though the design appears satisfactory from a steady-state point of view. A more desirable approach is to consider process dynamics and control issues early in the process design. The interaction between process design and control is analyzed in more detail in Chapters 10, 23, and 24.

Next, we consider two general approaches to control system design:

1. ***Traditional Approach.*** The control strategy and control system hardware are selected based on knowledge of the process, experience, and insight. After the control system is installed in the plant, the controller settings (such as controller gain K_c in Eq. 1-4) are adjusted. This activity is referred to as *controller tuning*.
2. ***Model-Based Approach.*** A dynamic model of the process is first developed that can be helpful in at least three ways: (i) It can be used as the basis for model-based controller design methods (Chapters 12 and 14); (ii) the dynamic model can be incorporated directly in the control law (for example, model predictive control); and (iii) the model can be used in a computer simulation to evaluate alternative control strategies and to determine preliminary values of the controller settings.

In this book, we advocate the philosophy that, for complex processes, a dynamic model of the process should be developed so that the control system can be properly designed. Of course, for many simple process control problems controller specification is relatively straightforward and a detailed analysis or an explicit model is not required. For complex processes, however, a process model is invaluable both for control system design and for an improved understanding of the process. As mentioned earlier, process control should be based on process understanding.

The major steps involved in designing and installing a control system using the model-based approach are shown in the flow chart of Fig. 1.9. The first step, formulation of the control objectives, is a critical decision. The formulation is based on the operating objectives for the plants and the process constraints. For example, in the distillation column control problem, the objective might be to regulate a key component in the distillate stream, the bottoms stream, or key components in both streams. An alternative would be to minimize energy consumption (e.g., heat input to the reboiler) while meeting product quality specifications on one or both product streams. The inequality constraints should include upper and lower limits on manipulated variables, conditions that lead to flooding or weeping in the column, and product impurity levels.

After the control objectives have been formulated, a dynamic model of the process is developed. The dynamic model can have a theoretical basis, for example, physical and chemical principles such as conservation laws and rates of reactions (Chapter 2), or the model can be developed empirically from experimental data (Chapter 7). If experimental data are available, the dynamic model should be validated, with the data and the model accuracy characterized. This latter information is useful for control system design and tuning.

The next step in the control system design is to devise an appropriate control strategy that will meet the control objectives while satisfying process constraints. As indicated in Fig. 1.9, this design activity is

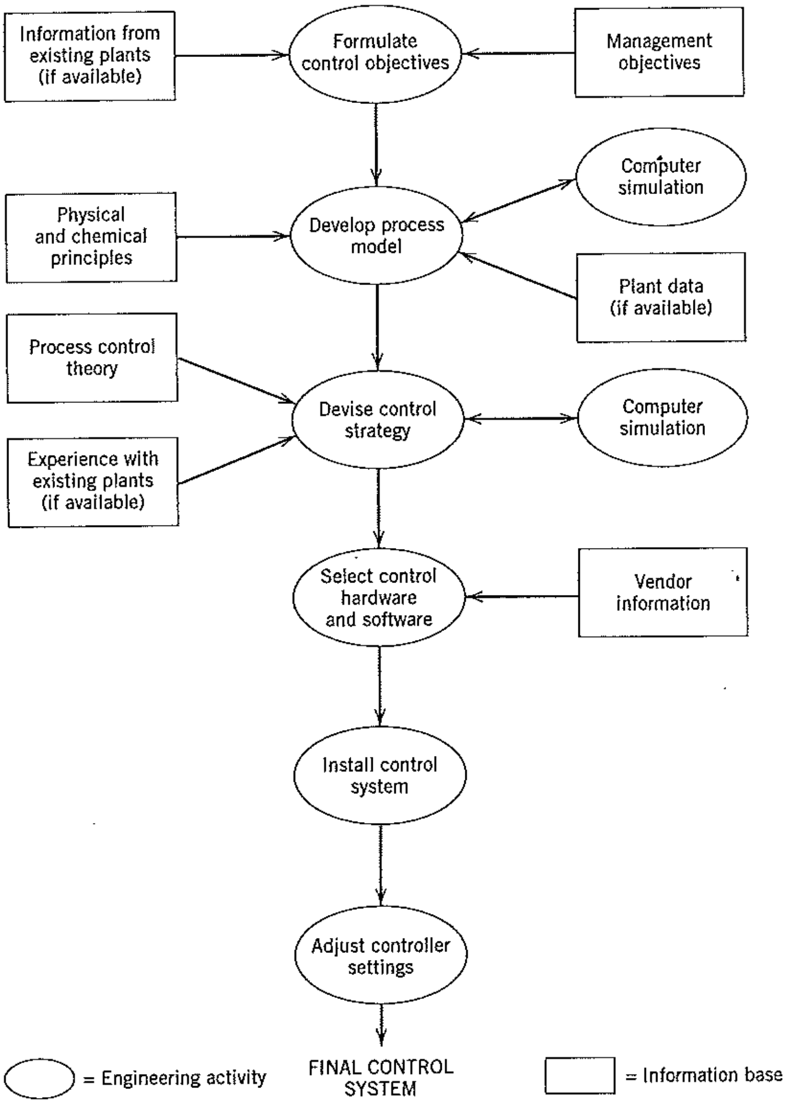


Figure 1.9 Major steps in control system development.

both an art and a science. Process understanding and the experience and preferences of the design team are key factors. Computer simulation of the controlled process is used to screen alternative control strategies and to provide preliminary estimates of appropriate controller settings.

Finally, the control system hardware and instrumentation are selected, ordered, and installed in the plant. Then the control system is tuned in the plant using the preliminary estimates from the design step as a starting point. Controller tuning usually involves trial-and-error procedures as described in Chapter 12.

SUMMARY

In this chapter we have introduced the basic concepts of process dynamics and process control. The process dynamics determine how a process responds during transient conditions, such as plant startups and shutdowns, grade changes, and unusual disturbances. Process control enables the process to be maintained at the desired operating conditions, safely and efficiently, while satisfying environmental and product quality requirements. Without effective process control, it would be impossible to operate large-scale industrial plants.

Two physical examples, a continuous blending system and a distillation column, have been used to introduce basic control concepts, notably, feedback and feedforward control. We also motivated the need for a systematic approach for the design of control systems for complex processes. Control system development consists of a number of separate activities that are shown in Fig. 1.9. In this book we advocate the design philosophy that, for complex processes, a dynamic model of the process should be developed so that the control system can be properly designed.

A hierarchy of process control activities was presented in Fig. 1.7. Process control plays a key role in ensuring process safety and protecting personnel, equipment, and the environment. Controlled variables are maintained near their set points by the application of regulatory control techniques and advanced control techniques such as multivariable and constraint control. Real-time optimization can be employed to determine the optimum controller set points for current operating conditions and constraints. The highest level of the process control hierarchy is concerned with planning and scheduling operations for the entire plant. The different levels of process control activity in the hierarchy are related and should be carefully coordinated.

EXERCISES

1.1 Which of the following statements are true?
  (a) *Feedback* and *feedforward control* both require a measured variable.
  (b) The process variable to be controlled is measured in *feedback control.*
  (c) *Feedforward control* can be perfect in the theoretical sense that the controller can take action via the manipulated variable even while the controlled variable remains equal to its desired value.
  (d) *Feedforward control* can provide perfect control; that is, the output can be kept at its desired value, even with an imperfect process model.
  (e) *Feedback control* will always take action regardless of the accuracy of any process model that was used to design it and the source of a disturbance.

1.2 Consider a home heating system consisting of a natural gas-fired furnace and a thermostat. In this case the process consists of the interior space to be heated. The thermostat contains both the measuring element and the controller. The furnace is either on (heating) or off. Draw a schematic diagram for this control system. On your diagram, identify the controlled variables, manipulated variables, and disturbance variables. Be sure to include several possible sources of disturbances that can affect room temperature.

1.3 In addition to a thermostatically-operated home heating system, identify two other feedback control systems that can be found in most residences. Describe briefly how each of them works: include sensor, actuator, and controller information.

1.4 Does a typical microwave oven utilize feedback control to set cooking temperature or to determine if the food is "cooked"? If not, what mechanism is used? Can you think of any disadvantages to this approach, for example, in thawing and cooking foods?

1.5 Driving an automobile safely requires a large amount of individual skill. Even if not generally recognized, the driver needs an intuitive ability to utilize feedforward and feedback control methods.

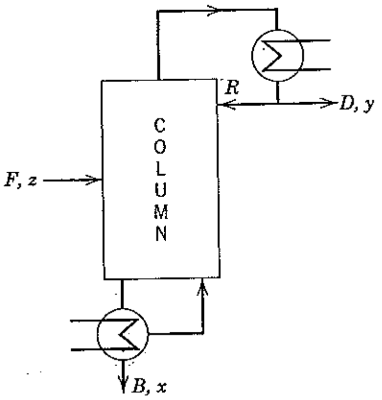
(a) In the process of steering a car, the objective is to keep the vehicle generally centered in the proper traffic lane. Thus, the controlled variable is some measure of that distance. If so, how is feedback control used to accomplish this objective? Identify the sensor(s), the actuator, how the appropriate control action is determined, and some likely disturbances.

(b) The process of braking/accelerating an auto is highly complex, requiring the skillful use of both feedback and feedforward mechanisms to drive safely. For feedback control, the driver normally uses distance to the vehicle ahead as the measured variable. The "set point" then is often recommended to be some distance related to speed, for example, one car length separation for each 10 mph. If this assertion is correct, how does feedforward control come into the accelerating/braking process when one is attempting to drive in traffic at a constant speed? In other words, what other information—in addition to distance separating the two vehicles, which obviously should never equal zero—does the driver utilize to avoid colliding with the car ahead?

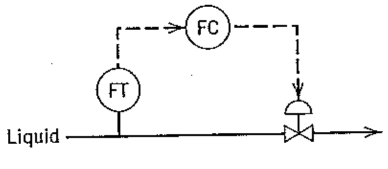
1.6 The distillation column shown in the drawing is used to distill a binary mixture. Symbols x, y, and z denote mole fractions of the more volatile component, while B, D, R, and F represent molar flow rates. It is desired to control distillate composition y despite disturbances in feed flow rate F. All flow rates can be measured and manipulated with the exception of F, which can only be measured. A composition analyzer provides measurements of y.

(a) Propose a feedback control method and sketch the schematic diagram.

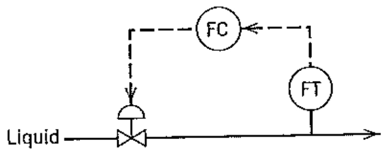
(b) Suggest a feedforward control method and sketch the schematic diagram.



1.7 Two flow control loops are shown in the drawing. Indicate whether each system is either a feedback or a feedforward control system. Justify your answer. It can be assumed that the distance between the flow transmitter (FT) and the control valve is quite small in each system.



System I



System II

1.8 I. M. Appelpolscher, supervisor of the process control group of the Ideal Gas Company, has installed a 25 × 40 × 5-ft swimming pool in his backyard. The pool contains level and temperature sensors used with feedback controllers to maintain the pool level and temperature at desired values. Appelpolscher is satisfied with the level control system, but he feels that the addition of one or more feedforward controllers would help maintain the pool temperature more nearly constant. As a new member of the process control group, you have been selected to check Appelpolscher's mathematical analysis and to give your advice. The following information may or may not be pertinent to your analysis:

(i) Appelpolscher is particular about cleanliness and thus has a high-capacity pump that continually recirculates the water through an activated charcoal filter.

(ii) The pool is equipped with a natural gas-fired heater that adds heat to the pool at a rate $Q(t)$ that is directly proportional to the output signal from the controller $p(t)$.

(iii) There is a leak in the pool, which Appelpolscher has determined is constant equal to F (volumetric flow rate). The liquid-level control system adds water from the city supply system to maintain the level in the pool exactly at the specified level. The temperature of the water in the city system is T_w, a variable.

(iv) A significant amount of heat is lost by conduction to the surrounding ground, which has a constant, year-round temperature T_G. Experimental tests by Appelpolscher showed that essentially all of the temperature drop between the pool and the ground occurred across the homogeneous layer of gravel that surrounded his pool. The gravel thickness is Δx, and the overall thermal conductivity is k_G.

(v) The main challenge to Appelpolscher's modeling ability was the heat loss term accounting for convection, conduction, radiation, and evaporation to the atmosphere. He determined that the heat losses per unit area of open water could be represented by

$$\text{losses} = U(T_p - T_a)$$

where

T_p = temperature of pool
T_a = temperature of the air, a variable
U = overall heat transfer coefficient

Appelpolscher's detailed model included radiation losses and heat generation due to added chemicals, but he determined that these terms were negligible.

(a) Draw a schematic diagram for the pool and all control equipment. Show all inputs and outputs, including all disturbance variables.
(b) What additional variable(s) would have to be measured to add feedforward control to the existing pool temperature feedback controller?
(c) Write a steady-state energy balance. How can you determine which of the disturbance variables you listed in part (a) are most/least likely to be important?
(d) What recommendations concerning the prospects of adding feedforward control would you make to Appelpolscher?