Control Objectives and Benefits

CHAPTER 2

2.1 INTRODUCTION

The first chapter provided an overview of process control in which the close association between process control and plant operation was noted. As a consequence, control objectives are closely tied to process goals, and control benefits are closely tied to attaining these goals. In this chapter the control objectives and benefits are discussed thoroughly, and several process examples are presented. The control objectives provide the basis for all technology and design methods presented in subsequent chapters of the book.

While this book emphasizes the contribution made by automatic control, control is only one of many factors that must be considered in improving process performance. Three of the most important factors are shown in Figure 2.1, which indicates that proper equipment design, operating conditions, and process control should all be achieved simultaneously to attain safe and profitable plant operation. Clearly, equipment should be designed to provide good dynamic responses in addition to high steady-state profit and efficiency, as covered in process design courses and books. Also, the plant operating conditions, as well as achieving steady-state plant objectives, should provide flexibility for dynamic operation. Thus, achieving excellence in plant operation requires consideration of all factors. This book addresses all three factors; it gives guidance on how to design processes and select operating conditions favoring good dynamic performance, and it presents automation methods to adjust the manipulated variables.

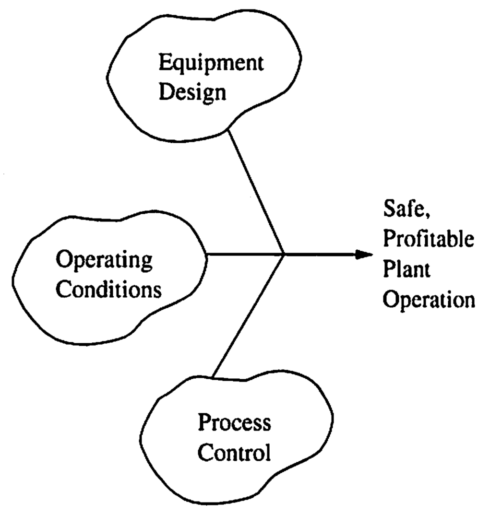

FIGURE 2.1

Schematic representation of three critical elements for achieving excellent plant performance.

Control Objectives

1. Safety
2. Environmental Protection
3. Equipment Protection
4. Smooth Operation and Production Rate
5. Product Quality
6. Profit
7. Monitoring and Diagnosis

2.2 CONTROL OBJECTIVES

The seven major categories of control objectives were introduced in Chapter 1. They are discussed in detail here, with an explanation of how each influences the control design for the example process shown in Figure 2.2. The process separates two components based on their different vapor pressures. The liquid feed stream, consisting of components A and B, is heated by two exchangers in series. Then the stream flows through a valve to a vessel at a lower pressure. As a result of the higher temperature and lower pressure, the material forms two phases, with most of the A in the vapor and most of the B in the liquid. The exact compositions can be determined from an equilibrium flash calculation, which simultaneously solves the material, energy, and equilibrium expressions. Both streams leave the vessel for further processing, the vapor stream through the overhead line and the liquid stream out from the bottom of the vessel. Although a simple process, the heat exchanger with flash drum provides examples of all control objectives, and this process is analyzed quantitatively with control in Chapter 24.

A control strategy is also shown in Figure 2.2. Since we have not yet studied the calculations used by feedback controllers, you should interpret the controller as a linkage between a measurement and a valve. Thus, you can think of the feedback pressure control (PC) system as a system that measures the pressure and maintains the pressure close to its desired value by adjusting the opening of the valve in the overhead vapor pipe. The type of control calculation, which will be covered in depth in later chapters, is not critical for the discussions in this chapter.

Safety

The safety of people in the plant and in the surrounding community is of paramount importance. While no human activity is without risk, the typical goal is that working at an industrial plant should involve much less risk than any other activity in a person's life. *No compromise with sound equipment and control safety practices is acceptable.*

Plants are designed to operate safely at expected temperatures and pressures; however, improper operation can lead to equipment failure and release of potentially hazardous materials. Therefore, the process control strategies contribute to the overall plant safety by maintaining key variables near their desired values. Since these control strategies are important, they are automated to ensure rapid and complete implementation. In Figure 2.2, the equipment could operate at high pressures under normal conditions. If the pressure were allowed to increase too far beyond the normal value, the vessel might burst, resulting in injuries or death. Therefore, the control strategy includes a controller labelled "PC-1" that controls the pressure by adjusting the valve position (i.e., percent opening) in the vapor line.

Another consideration in plant safety is the proper response to major incidents, such as equipment failures and excursions of variables outside of their acceptable bounds. Feedback strategies cannot *guarantee* safe operation; a very large disturbance could lead to an unsafe condition. Therefore, an additional layer of control, termed an *emergency system*, is applied to enforce bounds on key variables. Typically, this layer involves either safely diverting the flow of material or shutting down the process when unacceptable conditions occur. The control strategies are usually not complicated; for example, an emergency control might stop the feed to a vessel when the liquid level is nearly overflowing. Proper design of these

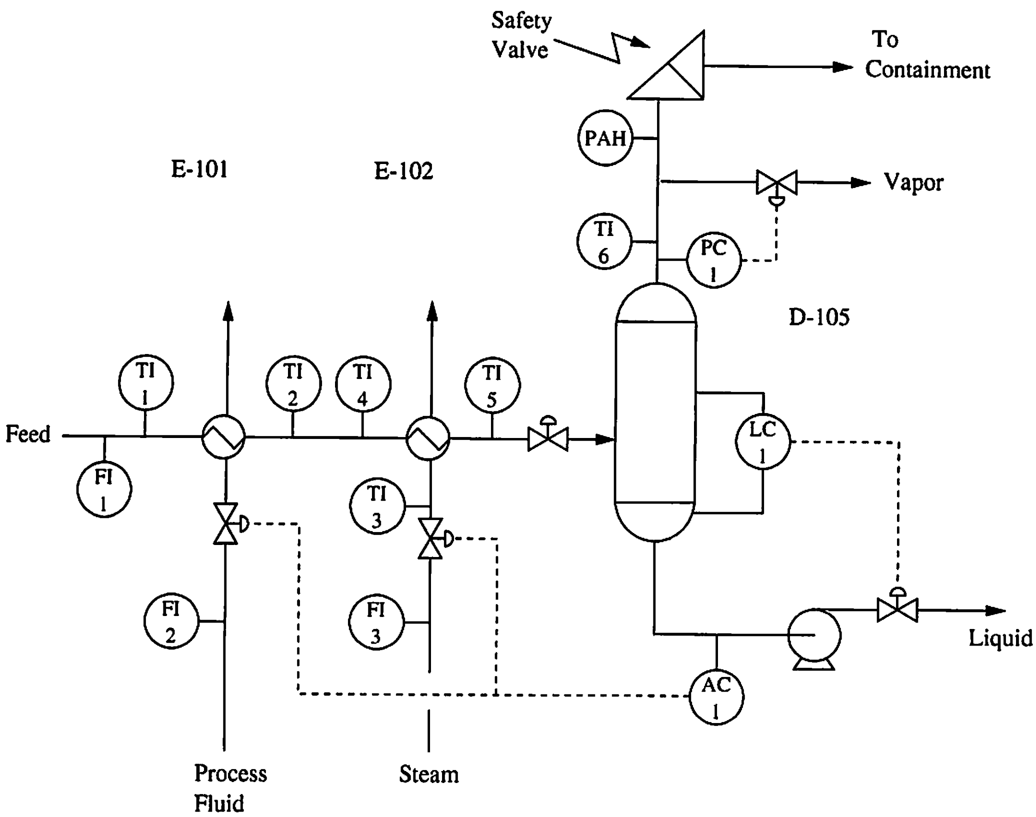

FIGURE 2.2

Flash separation process with control strategy.

emergency systems is based on a structured analysis of hazards (Battelle Laboratory, 1985; Warren Centre, 1986) that relies heavily on experience about expected incidents and on the reliability of process and control equipment.

In Figure 2.2, the pressure is controlled by the element labelled "PC." Normally, it maintains the pressure at or near its desired value. However, the control strategy relies on the proper operation of equipment like the pressure sensor and the valve. Suppose that the sensor stopped providing a reliable measurement; the control strategy could improperly close the overhead valve, leading to an unsafe pressure. The correct control design would include an additional strategy using independent equipment to prevent a very high pressure. For example, the safety valve shown in Figure 2.2 is closed unless the pressure rises above a specified maximum; then, it opens to vent the excess vapor. It is important to recognize that this safety relief system is called on to act infrequently, perhaps once per year or less often; therefore, its design should include highly reliable components to ensure that it performs properly when needed.

Environmental Protection

Protection of the environment is critically important. This objective is mostly a process design issue; that is, the process must have the capacity to convert potentially toxic components to benign material. Again, control can contribute to the proper operation of these units, resulting in consistently low effluent concentrations. In addition, control systems can divert effluent to containment vessels should any extreme disturbance occur. The stored material could be processed at a later time when normal operation has been restored.

In Figure 2.2, the environment is protected by containing the material within the process equipment. Note that the safety release system directs the material for containment and subsequent "neutralization," which could involve recycling to the process or combusting to benign compounds. For example, a release system might divert a gaseous hydrocarbon to a flare for combustion, and it might divert a water-based stream to a holding pond for subsequent purification through biological treatment before release to a water system.

Equipment Protection

Much of the equipment in a plant is expensive and difficult to replace without costly delays. Therefore, operating conditions must be maintained within bounds to prevent damage. The types of control strategies for equipment protection are similar to those for personnel protection, that is, controls to maintain conditions near desired values and emergency controls to stop operation safely when the process reaches boundary values.

In Figure 2.2, the equipment is protected by maintaining the operating conditions within the expected temperatures and pressures. In addition, the pump could be damaged if no liquid were flowing through it. Therefore, the liquid level controller, by ensuring a reservoir of liquid in the bottom of the vessel, protects the pump from damage. Additional equipment protection could be provided by adding an emergency controller that would shut off the pump motor when the level decreased below a specified value.

Smooth Operation and Production Rate

A chemical plant includes a complex network of interacting processes; thus, the smooth operation of a process is desirable, because it results in few disturbances to all integrated units. Naturally, key variables in streams leaving the process should be maintained close to their desired values (i.e., with small variation) to prevent disturbances to downstream units. In Figure 2.2, the liquid from the vessel bottoms is processed by downstream equipment. The control strategy can be designed to make slow, smooth changes to the liquid flow. Naturally, the liquid level will not remain constant, but it is not required to be constant; the level must only remain within specified limits. By the use of this control design, the downstream units would experience fewer disturbances, and the overall plant would perform better.

There are additional ways for upsets to be propagated in an integrated plant. For example, when the control strategy increases the steam flow to heat exchanger E-102, another unit in the plant must respond by generating more steam. Clearly, smooth manipulations of the steam flow require slow adjustments in the boiler operation and better overall plant operation. Therefore, we are interested in *both the controlled variables and the manipulated variables.* Ideally, we would like to have tight regulation of the controlled variables and slow, smooth adjustment of the manipulated variables. As we will see, this is not usually possible, and some compromise is required.

People who are operating a plant want a simple method for maintaining the production rate at the desired value. We will include the important production rate goal in this control objective. For the flash process in Figure 2.2, the natural method for achieving the desired production rate is to adjust the feed valve located before the flash drum so that the feed flow rate F_1 has the desired value.

Product Quality

The final products from the plant must meet demanding quality specifications set by purchasers. The specifications may be expressed as compositions (e.g., percent of each component), physical properties (e.g., density), performance properties (e.g., octane number or tensile strength), or a combination of all three. Process control contributes to good plant operation by maintaining the operating conditions required for excellent product quality. Improving product quality control is a major economic factor in the application of digital computers and advanced control algorithms for automation in the process industries.

In Figure 2.2, the amount of component A, the material with the higher vapor pressure, is to be controlled in the liquid stream. Based on our knowledge of thermodynamics, we know that this value can be controlled by adjusting the flash temperature or, equivalently, the heat exchanged. Therefore, a control strategy would be designed to measure the composition in real time and adjust the heating medium flows that exchange heat with the feed.

Profit

Naturally, the typical goal of the plant is to return a profit. In the case of a utility such as water purification, in which no income from sales is involved, the equivalent goal is to provide the product at lowest cost. Before achieving the profit-oriented goal, selected independent variables are adjusted to satisfy the first five higher-priority control objectives. Often, some independent operating variables are not specified after the higher objectives (that is, including product quality but excepting profit) have been satisfied. When additional variables (degrees of freedom) exist, the control strategy can increase profit *while satisfying all other objectives.*

In Figure 2.2 all other control objectives can be satisfied by using exchanger E-101, exchanger E-102, or a combination of the two, to heat the inlet stream. Therefore, the control strategy can select the correct exchanger based on the cost of the two heating fluids. For example, if the process fluid used in E-101 were less costly, the control strategy would use the process stream for heating preferentially and use steam only when required for additional heating. How the control strategy would implement this policy, based on a selection hierarchy defined by the engineer, is covered in Chapter 22.

Monitoring and Diagnosis

Complex chemical plants require monitoring and diagnosis by people as well as excellent automation. Plant control and computing systems generally provide monitoring features for two sets of people who perform two different functions: (1) the immediate safety and operation of the plant, usually monitored by plant operators, and (2) the long-term plant performance analysis, monitored by supervisors and engineers.

The plant operators require very rapid information so that they can ensure that the plant conditions remain within acceptable bounds. If undesirable situations occur—or, one hopes, before they occur—the operator is responsible for rapid recognition and intervention to restore acceptable performance. While much of this routine work is automated, the people are present to address complex issues that are difficult to automate, perhaps requiring special information not readily available to the computing system. Since the person may be responsible for a plant section with hundreds of measured variables, excellent displays are required. These are usually in the form of trend plots of several associated variables versus time and of indicators in bar-chart form for easy identification of normal and abnormal operation. Examples are shown in Figure 2.3.

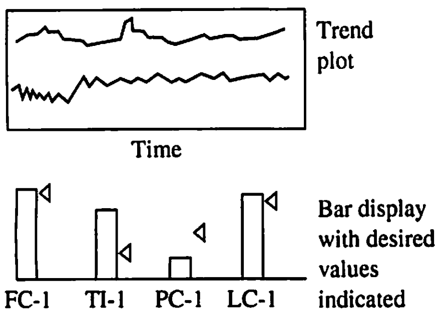

FIGURE 2.3

Examples of displays presented to a process operator.

Since the person cannot monitor all variables simultaneously, the control system includes an alarm feature, which draws the operator's attention to variables that are near limiting values selected to indicate serious maloperation. For example, a high pressure in the flash separator drum is undesirable and would at the least result in the safety valve opening, which is not desirable, because it diverts material and results in lost profit and because it may not always reclose tightly. Thus, the system in Figure 2.2 has a high-pressure alarm, PAH. If the alarm is activated, the operator might reduce the flows to the heat exchanger or of the feed to reduce pressure. This operator action might cause a violation of product specifications; however, maintaining the pressure within safe limits is more important than product quality. Every measured variable in a plant must be analyzed to determine whether an alarm should be associated with it and, if so, the proper value for the alarm limit.

Another group of people monitors the longer-range performance of the plant to identify opportunities for improvement and causes for poor operation. Usually, a substantial sample of data, involving a long time period, is used in this analysis, so that the effects of minor fluctuations are averaged out. Monitoring involves important measured and calculated variables, including equipment performances (e.g., heat transfer coefficients) and process performances (e.g., reactor yields and material balances). In the example flash process, the energy consumption would be monitored. An example trend of some key variables is given in Figure 2.4, which shows that the ratio of expensive to inexpensive heating source had an increasing trend. If the feed flow and composition did not vary significantly, one might suspect

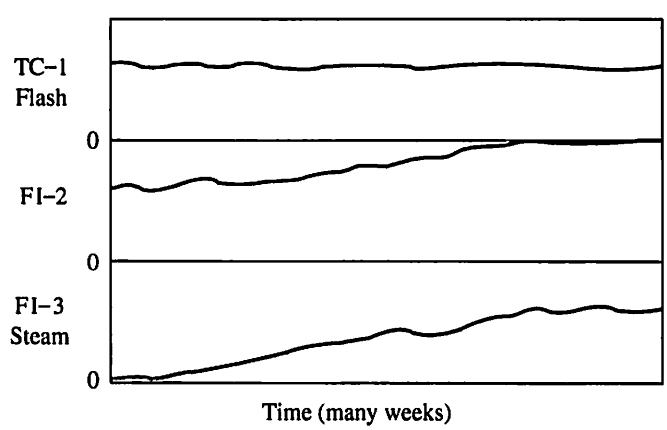

FIGURE 2.4

Example of long-term data, showing the increased use of expensive steam in the flash process.

that the heat transfer coefficient in the first heat exchanger, E-101, was decreasing due to fouling. Careful monitoring would identify the problem and enable the engineer to decide when to remove the heat exchanger temporarily for mechanical cleaning to restore a high heat transfer coefficient.

Previously, this monitoring was performed by hand calculations, which was a tedious and inefficient method. Now, the data can be collected, processed if additional calculations are needed, and reported using digital computers. This combination of ease and reliability has greatly improved the monitoring of chemical process plants.

Note that both types of monitoring—the rapid display and the slower process analysis—require people to make and implement decisions. This is another form of feedback control involving personnel, sometimes referred to as having "a person in the loop," with the "loop" being the feedback control loop. While we will concentrate on the automated feedback system in a plant, we must never forget that many of the important decisions in plant operation that contribute to longer-term safety and profitability are based on monitoring and diagnosis and implemented by people "manually."

Therefore,

> All seven categories of control objectives must be achieved simultaneously; failure to do so leads to unprofitable or, worse, dangerous plant operation.

Control Objectives

1. Safety
2. Environmental Protection
3. Equipment Protection
4. Smooth Operation and Production Rate
5. Product Quality
6. Profit
7. Monitoring and Diagnosis

In this section, instances of all seven goals were identified in the simple heater and flash separator. The analysis of more complex process plants in terms of the goals is a challenging task, enabling engineers to apply all of their chemical engineering skills. Often a team of engineers and operators, each with special experiences and insights, performs this analysis. Again, we see that control engineering skills are needed by all chemical engineers in industrial practice.

2.3 ▫ DETERMINING PLANT OPERATING CONDITIONS

A key factor in good plant operation is the determination of the best operating conditions, which can be maintained within small variation by automatic control strategies. Therefore, setting the control objectives requires a clear understanding of how the plant operating conditions are determined. A complete study of plant objectives requires additional mathematical methods for simulating and optimizing the plant operation. For our purposes, we will restrict our discussion in this section to small systems that can be analyzed graphically.

Determining the best operating conditions can be performed in two steps. First, the region of possible operation is defined. The following are some of the factors that limit the possible operation:

- Physical principles; for example, all concentrations ≥ 0
- Safety, environmental, and equipment protection
- Equipment capacity; for example, maximum flow
- Product quality

The region that satisfies all bounds is termed the feasible operating region or, more commonly, the *operating window.* Any operation within the operating window is possible. Violation of some of the limits, called *soft constraints,* would lead to poor product quality or reduction of long-term equipment life; therefore, short-term violations of soft constraints are allowed but are to be avoided. Violation of critical bounds, called *hard constraints,* could lead to injury or major equipment damage; violations of hard constraints are not acceptable under any foreseeable circumstances. The control strategy must take aggressive actions, including shutting down the plant, to prevent hard constraint violations. For both hard and soft constraints, debits are incurred for violating constraints, so the control system is designed to maintain operation within the operating window. While any operation within the window is possible and satisfies minimum plant goals, a great difference in profit can exist depending on the conditions chosen. Thus, the plant economics must be analyzed to determine the best operation within the window. The control strategy should be designed to maintain the plant conditions near their most profitable values.

The example shown in Figure 2.5 demonstrates the operating window for a simple, one-dimensional case. The example involves a fired heater (furnace) with a chemical reaction occurring as the fluid flows through the pipe or, as it is often called, the *coil.* The temperature of the reactor must be held between minimum (no reaction) and maximum (metal damage or excessive side reactions) temperatures. When economic objectives favor increased conversion of feed, the profit function monotonically increases with increasing temperature; therefore, the best operation would be at the maximum allowable temperature. However, the dynamic data show that the temperature varies about the desired value because of disturbances such as those in fuel composition and pressure. Therefore, the effectiveness of the control strategy in maximizing profit depends on reducing the variation of the temperature. A small variation means that the temperature can be operated very close to, without exceeding, the maximum constraint.

Another example is the system shown in Figure 2.6, where fuel and air are mixed and combusted to provide heat for a boiler. The ratio of fuel to air is important. Too little air (oxygen) means that some of the fuel is uncombusted and wasted, whereas excess air reduces the flame temperature and, thus, the heat trans-

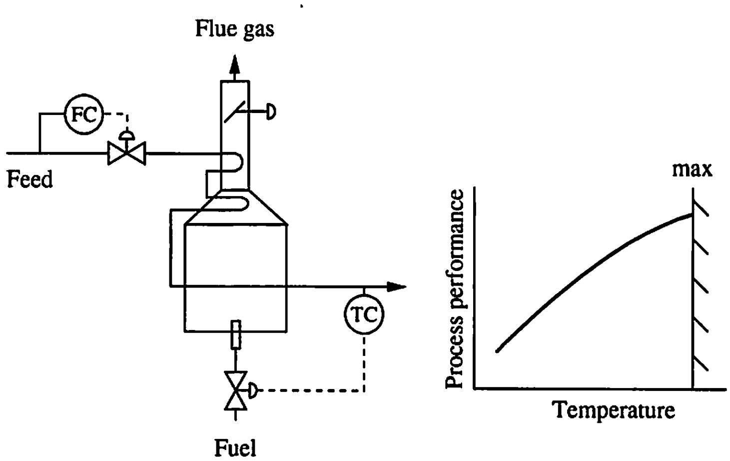

FIGURE 2.5

Example of operating window for fired-heater temperature.

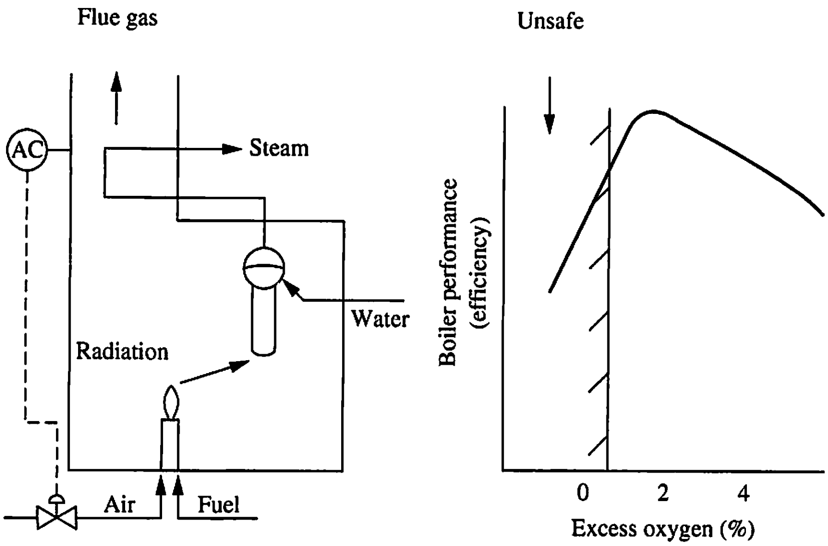

FIGURE 2.6

Example of operating window for boiler combustion flue gas excess oxygen.

fer. Therefore, the highest efficiency and most profitable operation are near the stoichiometric ratio. (Actually, the best value is usually somewhat above the stoichiometric ratio because of imperfect mixing, leakage, and complex combustion chemistry.) The maximum air flow is determined by the air compressor and is usually not a limitation, but a large excess of air leads to extremely high fuel costs. Therefore, the best plant operation is at the peak of the efficiency curve. An effective control strategy results in a small variation in the excess oxygen in the flue gas, allowing operation near the peak.

However, a more important factor is safety, which provides another reason for controlling the excess air. A deficiency of oxygen could lead to a dangerous condition because of unreacted fuel in the boiler combustion chamber. Should this situation occur, the fuel could mix with other air (that leaks into the furnace chamber) and explode. Therefore, the air flow should never fall below the stoichiometric value. Note that the control sketch in Figure 2.6 is much simpler than actual control designs for combustion systems (for example, API, 1977).

Finally, a third example demonstrates that this analysis can be extended to more than one dimension. We now consider the chemical reactor in Figure 2.5 with two variables: temperature and product flow. The temperature bounds are the same, and the product flow has a maximum limitation because of erosion of the pipe at the exit of the fired heater. The profit function, which would be calculated based on an analysis of the entire plant, is given as contours in the operating window in Figure 2.7. In this example, the maximum profit occurs outside the operating window and therefore cannot be achieved. The best operation inside the window would be at the maximum temperature and flow, which are found at the upper right-hand corner of the operating window. As we know, the plant cannot be operated exactly at this point because of unavoidable disturbances in variables such as feed pressure and fuel composition (which affects heat of combustion). However, good control designs can reduce the variation of temperature and flow so that desired values can be selected that nearly maximize the achievable profit while not violating the constraints. This situation is shown in Figure 2.7, where

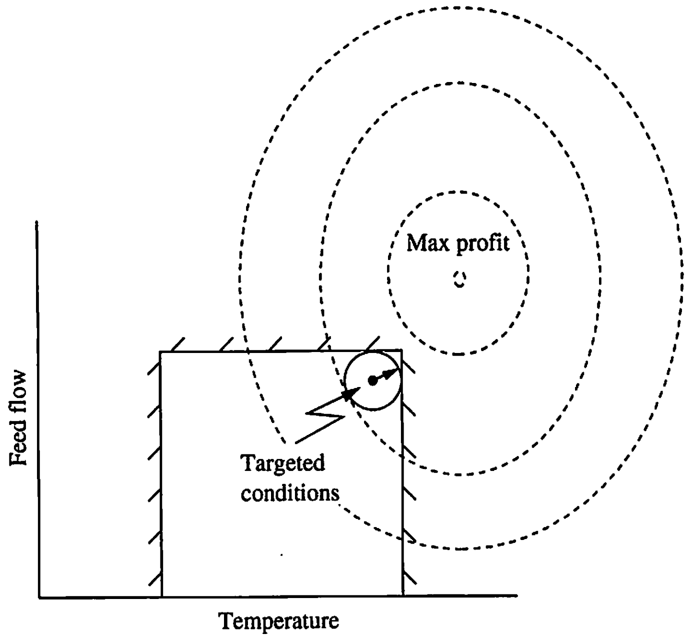

FIGURE 2.7

Example of operating window for the feed and temperature of a fired-heater chemical reactor.

a circle defines the variation expected about the desired values (Perkins, 1990; Narraway and Perkins, 1993). When control provides small variation, that is, a circle of small radius, the operation can be maintained closer to the best operation.

All of these examples demonstrate that

> Process control improves plant performance by reducing the variation of key variables. When the variation has been reduced, the desired value of the controlled variable can be adjusted to increase profit.

Note that simply reducing the variation does not always improve plant operation. The profit contours within the operating window must be analyzed to determine the best operating conditions that take advantage of the reduced variation. Also, it is important to recognize that the theoretical maximum profit cannot usually be achieved because of inevitable variation due to disturbances. This situation should be included in the economic analysis of all process designs.

2.4 ◻ BENEFITS FOR CONTROL

The previous discussion of plant operating conditions provides the basis for calculating the benefits for excellent control performance. In all of the examples discussed qualitatively in the previous section, the economic benefit resulted from reduced variation of key variables. Thus, the calculation of benefits considers the effect of variation on plant profit. Before the method is presented, it is emphasized that the highest-priority control objectives—namely, safety, environmental protection, and equipment protection—are *not* analyzed by the method described in this section. Although the control designs for these objectives often reduce variation, they are not selected for increasing profit but rather for providing safe, reliable plant operation.

Once the profit function has been determined, the benefit method needs to characterize the variation of key plant variables. This can be done through the calculation shown schematically in Figure 2.8. The plant operating data, which is usually given as a plot or trend versus time, can be summarized by a frequency distribution. The frequency distribution can be determined by taking many sample measurements of the process variable, usually separated by a constant time period, and counting the number of measurements whose values fall in each of several intervals within the range of data values. The total time period covered must be long compared to the dynamics of the process, so that the effects of time correlation in the variable and varying disturbances will be averaged out.

The resulting distribution is plotted as frequency; that is, as fraction or percent of measurements falling within each interval versus the midpoint value of that interval. Such a plot is called a *frequency distribution* or *histogram.* If the variable were constant, perhaps due to perfect control or the presence of no disturbances, the distribution would have one bar, at the constant value, rising to 1.0 (or 100%). As the variation in the values increases, the distribution becomes broader; thus, the frequency distribution provides a valuable summary of the variable variation.

The distribution could be described by its *moments;* in particular, the mean and standard deviation are often used in describing the behavior of variables in feedback systems (Snedecor and Cochran, 1980; Bethea and Rhinehart, 1991). These values can be calculated from the plant data according to the following

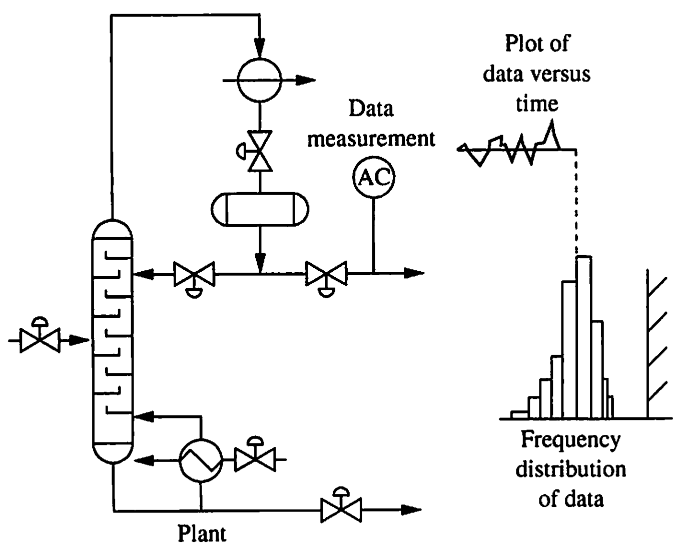

FIGURE 2.8

Schematic presentation of the method for representing the variability in plant data.

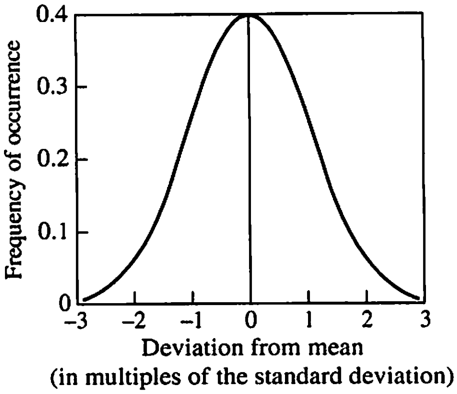

FIGURE 2.9

Normal distribution.

equations:

$$\text{Mean} = \bar{Y} = \frac{1}{n}\sum_{i=1}^{n} Y_i \tag{2.1}$$

$$\text{Standard deviation} = s_Y = \sqrt{\frac{\sum_{i=1}^{n}(Y_i - \bar{Y})^2}{n-1}} \tag{2.2}$$

where Y_i = measured value of variable
s_Y^2 = variance
n = number of data points

When the experimental distribution can be characterized by the standard normal distribution, the variation about the mean is characterized by the standard deviation as is shown in Figure 2.9. (Application of the central limit theorem to data whose underlying distribution is not normal often results in the valid use of the normal distribution.) When the number of data in the sample are large, the estimated (sample) standard deviation is approximately equal to the population standard deviation, and the following relationships are valid for the normally distributed variable:

About 68.2% of the variable values are within $\pm s$ of mean.
About 95.4% of the variable values are within $\pm 2s$ of mean.
About 99.7% of the variable values are within $\pm 3s$ of mean.

In all control performance and benefits analysis, the mean and standard deviation can be used in place of the frequency distribution when the distribution is normal. As is apparent, a narrow distribution is equivalent to a small standard deviation. Although the process data can often be characterized by a normal distribution, the method for calculating benefits does *not depend on the normal distribution*, which was introduced here to relate the benefits method to statistical terms often used to describe the variability of data.

The empirical histogram provides how often—that is, what percentage of the time—a variable has a certain value, with the value for each histogram entry taken as the center of the variable interval. The performance of plant operation at each variable value can be determined from the *performance function*. Depending on the plant, the performance function could be reactor conversion, efficiency, production rate, profit, or other variable that characterizes the quality of operation. The average performance for a set of representative data (that is, frequency distribution) is calculated by combining the histogram and profit function according to the following equation (Bozenhardt and Dybeck, 1986; Marlin et al., 1991; and Stout and Cline, 1976).

$$P_{\text{ave}} = \sum_{j=1}^{M} F_j P_j \tag{2.3}$$

where P_{ave} = average process performance
F_j = fraction of data in interval $j = N_j/N_T$
N_j = number of data points in interval j
N_T = total number of data points
P_j = performance measured at the midpoint of interval j
M = number of intervals in the frequency distribution

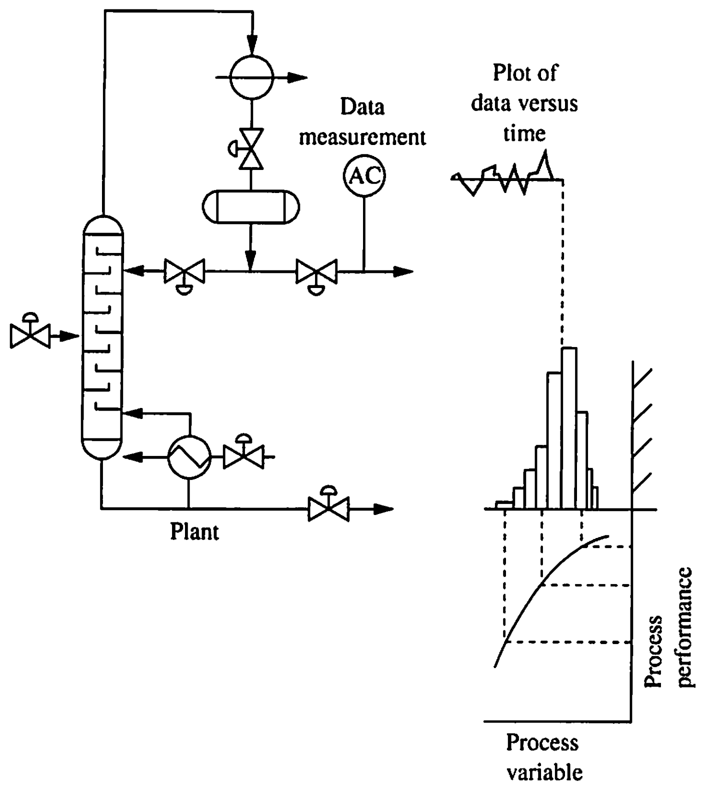

FIGURE 2.10

Schematic presentation of the method for calculating the average process performance from plant data.

This calculation is schematically shown in Figure 2.10. The calculation is tedious when done by hand but is performed easily with a spreadsheet or other computer program.

Note that methods for predicting how improved control affects the frequency distribution require technology covered in Part III of the book. These methods require a sound understanding of process dynamic responses and typical control calculations. For now, we will assume that the improved frequency distribution can be predicted.

EXAMPLE 2.1.

This example presents data for a reactor of the type shown in Figure 2.5. The reaction taking place is the pyrolysis of ethane to a wide range of products, one of which is the desired product, ethylene. The goal for this example is to maximize the conversion of feed ethane. This could be achieved by increasing the reactor temperature, but a hard constraint, the maximum temperature of 864°C, must not be exceeded, or damage will occur to the furnace. Control performance data is provided in Table 2.1.

In calculating benefits for control improvement, the calculation is performed twice. The first calculation uses the *base case* distribution, which represents the plant performance with poor control. The base case reactor temperature, shown as the top graph in Figure 2.11, might result from control via the plant operator occasionally adjusting the fuel flow. The second calculation uses the tighter distribution shown in the middle graph, which results from improved control using methods described in Parts III and IV. The process performance correlation, which is required to relate the temperature to conversion, is given in the bottom graph. The data for the graphs, along with the calculations for the averages, are given in Table 2.1.

The difference between the two average performances, a conversion increase of 4.4 percent, is the benefit for improved control. Note that the benefit is achieved by *reducing the variance* and *increasing the average temperature.* Both are required in this example; simply reducing variance with the same mean would not be a worthwhile achievement! Naturally, this benefit must be related to dollars and compared with the costs for equipment and personnel time when deciding whether this investment is justified. The economic benefit would be calculated as follows:

$$\Delta\text{profit} = (\text{feed flow})\ (\Delta\ \text{conversion})\ (\$/\text{kg products}) \tag{2.4}$$

In a typical ethylene plant, the benefits for even a small increase in conversion would be much greater than the costs. Additional benefits would result from fewer disturbances to downstream units and longer operating life of the fired heater due to reduced thermal stress.

EXAMPLE 2.2.

A second example is given for the boiler excess oxygen shown in Figure 2.6. The discussion in the previous section demonstrated that the profit is maximized when the excess oxygen is maintained slightly above the stoichiometric ratio, where the efficiency is at its maximum. Again, the process performance function, here efficiency, is used to evaluate each operating value, and frequency distributions are used to characterize the variation in performance.

The performance is calculated for the base case and an improved control case, and the benefit is calculated as shown in Figure 2.12 for an example with

TABLE 2.1

Frequency data for Example 2.1

Temperature midpoint (°C)	Conversion P_j (%)	Initial data F_j	Initial data P_j*F_j	Data with improved control F_j	Data with improved control P_j*F_j
842	50	0	0	0	0
844	51	0.0666	3.4	0	0
846	52	0.111	5.778	0	0
848	53	0.111	5.889	0	0
850	54	0.156	8.4	0	0
852	55	0.244	13.44	0	0
854	56	0.133	7.467	0	0
856	57	0.111	6.333	0	0
858	58	0.044	2.578	0.25	14.5
860	59	0.022	1.311	0.50	29.5
862	60	0	0	0.25	15
Average conversion (%) = $\sum P_j * F_j$ =			54.6		59

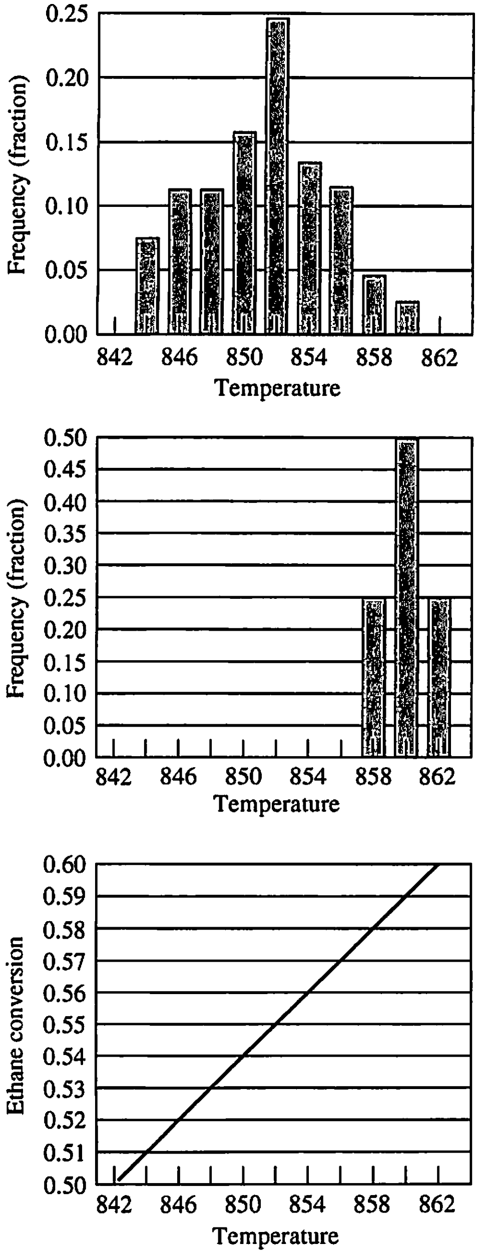

FIGURE 2.11

Data for Example 2.1 in which the benefits of reduced variation and closer approach to the maximum temperature limit in a chemical reactor are calculated.

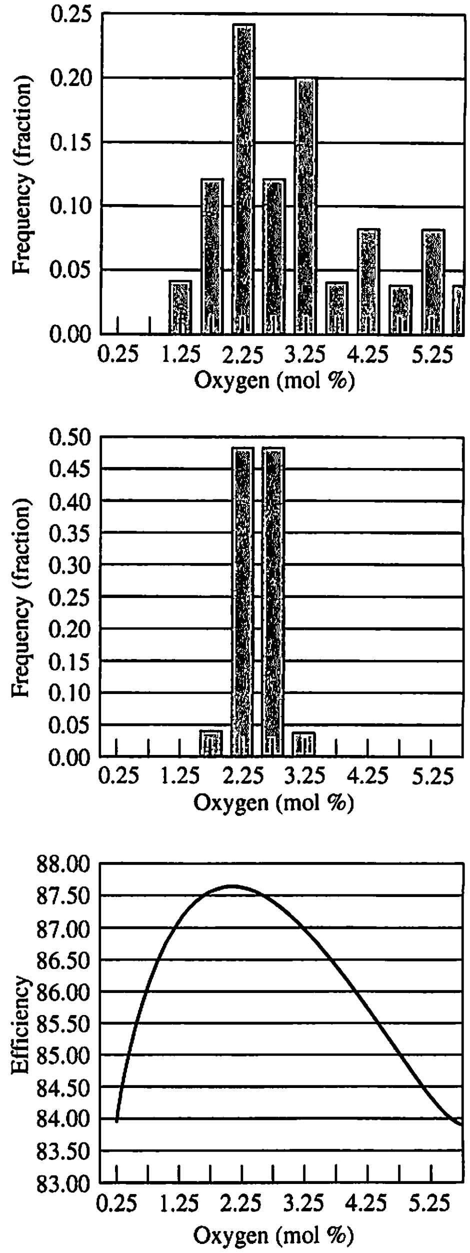

FIGURE 2.12

Data for Example 2.2 in which the benefits of reducing the variation of excess oxygen in boiler flue gas are calculated.

realistic data. The data for the graphs, along with the calculations for the averages, are given in Table 2.2. The average efficiency increased by almost 1 percent with better control and would be related to profit as follows:

$$\Delta\text{profit} = (\Delta\ \text{efficiency}/100)\ (\text{steam flow})\ (\Delta H_{\text{vap}})\ (\$/\text{energy}) \qquad (2.5)$$

This improvement would result in fuel savings worth tens of thousands of dollars per year in a typical industrial boiler. In this case, the average of the process variable (excess oxygen) is the same for the initial and improved operations, because the improvement is *due entirely to the reduction in the variance* of the excess oxygen. The difference between the chemical reactor and the boiler results from the different process performance curves. Note that the improved control case has its desired value at an excess oxygen value slightly greater than where the maximum profit occurs, so that the chance of a dangerous condition is negligibly small.

A few important assumptions in this benefits calculation method may not be obvious, so they are discussed here. First, the frequency distributions can never be *guaranteed* to remain within the operating window. If a large enough data set were collected, some data would be outside of the operating window due to infrequent, large disturbances. Therefore, some small probability of exceeding the constraints always exists and must be accepted. For soft constraints, it is common to select an average value so that no more than a few percent of the data exceeds the constraint; often the target is two standard deviations from the limit. For important hard constraints, an average much farther from the constraint can be selected, since the emergency system will activate each time the system reaches a boundary.

A second assumption concerns the mixing of steady-state and dynamic relationships. Remember that the process performance function is developed from steady-state analysis. The frequency distribution is calculated from plant data, which is inherently dynamic. Therefore, the two correlations cannot strictly be used together, as they are in equation (2.3). The difficulty is circumvented if the plant is assumed to have operated at quasi-steady state at each data point, then varied to the next quasi-steady state for the subsequent data point. When this assumption is valid, the plant data is essentially from a series of steady-state operations, and equation (2.3) is valid, because all data and correlations are consistently steady-state.

TABLE 2.2

Frequency data for Example 2.2

		Initial data		Data with improved control	
Excess oxygen midpoint (mol fraction)	**Boiler efficiency P_j (%)**	F_j	$P_j{*}F_j$	F_j	$P_j{*}F_j$
0.25	83.88	0	0	0	0
0.75	85.70	0	0	0	0
1.25	86.85	0.04	3.47	0	0
1.75	87.50	0.12	10.50	0.250	2.19
2.25	87.70	0.24	21.05	0.475	41.66
2.75	87.54	0.12	10.50	0.475	41.58
3.25	87.10	0.20	17.42	0.025	2.18
3.75	86.48	0.04	3.46	0	0
4.25	85.76	0.08	6.86	0	0
4.75	85.02	0.04	3.40	0	0
5.25	84.36	0.08	6.75	0	0
5.75	83.86	0.04	3.35	0	0
Average efficiency (%) $= \sum P_j * F_j =$			86.77		87.70

Third, the approach is valid for modifying the behavior of one process variable, with all other variables unchanged. If many control strategies are to be evaluated, the interaction among them must be considered. The alterations to the procedure depend on the specific plant considered but would normally require a model of the integrated plant.

> The analysis method presented in this section demonstrates that information on the *variability* of key variables is required for evaluating the performance of a process—average values of process variables are *not adequate*.

The method explained in this section clearly demonstrates the importance of understanding the goals of the plant prior to evaluating and designing the control strategies. It also shows the importance of reducing the variation in achieving good plant operation and is a practical way to perform economic evaluations of potential investments.

2.5 ▫ IMPORTANCE OF CONTROL ENGINEERING

Good control performance yields substantial benefits for safe and profitable plant operation. By applying the process control principles in this book, the engineer will be able to design plants and control strategies that achieve the control objectives. Recapitulating the material in Chapter 1, control engineering facilitates good control by ensuring that the following criteria are satisfied.

Control Is Possible

The plant must be designed with control strategies in mind so that the appropriate measurements and manipulated variables exist. Control of the composition of the liquid product from the flash drum in Figure 2.2 requires the flexibility to adjust the valves in the heating streams. Even if the valve can be adjusted, the total heat exchanger areas and utility flows must be large enough to satisfy the demands of the flash process. Thus, the chemical engineer is responsible for ensuring that the process equipment and control equipment provide sufficient flexibility.

The Plant Is Easy to Control

Clearly, reduction in variation is desired. Typically, plants that are subject to few disturbances, due to inventory (buffer) between the disturbance and the controlled variable, are easier to control. Unfortunately, this is contradictory to many modern designs, which include energy-saving heat integration schemes and reduced plant inventories. Therefore, the dynamic analysis of such designs is important to determine how much (undesired) variance results from the (desired) lower capital costs and higher steady-state efficiency. Also, the plant should be "responsive"; that is, the dynamics between the manipulated and controlled variables should be fast—the faster the better. Plant design can influence this important factor substantially.

Proper Control Calculations Are Used

Properly designed control calculations can improve the control performance by reducing the variation of the controlled variable. Some of the desired characteristics for these calculations are simplicity, generality, reliability, and flexibility. The basic control algorithm is introduced in Chapter 8.

Control Equipment Is Properly Selected

Equipment for process control involves considerable cost and must be selected carefully to avoid wasteful excess equipment. Information on equipment cost can be obtained from the references in Chapter 1.

EXAMPLE 2.3.

Control performance depends on process and control equipment design. The plant section in Figure 2.13*a* and *b* includes different designs for a packed-bed chemical reactor and two distillation towers. The feed to the plant section experiences composition variation, which results in variation in the product composition, which should be maintained as constant as possible.

The lower-cost plant design in Figure 2.13*a* has no extra tankage and a low-cost analyzer that must be placed after the distillation towers. The more costly design has a feed tank, to reduce the effects of the feed compositions through mixing, and a more expensive analyzer located at the outlet of the reactor for faster sensing. Thus, the design in Figure 2.13*b* has smaller disturbances to the reactor and faster control. The dynamic responses show that the control performance of the more costly plant is much better. Whether the investment is justified requires an economic analysis of the entire plant. As this example demonstrates, good control engineering involves proper equipment design as well as control calculations.

EXAMPLE 2.4.

Control contributes to safety by maintaining process variables near their desired values. The chemical reactor with highly exothermic reaction in Figure 2.14 demonstrates two examples of safety through control. Many input variables, such as feed composition, feed temperature, and cooling temperature, can vary, which could lead to dangerous overflow of the liquid and large temperature excursions (runaway). The control design shown in Figure 2.14 maintains the level near its desired value by adjusting the outlet flow rate, and it maintains the temperature near its desired value by adjusting the coolant flow rate. If required, these controls could be supplemented with emergency control systems.

EXAMPLE 2.5.

The type of control calculation can affect the dynamic performance of the process. Consider the system in Figure 2.15*a* through *c*, which has three different control designs, each giving a different control performance. The process involves mixing two streams to achieve a desired concentration in the exit stream by adjusting one of the inlet streams. The first design, in Figure 2.15*a*, gives the result of a very simple feedback control calculation, which keeps the controlled variable from varying too far from but does not return the controlled variable to the desired value; this deviation is termed *offset* and is generally undesirable. The second design, in Figure 2.15*b*, uses a more complex feedback control calculation, which provides

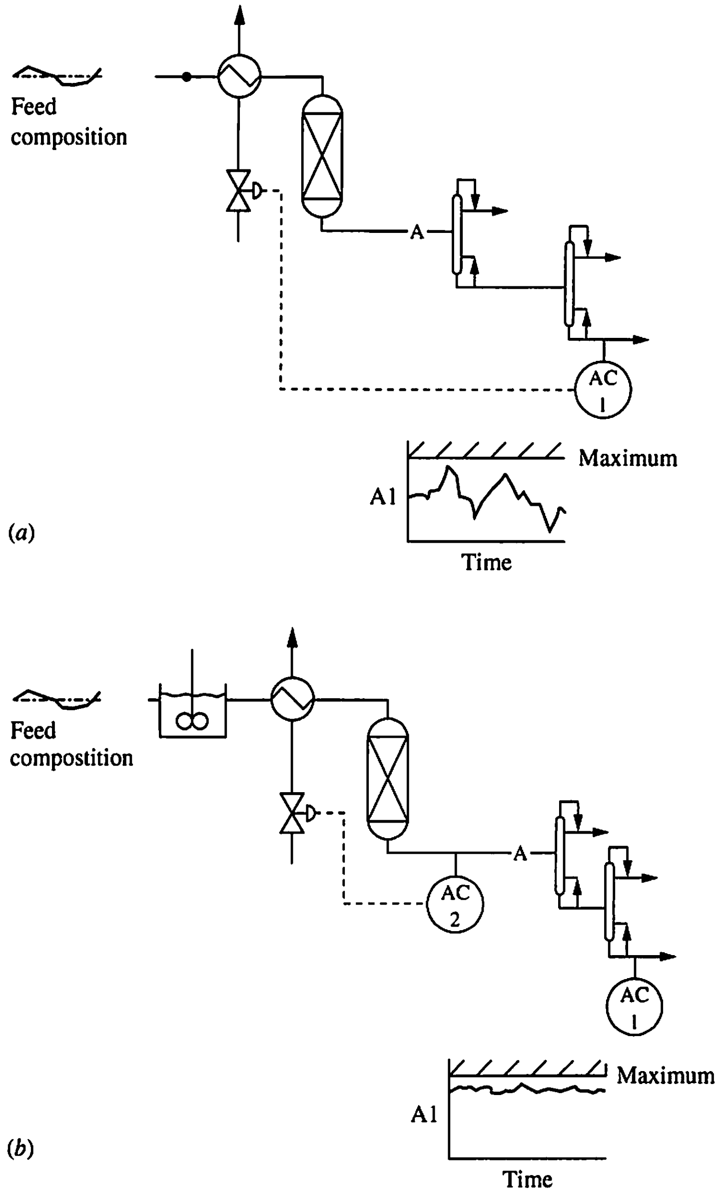

FIGURE 2.13

(*a*) Example of a process design that is difficult to control. (*b*) Example of a process that is easier to control.

response to disturbances that returns the controlled variable to its desired value. Since the second design relies on feedback principles, the controlled variable experiences a rather large initial deviation, which cannot be reduced by improved feedback calculations. The third design combines feedback with a predicted correction based on a measurement of the disturbance, which is called *feedforward*. The third design provides even better performance by reducing the magnitude of the initial response along with a return to the desired value. The calculations used for these designs, along with criteria for selecting among possible designs, are covered in later chapters. This example simply demonstrates that the type of calculation can substantially affect the dynamic response of a control system.

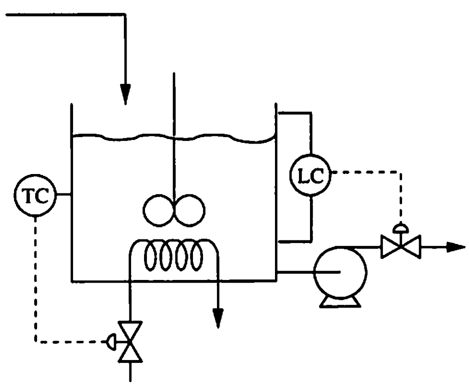

FIGURE 2.14

Control for stirred-tank reactor.

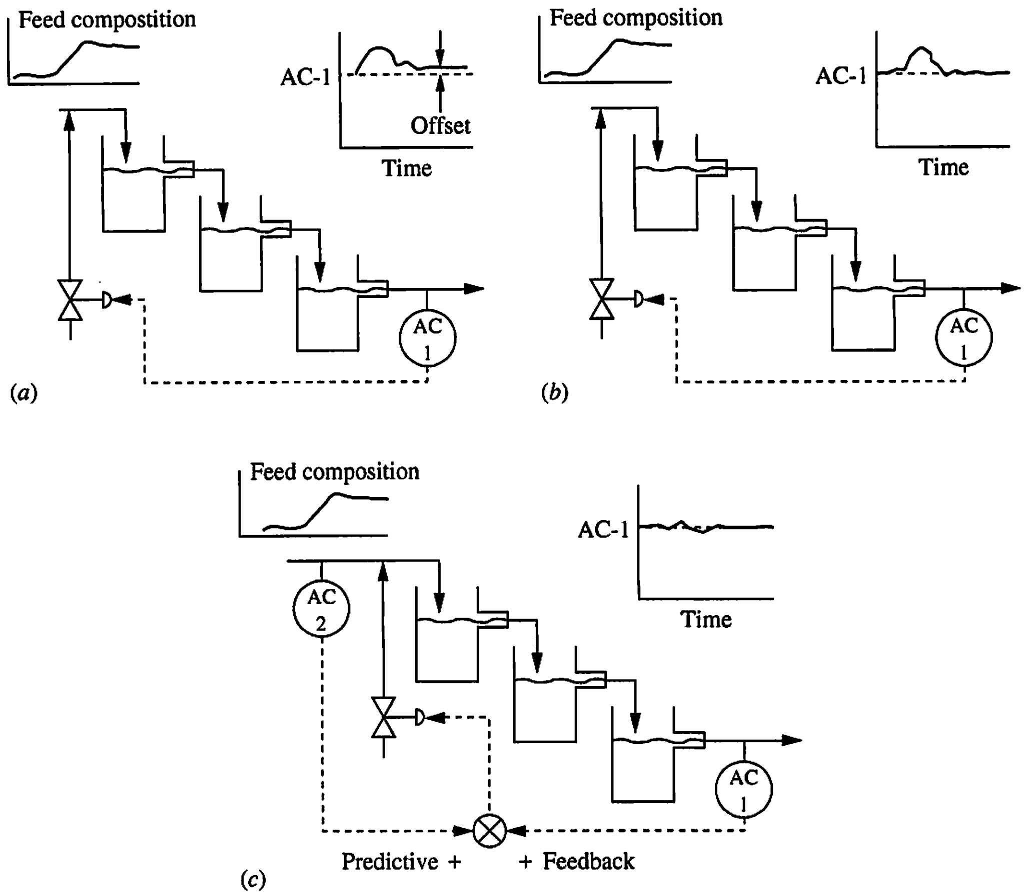

FIGURE 2.15

2.6 ■ CONCLUSIONS

Good control design addresses a hierarchy of control objectives, ranging from safety to product quality and profit, which depend on the operating objectives for the plant. The objectives are determined by both steady-state and dynamic analysis of the plant performance. The steady-state feasible operating region is defined by the operating window; plant operation should remain within the window, because constraint violations involve severe penalties. Within the operating window, the condition that results in the highest profit is theoretically the best operation. However, because the plant cannot be maintained at an exact value of each variable due to disturbances, variation must be considered in selecting an operating point that does not result in (unacceptably frequent) constraint violations yet still achieves a high profit. Process control reduces the variation and results in consistently high product quality and close approach to the theoretical maximum profit. Methods for quantitatively analyzing these factors are presented in this chapter.

As we have learned, good performance provides "tight" control of key variables; that is, the variables vary only slightly from their desired values. Clearly, understanding the dynamic behavior of processes is essential in designing control strategies. Therefore, the next part of the book addresses process dynamics and modelling. Only with a thorough knowledge of the process dynamics can we design control calculations that meet demanding objectives and yield large benefits.

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ADDITIONAL RESOURCES

The following references provide guidance on performing benefits studies in industrial plants, and Marlin et al. (1987) gives details on studies in seven industrial plants.

Marlin, T., J. Perkins, G. Barton, and M. Brisk, *Advanced Process Control Applications—Opportunities and Benefits*, Instrument Society of America, Research Triangle Park, NC, 1987.

Shunta, J., *Achieving World Class Manufacturing through Process Control*, Prentice-Hall PTR, Englewood Cliffs, NJ, 1995.

For further examples of operating windows and how they are used in setting process operating policies, see

Arkun, Y., and M. Morari, "Studies in the Synthesis of Control Structures for Chemical Processes, Part IV," *AIChE J.*, *26*, 975–991 (1980).

Fisher, W., M. Doherty, and J. Douglas, "The Interface between Design and Control," *IEC Res.*, *27*, 597–615 (1988).

Maarleveld, A., and J. Rijnsdorp, "Constraint Control in Distillation Columns," *Automatica*, *6*, 51–58 (1970).

Morari, M., Y. Arkun, and G. Stephanopoulos, "Studies in the Synthesis of Control Structures for Chemical Processes, Part III," *AIChE J.*, *26*, 220 (1980).

Roffel, B., and H. Fontien, "Constraint Control of Distillation Processes," *Chem. Eng. Sci.*, *34*, 1007–1018 (1979).

These questions provide exercises in relating process variability to performance. Much of the remainder of the book addresses *how* process control can reduce the variability of key variables.

QUESTIONS

2.1. For each of the following processes, identify at least one control objective in each of the seven categories introduced in Section 2.2. Describe a feedback approach appropriate for achieving each objective.

(*a*) The reactor-separator system in Figure 1.8
(*b*) The boiler in Figure 14.17
(*c*) The distillation column in Figure 15.18
(*d*) The fired heater in Figure 17.17

2.2. The best distribution of variable values depends strongly on the performance function of the process. Three different performance functions are given in Figure Q2.2. In each case, the average value of the variable (x_{ave}) must remain at the specified value, although the distribution around the average is not specified. The performance function, P, can be assumed to be

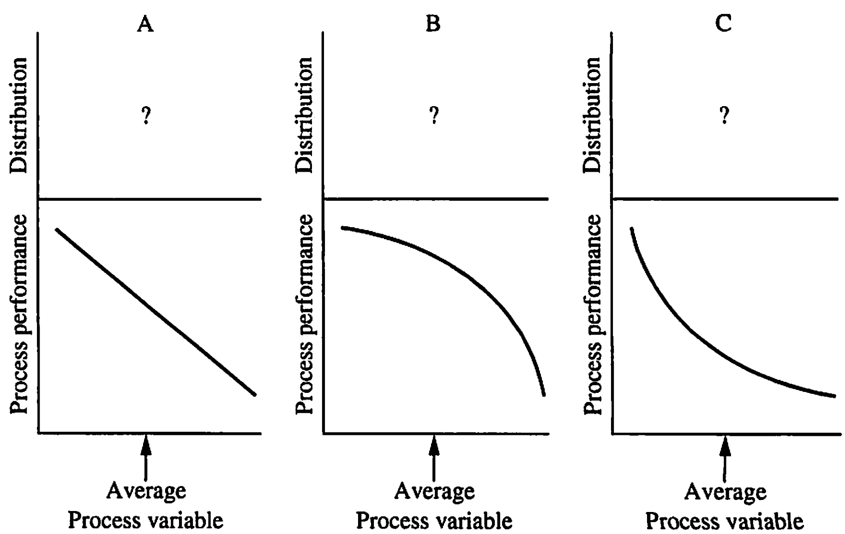

FIGURE Q2.2

a quadratic function of the variable, x, in every segment of the distribution.

$$P_i = a + b\,(x_i - x_{\text{ave}}) + c\,(x_i - x_{\text{ave}})^2$$

For each of the cases in Figure Q2.2, discuss the relationship between the distribution and the average profit, and determine the distribution that will maximize the average performance function. Provide quantitative justification for your result.

2.3. The fired heater example in Figure 2.11 had a hard constraint.

(*a*) Sketch the performance function for this situation, including the performance when violations occur, on the figure.

(*b*) Assume that the distribution of the temperature would have 0.005 fraction of its operation exceeding the limit of 864°C and that each time the limit is exceeded, the plant incurs a cost of $1,000 to restart the equipment. Can you calculate the total cost per year for exceeding the limit?

(*c*) Make any additional assumptions and complete the calculation.

2.4. Sometimes there is no active hard constraint. Assume that the fired heater in Figure 2.11 has no hard constraint, but that a side reaction forming undesired products begins to occur significantly at 850°C. This side reaction has an activation energy with larger magnitude than the product reaction. Sketch the shape of the performance function for this situation. How would you determine the best desired (average) value of the temperature and the best temperature distribution?

2.5. Sometimes engineers use a shortcut method for determining the average process performance. In this shortcut, the average variable value is used, rather than the full distribution, in calculating the performance. Discuss the assumptions implicit in this shortcut and when it is and is not appropriate.

2.6. A chemical plant produces vinyl chloride monomer for subsequent production of polyvinyl chloride. This plant can sell all monomer it can produce within quality specifications. Analysis indicates that the plant can produce 175 tons/day of monomer with perfect operation. A two-month production record is given in Figure Q2.6. Calculate the profit lost by not operating at the highest value possible. Discuss why the plant production might not always be at the highest possible value.

2.7. A blending process, shown in Figure Q2.7, mixes component A into a stream. The objective is to maximize the amount of A in the stream without exceeding the upper limit of the concentration of A, which is 2.2 mole/m^3. The current operation is "open-loop," with the operator occasionally looking at the analyzer value and changing the flow of A. The flow during the period that the data was collected was essentially constant at 1053 m^3/h. How much more A could have been blended into the stream with perfect control, that is, if the concentration of A had been maintained exactly at its maximum? What would be the improvement if the new distribution were normal with a standard deviation of 0.075 mole/m^3?

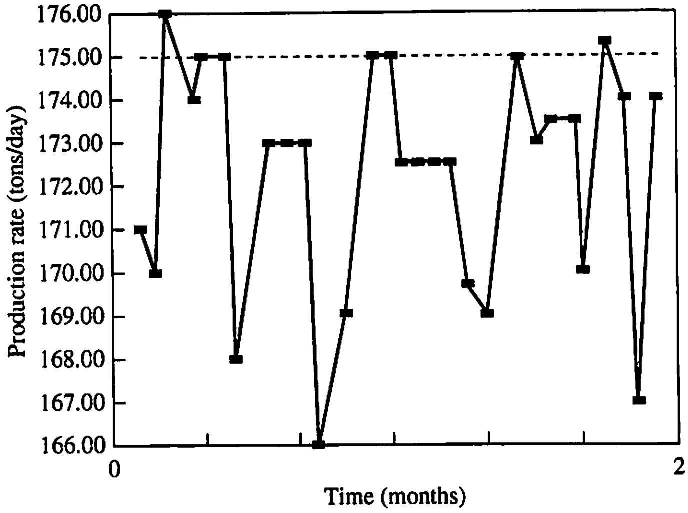

FIGURE Q2.6

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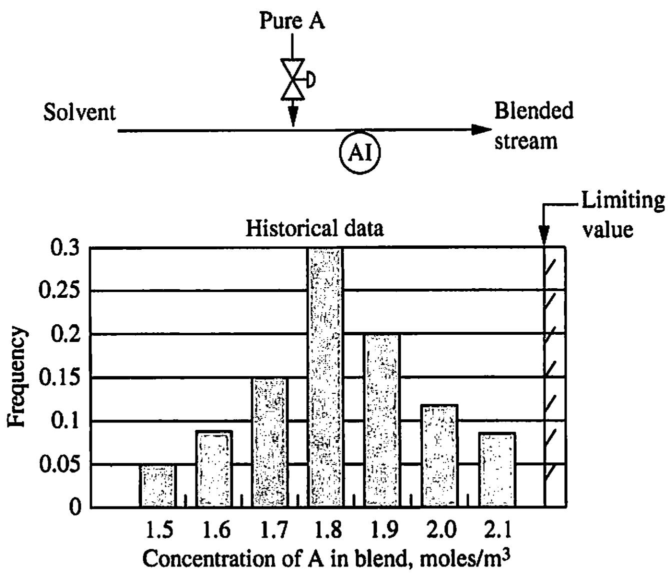

FIGURE Q2.7

2.8. The performance function for a distillation tower is given in Figure Q2.8 in terms of lost profit from the best operation as a function of the bottoms impurity, x_B (Stout and Cline, 1978). Calculate the average performance for the four distributions (A through D) given in Table Q2.8 along with the average and standard deviation of the concentration, x_B. *Discuss the* relationship between the distributions and the average performance.

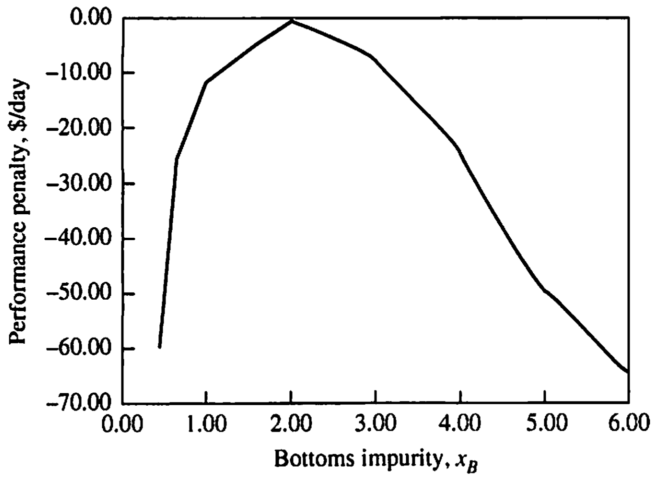

FIGURE Q2.8

(Reprinted by permission. Copyright ©1976, Instrument Society of America. From Stout, T., and R. Cline, "Control Systems Justification." *Instr. Techn.*, September 1976, pp. 51–58.)

TABLE Q2.8

	Fraction of time at x_B			
x_B	A	B	C	D
0.25	0	0	0	0
0.5	0.25	0.05	0	0
0.75	0.50	0.05	0	0
1.0	0.25	0.10	0	0
1.5	0	0.20	0	0.333
2.0	0	0.30	0	0.333
3.0	0	0.20	0.25	0.333
4.0	0	0.10	0.50	0
5.0	0	0	0.25	0
6.0	0	0	0	0

2.9. Profit contours similar to those in Figure Q2.9 have been reported by Gorzinski (1983) for a distillation tower separating normal butane and isobutane in an alkylation process for a petroleum refinery. Based on the shape of the profit contours, discuss the selection of desired values for the distillate and bottoms impurity variables to be used in an automation strategy. (Recall that some variation about the desired values is inevitable.) If only one product purity can be controlled tightly to its desired value, which would be the one you would select to control tightly?

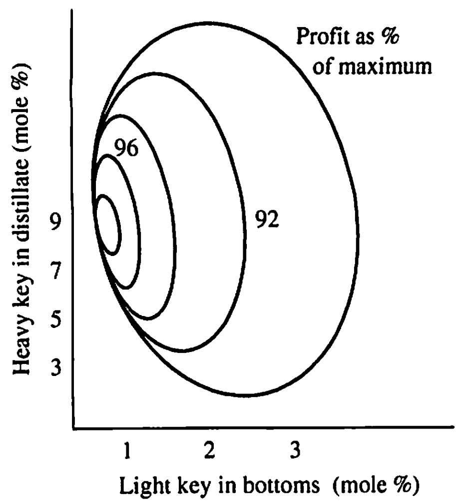

FIGURE Q2.9