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Adapting Single-Loop Control Systems for Nonlinear Processes

CHAPTER 16

16.1 ■ INTRODUCTION

Linear control theory provides methods for the analysis and design of many successful control strategies. Control systems based on these linear methods are generally successful in the process industries because (1) the control system maintains the process in a small range of operating variables, (2) many processes are not highly nonlinear, and (3) most control algorithms and designs are not sensitive to reasonable (±20%) model errors due to nonlinearities. These three conditions are satisfied for many processes, but they are not satisfied by all; therefore, the control of nonlinear processes must be addressed.

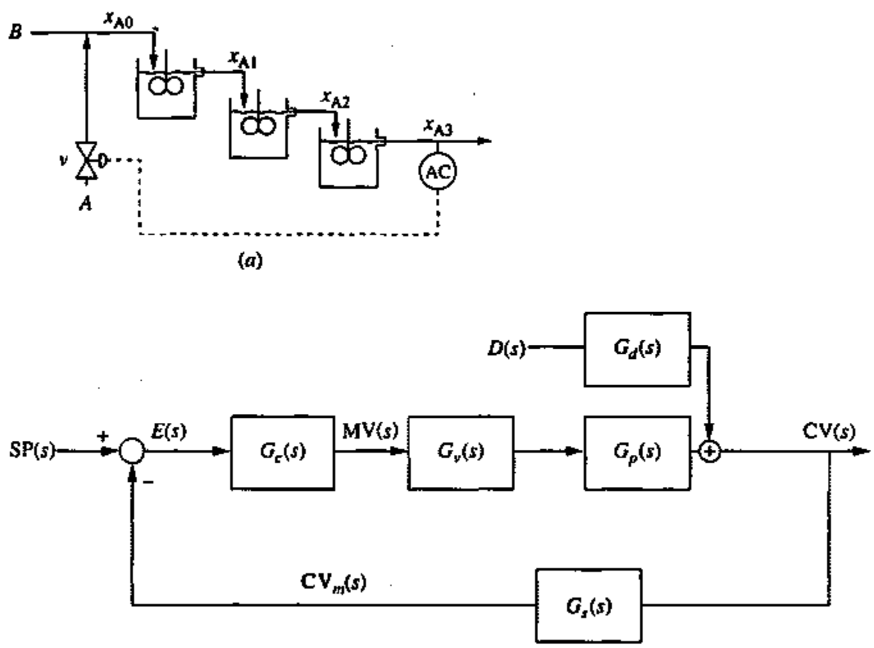
It is possible that the response of a nonlinear system could give better performance than a linear system and, therefore, a nonlinear control calculation might be better than any linear algorithm. However, there is no recognized, *general* nonlinear control theory that has been widely applied in the process industries. (An example of a nonlinear algorithm applied to level control is given in Chapter 18.) Therefore, the goal of the approaches in this chapter is to attain the performance achieved with a well-tuned linear controller. To reach this goal, the control methods in this chapter attempt to achieve a system that has a linear closed-loop relationship. If an element in the control loop is nonlinear, the approach applied here is to introduce a *compensating nonlinearity*, so that the overall closed-loop system behaves approximately linearly. This compensating nonlinearity may be introduced in the control algorithm or in physical equipment, such as a sensor or final element.

The next section begins the analysis by introducing a method for determining when nonlinearities significantly affect a control system. This analysis is extended to evaluate the proper fixed set of tuning constants for a linear PID controller applied to a nonlinear process. If a fixed set of tuning constants and linear instrumentation are not satisfactory, improvements can be achieved by adapting either the control calculation or the equipment responses. First, a common method for adapting the controller tuning in real time to compensate for nonlinearities is presented. Then the same concept is applied to introduce compensating nonlinearities in selected instrumentation, such as the control valve, to improve performance.

16.2 ■ ANALYZING A NONLINEAR PROCESS WITH LINEAR FEEDBACK CONTROL

A relatively simple process is analyzed in this section so that analytical models can be derived; the general approach is applicable to more complex processes. The process is shown in Figure 16.1*a*, which is the three-tank mixer considered in Examples 7.2 and 9.2. The outlet concentration of the last tank is controlled by adjusting the addition of component A to the feed to the first reactor. The equations describing the system are derived in Example 7.2 and summarized as follows:

$$x_{A0} = \frac{F_B(x_A)_B + F_A(x_A)_A}{F_B + F_A} \quad \text{with } F_A = K_v v \tag{16.1}$$



Transfer Functions	**Variables**
$G_c(s)$ = Controller	CV(s) = Controlled variable
$G_v(s)$ = Transmission, transducer, and valve	$\text{CV}_m(s)$ = Measured value of controlled variable
$G_p(s)$ = Process	D(s) = Disturbance
$G_s(s)$ = Sensor, transducer, and transmission	E(s) = Error
$G_d(s)$ = Disturbance	SP(s) = Set point

(b)

FIGURE 16.1

Mixing process: (a) schematic; (b) control system block diagram.

$$V_i \frac{dx_{Ai}}{dt} = (F_A + F_B)(x_{A(i-1)} - x_{Ai}) \qquad \text{for } i = 1, 3 \tag{16.2}$$

Note that the differential equations are nonlinear. We can linearize these equations, express the variables as deviations from the initial steady state, and take the Laplace transforms to yield the transfer function model:

$$\frac{x_{A3}}{\mathrm{SP}(s)} = \frac{G_p(s)G_v(s)G_c(s)}{1 + G_p(s)G_v(s)G_c(s)G_s(s)} \approx \frac{G_p(s)G_c(s)}{1 + G_{OL}(s)} \tag{16.3}$$

where the valve transfer function is a constant lumped into $G_p(s)$ and the sensor $G_s(s) = 1.0$.

$$\begin{aligned} G_{OL}(s) &= G_p(s)G_v(s)G_c(s)G_s(s) \approx G_p(s)G_c(s) \\ &= \frac{K_p}{(\tau s + 1)^3} G_c(s) \end{aligned} \tag{16.4}$$

where $$K_p = K_v \left[\frac{F_{Bs}(x_{AA} - x_{AB})_s}{(F_{As} + F_{Bs})^2} \right] \qquad K_v = 0.0028 \tag{16.5}$$

$$\tau = \frac{V}{F_{Bs} + F_{As}} \tag{16.6}$$

This linearized model clearly demonstrates how the gain and time constants depend on the volumes, total flow, and compositions. We will consider the response of the system for various values of one operating variable, the total flow rate $(F_A + F_B)$, which has the greatest variability for the situation considered here. In the scenario, the production rate changes periodically and remains nearly constant for a long time (relative to the feedback dynamics) at each production rate. The process dynamics are summarized in Table 16.1 for the range of flow variability (i.e., production rates) expected. The variation in the process dynamics due to the nonlinearity is not randomly distributed, because in this example the effect of an increase in flow rate is to decrease the process gain and time constants concurrently. This type of correlation is typical for nonlinear processes and demonstrates the need for careful analysis of the dynamic responses at different operating conditions.

TABLE 16.1

Summary of process dynamics and tuning for the three-tank mixing process*

Case	Process parameters			Controller parameters		
	F_B	K_p	τ	K_c	T_I	T_d
A	3.0	0.087	11.4	13.8	25.1	1.82
B	4.0	0.064	8.6	18.6	19.0	1.4
C	5.0	0.052	6.9	23.1	15.2	1.10
D	6.0	0.043	5.7	27.9	12.7	0.92
E	6.9	0.039	5.0	30.0	11.0	0.80

*For the combinations of process dynamics and tuning in this table, the gain margin for each case is 1.7.

Naturally, other factors, such as incorrect data in fundamental models or m... in empirical models, contribute to the modelling errors, but change in operating conditions is often the dominant factor causing the difference between models and true process behavior.

> The values in Table 16.1 demonstrate that the changes in dynamic model parameters due to nonlinearity can be *highly correlated*.

The effects of nonlinearity on two important system characteristics, stability and performance, are now investigated. As demonstrated in Chapter 10, the process dynamics influence the stability of a closed-loop system, and to achieve a stable control system the tuning parameters are adjusted to be compatible with the process dynamics. The PID feedback controller tuning for this process has been determined for five different flow rates that span the expected range of operation. (Note that Case E in Table 16.1 is the same as Example 9.2.) The tuning was determined by evaluating the process reaction curve, fitting an approximate first-order-with-dead-time model, and using the Ciancone correlations. Similar trends would be obtained for other tuning methods such as Ziegler-Nichols. It is important to recognize that the tuning reported in Table 16.1 has a reasonable margin from the stability limit for each case. In fact, the gain margin for all cases is about 1.7. The results in the table clearly indicate that the values of "good" tuning constants change significantly, over 50%, for the range of process operating conditions considered. This analysis indicates that the nonlinearity is significant for the changes in flow considered in this scenario.

> Calculating the controller tuning for cases covering the range of dynamics occurring in the process provides a basis for determining whether the controller tuning should be adapted.

The simplest control design approach would be to use a *single set* of tuning constants for all the operating conditions. The results in Table 16.1 provide the basic information needed to decide whether to use this tuning approach. If the tuning constants were not very different, it would be concluded that either the nonlinearities are mild or the operating conditions do not change much from the base case. For either situation, a constant set of tuning constants, which could be taken as the average values, would yield good PID feedback control performance.

If the proper values of the tuning constants differ significantly, as they do in Table 16.1, further analysis is necessary. Recall that the tuning for each case was determined to give good dynamic response and a proper gain margin for the nominal process model in that case. The single set of tuning constants to be used for all process models in the table must provide acceptable (if not good) feedback control performance for all cases. Since the process dynamics change, the stability margin of the closed-loop system can change, and the closed-loop system can become highly oscillatory or unstable for an improper choice of fixed tuning. Since instability and severe oscillations are to be avoided, the overriding concern is maintaining a reasonable stability margin for all expected process dynamics.

To ensure that the control system with varying process dynamics performs acceptably over the expected range of operation, the *worst-case dynamics* must be identified. This worst case gives the poorest control performance under the feedback controller and is usually the closed-loop system closest to the stability limit. The Bode plots of $G_p(s)$ for three of the cases in Table 16.1 are given in Figure 16.2. The results show that Case A has the lowest critical frequency and the highest amplitude ratio at its critical frequency. This result conforms to our experience that processes with longer time constants are more difficult to control. Thus, Case A would be selected as the most difficult process operation, or the worst case, within the scenarios.

The Bode analysis of $G_p(s)$ is substantiated by the results in Table 16.1, which indicate that Case A has the least aggressive feedback controller, because the controller gain is smallest and integral time is largest. Applying the controller tuning from Case A would result in a stable system for all cases, albeit with poor performance for some cases. Using a more aggressive set of tuning constants, Case E, for example, would lead to good performance in some cases, but the closed-loop performance would be very poor, and perhaps unstable, for other cases.

Dynamic simulations of closed-loop systems with various tunings are shown in Figures 16.3 and 16.4. The results in Figure 16.3*a* and *b* give the dynamic responses of the closed-loop system, with controller tuning from Case A, for two

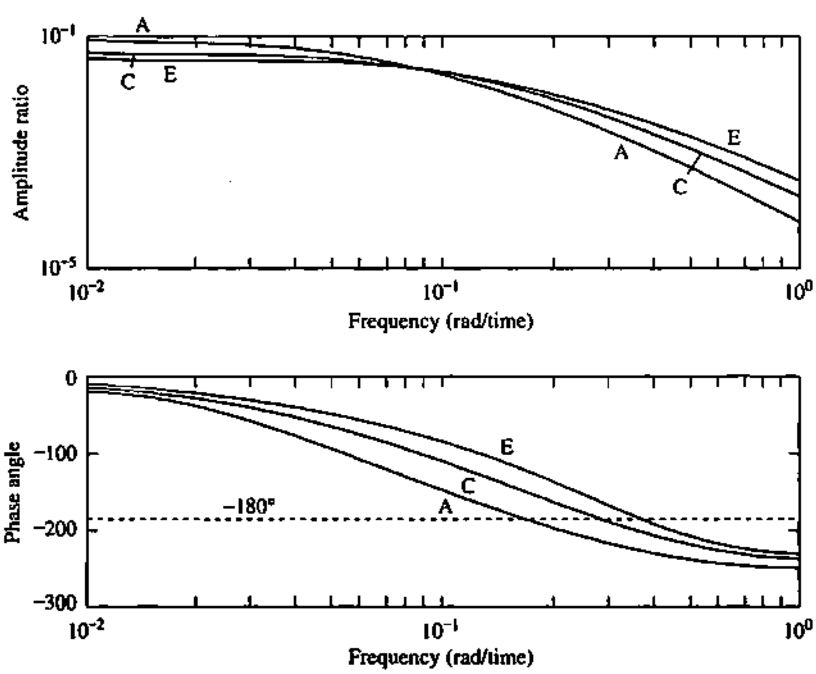


FIGURE 16.2

Bode plot for three-tank mixing system (cases defined in Table 16.1).

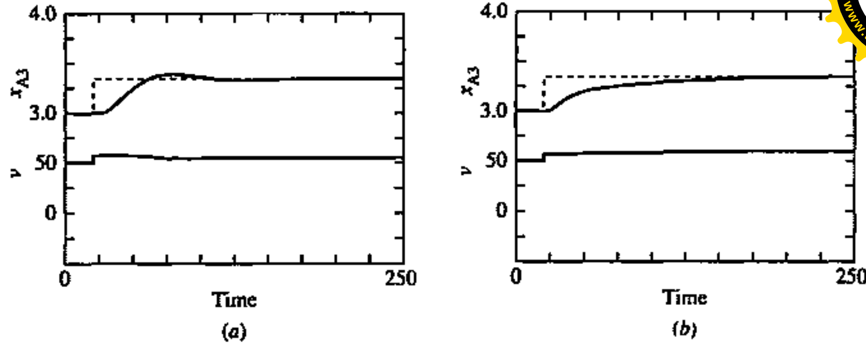


FIGURE 16.3

Dynamic respones for the mixing system with tuning from Case A ($K_c = 13.8$, $T_I = 25.1$, and $T_d = 1.82$). (*a*) Case A process dynamics, $F_B = 3$, gain margin $= 1.7$; (*b*) Case E process dynamics, $F_B = 6.9$, gain margin $= 4.5$.

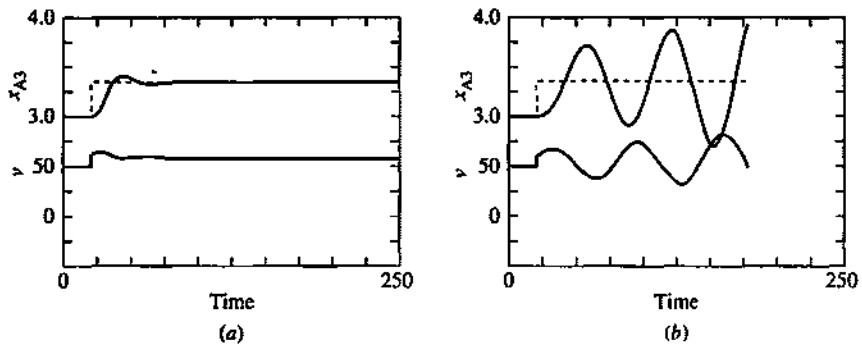


FIGURE 16.4

Dynamic respones for the mixing system with tuning from Case E ($K_c = 30.0$, $T_I = 11.0$, and $T_d = 0.80$): (*a*) Case E process dynamics, $F_B = 6.9$, gain margin $= 1.7$; (*b*) Case A process dynamics, $F_B = 3.0$, gain margin < 1.0, indicating instability.

different process dynamics. Note that the response, when controlling the plant with dynamics for Case A (the most difficult plant to control), is well behaved. The performance when controlling process E is rather poor, with a long time required to return to set point, but at least the response is stable.

The results in Figure 16.4*a* and *b* give the closed-loop dynamic responses for the controller tuning from Case E and the same two plant dynamics. Although the performance for the process dynamics from Case E is good, the performance for the process dynamics from Case A is unacceptable because the system is *unstable.* Since excessive oscillations and instability are to be avoided at all cost, the controller tuning used in Figure 16.4, based on the dynamics in Case E, is deemed unacceptable.

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When the feedback controller tuning constants are fixed and the process dynamics change, the fixed set of tuning constants selected should have the proper gain margin for the most difficult process dynamics in the range considered. This approach will ensure stability, but it may not provide satisfactory performance. For the example, the tuning from Case A is selected.

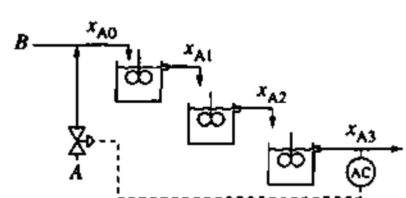
This section has presented a manner for determining whether nonlinearities significantly affect stability and control performance. The method is based on the tuning and stability analysis of the system linearized about various operating conditions. Also, a tuning selection criterion is given that is applicable when a fixed set of tuning constants is used as the process dynamics vary. The goal of this criterion is to provide the best possible control performance, with constant tuning, while preventing instability or excessive oscillations. The resulting control performance may be unacceptably poor, providing sluggish compensation for some cases; therefore, the next sections present common methods for improving performance, while preventing instability, by compensating for the nonlinearity.

16.3 ▣ IMPROVING NONLINEAR PROCESS PERFORMANCE THROUGH DETERMINISTIC CONTROL LOOP CALCULATIONS

The approach described in the previous section can lead to poor control performance for two reasons. First, some process operating conditions lead to poor performance because of increased feedback dynamics (e.g., longer dead time and time constants). Second, the fixed values for the feedback controller tuning constants are too "conservative" for some process operations. Clearly, one set of tuning values cannot prevent degradation in feedback control performance arising from the changes in plant dynamics. However, modifying the tuning to be compatible with the current process dynamics can maintain the feedback control performance close to the best possible with the PID algorithm for whatever plant dynamics exist.

The approach for modifying the controller tuning constants through deterministic calculations can be applied to improve the control of some nonlinear processes. The term "deterministic" is used to designate an unchanging relationship between the operating condition and the tuning constant values. The operating condition is determined by measuring a process variable that is directly related to the feedback dynamics. Then the control constants can be expressed as a function of this measured variable, PV, as shown in the following equation:

$$\mathrm{MV} = K_c(\mathrm{PV})\left[E + \frac{1}{T_I(\mathrm{PV})}\int_0^t E(t')\,dt' + T_d(\mathrm{PV})\frac{d\mathrm{CV}}{dt}\right] + I \qquad (16.7)$$



The resulting controller is *nonlinear.* The stability analysis presented in Chapter 10 can be applied to this system assuming that the value of PV in equation (16.7) changes slowly; that is, it has a much lower frequency than the closed-loop critical frequency. When this condition is satisfied, the tuning can be considered constant for the stability calculation.

This approach is demonstrated by applying it to the three-tank mixing system introduced in the previous section. The correlations between the tuning constants and the measured variable that indicates the change in process dynamics—in the example the flow—can be fitted by an equation or arranged in a lookup table. Equations for the example are given below for the range of operation in Table 16.1 [$3.1 < (F_A + F_B) < 7.0$] with the parameters determined by a least squares fit.

$$
\begin{aligned}
K_c &= -5.64 + 7.368(F_A + F_B) - 0.3135(F_A + F_B)^2 \\
T_I &= 50.37 - 10.626(F_A + F_B) + 0.7164(F_A + F_B)^2 \\
T_d &= 3.66 - 0.776(F_A + F_B) + 0.0525(F_A + F_B)^2
\end{aligned}
\tag{16.8}
$$

The controller using tuning calculated by equations (16.8) can be applied to the nonlinear mixing process. The resulting dynamic responses are essentially the same as in Figure 16.3*a* and Figure 16.4*a*. The control performance is good for the different flow rates because the tuning is modified to be compatible with the process dynamics. Note that the performance in Figure 16.4*a* is better than in Figure 16.3*a*, even though the controllers in both systems are well tuned, because the feedback dynamics in Case E are faster. Comparison with Figure 16.3*b* and Figure 16.4*b* demonstrates the potential performance advantage of this approach over maintaining the tuning constants at fixed values. The procedure introduced in this example is summarized in Table 16.2.

The use of controller tuning modifications described in this section is often referred to as ***gain scheduling***, because early applications adjusted only the controller gain. With digital computers, all tuning constants can be easily adjusted when required to achieve the desired control performance.

If adequate control performance is achieved through adapting only the controller gain and the controller gain should be proportional to feed flow, gain scheduling can be implemented as part of modified feedforward/feedback control design. An example is given in Figure 16.5*a* for the feedforward and feedback control of the simple mixing system. The model for the system is

$$
x_m = \frac{F_A x_A + F_B x_B}{F_A + F_B} \tag{16.9}
$$

The flows and compositions for this mixing process are assumed to be the same

TABLE 16.2

Criteria for the deterministic modification of controller tuning

Deterministic modification of tuning is appropriate when
1. Constant controller tuning values do not provide satisfactory control performance because of significant changes in operating conditions.
2. The nonlinearity can be predicted based on a process variable measured in real time.
3. The relationship between the measurement and the process dynamics can be determined either from a fundamental model or from empirically developed models.
4. The changes in the process dynamics are at a frequency much lower than the critical frequency of the control system.

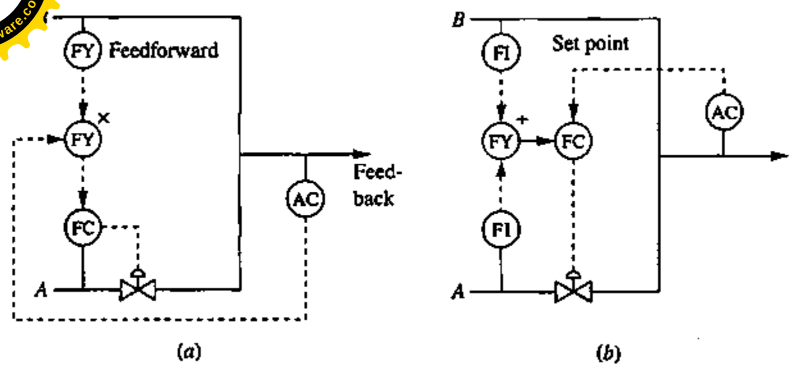


FIGURE 16.5

Examples of gain scheduling through feedforward control.

as the three-tank mixer. The steady-state equations describing the system are

$$G_p(s) = \frac{F_B[(x_A)_A - (x_A)_B]}{(F_A + F_B)^2} \approx \frac{K_p}{F_B}$$
$$G_{cff}(s) = F_B \qquad G_{cfb}(s) = K_c\left(1 + \frac{1}{T_I s}\right) \tag{16.10}$$

with K_p and K_c constant. The stability margin is determined from the Bode analysis by referring to $G_{OL}(s)$, which follows for the example:

$$\begin{aligned} G_{OL}(s) &= G_p(s) K_{cff} K_c \left(1 + \frac{1}{T_I s}\right) \\ &= \frac{K_p}{F_B} F_B K_c \left(1 + \frac{1}{T_I s}\right) \end{aligned} \tag{16.11}$$

To include the modification of the loop gain as indicated in equation (16.11), the outputs of the feedforward and feedback controllers are *multiplied*, rather than added as described in Chapter 15. Thus, as the feed flow increases, the effective gain of the feedback controller ($K_c F_B$) increases to compensate for the decrease in the process gain (K_p/F_B). Note that this design is an extension of the feedforward design shown in Figure 15.14*a* to include feedback and therefore retains the good disturbance response through feedforward control. This approach to controller gain modification is a simplification of the general approach described in Table 16.2.

16.4 ■ IMPROVING NONLINEAR PROCESS PERFORMANCE THROUGH CALCULATIONS OF THE MEASURED VARIABLE

In addition to the controller calculation, other elements in the control loop can also be modified in response to nonlinearities. Relationships between the sensor signal and the true process variable sometimes involve particularly simple nonlinearities that can be addressed by programmed calculations during the input processing phase of the control loop. Some examples are temperature (polynomial fit of thermocouple) and flow orifice (square root, density correction on ΔP). In addition to the linearization in the control loop, the availability of more accurate measured values for use in control and process monitoring is another important benefit of these calculations.

16.5 ▣ IMPROVING NONLINEAR PROCESS PERFORMANCE THROUGH FINAL ELEMENT SELECTION

Introducing a compensating nonlinearity in the control loop can be achieved by selecting appropriate control equipment to compensate for nonlinearities. The final control element, usually a valve, is the control loop element that is often modified in the process industries, because the modifications involve little cost. Again, the explanation in this section assumes that the desired closed-loop relationship is linear; if another relationship is required, the approach can be altered in a straightforward manner.

Since the valve is normally very fast relative to other elements in the control loop, only the gains of the elements are considered to vary. The linearized control system is shown in the block diagram in Figure 16.1*b*, and the loop gain for this system depends on the product of the individual gains.

$$\begin{aligned} G_{OL}(s) &= G_p(s)G_v(s)G_c(s)G_s(s) \\ &= K_pK_vK_cK_sG_p^*(s)G_v^*(s)G_c^*(s)G_s^*(s) \end{aligned} \tag{16.12}$$

where the gains (K_i) may be a function of operating conditions, and the dynamic elements of the transfer functions [$G_i^*(s)$ with $G_i^*(0) = 1.0$] do not change significantly with operating conditions. The manipulated variable in the majority of control loops is a valve stem position (v), also referred to as the *valve lift*, which influences a flow rate. The feedback system behaves as though it is linear if $G_{OL}(s)$ does not change as plant operating conditions change. Thus, linearity can be achieved, even if the process gain (K_p) changes, as long as the changes due to nonlinearities in the individual gains cancel. In this section, a method is described in which the valve nonlinearity is designed to cancel an undesirable process nonlinearity, with the controller (K_c) and sensor (K_s) gains assumed to be constant.

The final element selection is introduced through an example of flow control. The relationship between the controller output and the flow is often desired to be linear, so that the control system is linear. The relationship between the valve stem position and the flow is given below (Foust et al., 1960; Hutchinson, 1976).

$$F = F_{max}\frac{C_v(v)}{100}\sqrt{\frac{\Delta P_v}{\rho}} \tag{16.13}$$

where
F = flow
F_{max} = maximum flow through system with valve fully open
$C_v(v)$ = inherent valve characteristic, which is a function of v
v = valve stem position (% open or closed)
ΔP_v = pressure difference across the valve
ρ = fluid density

This is simply the expression for the flow through a restriction, with the variable v representing the valve stem position expressed in percent. The driving force for the flow is the difference between the pressures immediately before and after the valve, ΔP_v. The factor C_v is called the *inherent valve characteristic* and represents the percentage of maximum flow at any given valve stem position at a *constant pressure drop*, usually the design value. The C_v is a function of the valve design, basically the size and shape of the opening and plug, which can be linear or any of a selection of standard nonlinear relationships at the choice of the engineer. Three common inherent valve characteristics are shown in Figure 16.6.

In the typical process design, the pressures just before and after the valve change as the flow rate changes, as shown in Figure 16.7 (Quance, 1979). Typically, the pressure at the pump outlet is not constant; it decreases as the flow through the pump increases (Labanoff and Ross, 1985; Karassik and McGuire, 1998). Also, the pressure drop from the valve to the pipe outlet increases as the flow increases. For the example process in Figure 16.7, the pressure drop from the valve outlet to the end of the pipe could be calculated from the energy balance on the fluid, with losses determined from friction factor correlations (Foust et al., 1960):

$$P_2 = P_{\text{out}} + \Delta P_e + \sum_{i=1}^{2} \Delta P_{\text{Hx}i} + \Delta P_{\text{pipe}} + \Delta P_{\text{fit}} \qquad (16.14)$$

where P_{out} = outlet pressure (constant in this example)
ΔP_e = pressure drop due to change in elevation
$\Delta P_{\text{Hx}i}(F)$ = pressure drop due to heat exchangers ($i = 1, 2$)
$\Delta P_{\text{pipe}}(F)$ = pressure drop in the pipe due to skin friction
$\Delta P_{\text{fit}}(F)$ = pressure drop in elbows and expansions due to form friction

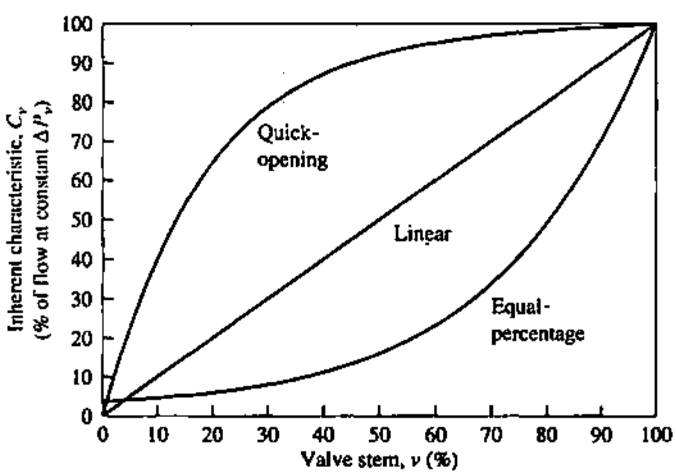


FIGURE 16.6

Three standard control valve inherent characteristics. (Reprinted by permission. Copyright ©1976, Instrument Society of America. From Hutchinson, J., ed., *ISA Handbook of Control Valves*, 2nd ed.)

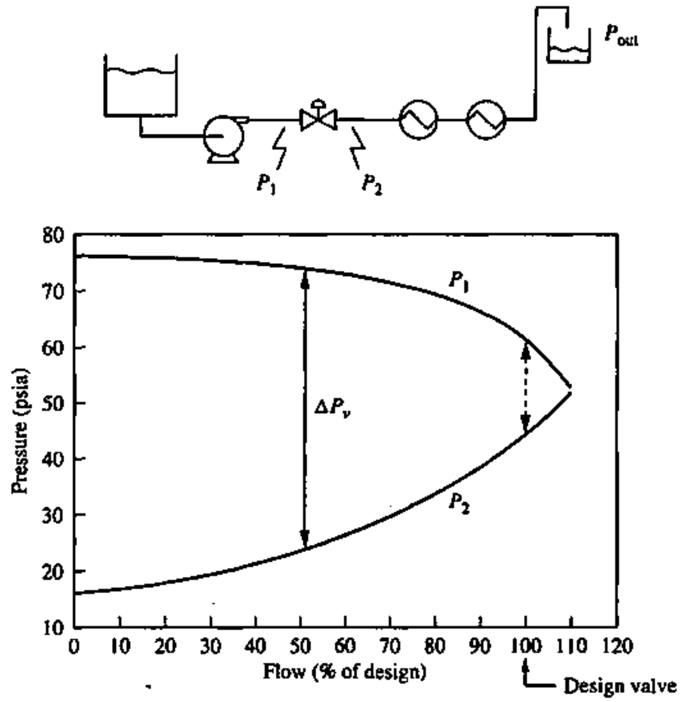


FIGURE 16.7

Relationship between pressures and flow for a typical system. (Reprinted by permission. Copyright ©1979, Instrument Society of America. From Quance, R., "Collecting Data for Control Valve Sizing," *In. Tech., 55*.)

Note that the last three pressure drop terms are functions of the flow rate (F). Due to the functional relationships for P_1 and P_2, the pressure drop across the valve ($\Delta P_v = P_1 - P_2$), decreases as the flow increases. This demonstrates that only part of the total pressure drop from the pump to the outlet is due to the valve; a considerable amount of the pressure drop is due to other frictional losses.

The goal of a linear system—a constant closed-loop relationship, $G_{OL}(s)$—is achieved when the relationship between the controller output and controlled variable is linear. In the case of flow control, the controller output can be taken to be the valve position, and the controlled variable is the measured flow rate. Since the pressure drop across the valve shown in Figure 16.7 is not constant, the relationship between the controller output and the valve opening must introduce a compensating nonlinearity for the overall gain to be constant. The nonlinearity can be introduced at low cost by selecting the appropriate valve characteristic C_v. The typical nonlinearity applied for situations similar to Figure 16.7 is the equal-percentage characteristic curve shown in Figure 16.6. The use of an equal-percentage valve in a process similar to that shown in Figure 16.7, in which the pressure drop decreases with flow, usually results in an approximately linear relationship between the valve stem position and flow. An experimental investigation of the application of an equal-percentage valve for the process described in Figure 16.7 resulted in the desired linear relationship shown in Figure 16.8. Note that the

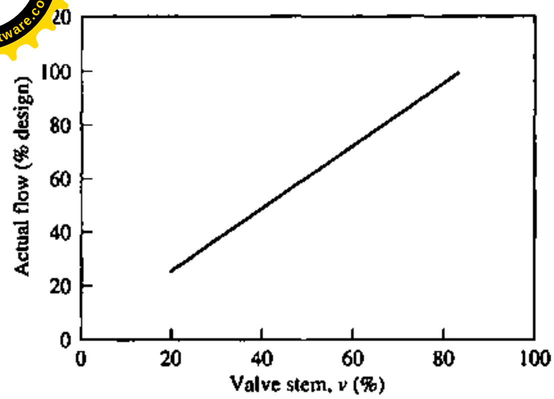


FIGURE 16.8

Linear overall relationship between v and flow for the example flow process with an equal-percentage valve. (Reprinted by permission. Copyright ©1979, Instrument Society of America. From Quance, R., "Collecting Data for Control Valve Sizing," *In. Tech.*, 55.)

valve is about 85 percent open at the design flow; this is a bit high, since most designers specify that the valve be about 70 to 80 percent open at the design flow. In all cases, the valve opening at design must be such that the required maximum flow can be achieved when the valve is fully open.

The general method employed for linearizing the flow control loop is now extended to a more complex process: a stirred-tank heat exchanger, shown in Figure 16.9. The energy balance for this system was derived in Example 3.7 and used in Example 8.5, and this example uses the same design parameters. The model is repeated here:

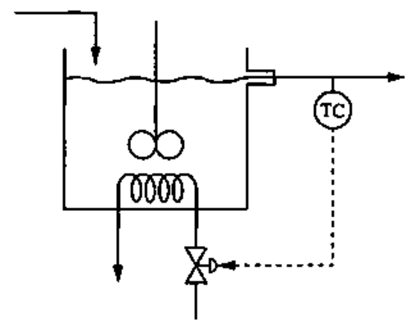


FIGURE 16.9

Heat exchanger control system.

$$V\rho C_p \frac{dT}{dt} = F\rho C_p(T_0 - T) - \frac{aF_c^{b+1}}{F_c + \dfrac{aF_c^b}{2\rho_c C_{pc}}}(T - T_{cin}) \tag{16.15}$$

$$F_c = F_{max}\left(\frac{C_v}{100}\sqrt{\frac{\Delta P}{\rho_c}}\right)v \tag{16.16}$$

In this example the pressure drop across the valve is assumed constant so that the analysis will highlight the effects of other process nonlinearities; however, if this were not the case, the same approach could be used, with an appropriate model for the coolant pressures included. This model can be used to evaluate the linearity of the steady-state process by calculating the steady-state value of the temperature at various coolant flow rates by setting $dT/dt = 0$. The results of this calculation are given in Figure 16.10*a*. This plot clearly shows the nonlinearity in the process gain, which changes by a factor of more than 5 over the range of operation considered.

The goal of compensation would be to achieve a linear relationship between the controller output and the temperature. The proper linear relationship would be

$$(K_p)_{ave} = \frac{\Delta T}{\Delta v} \approx \frac{79.8 - 122.5°\text{C}}{100 - 5\%} = -0.45\frac{°\text{C}}{\%} \tag{16.17}$$

which is the total change in temperature over the total change in valve position over the controllable range. To maintain the loop gain at this value while the process gain changes, the valve characteristic must change. From the value of the desired average gain in equation (16.17) and the process gain, $\Delta T/\Delta F_c$ in Figure 16.10*a*, the value of the characteristic can be evaluated as $C_v = (K_p)_{ave}/K_p$. The results of this calculation are given in Figure 16.10*b*. As expected, the plot of C_v versus valve stem has a slope with small magnitude where the process has a gain with large magnitude (i.e., at low coolant flows). Also, the figures show that the process gain, and therefore the slope of C_v versus valve stem, is nearly constant over the higher range of flows. The steady-state behavior of the system with the linearizing C_v installed gives Figure 16.10*c*; the linear relationship between the controller output and temperature indicates that a PID feedback controller with constant tuning parameters would be adequate.

It is important to understand the approach just demonstrated through these flow and heat exchanger examples: therefore, it is summarized in Table 16.3. The correct application of the procedure in this table frequently, but not always, results in an equal-percentage characteristic. An example of an exception occurs when the pressure drop across the valve is constant and the objective is flow control; then a linear valve characteristic is required to achieve a linear relationship between the controller output and the flow. Another exception occurs when a nonlinear relationship between the controller output and controlled variable is required in selected situations. For example, a cooling medium flow may normally be small or zero but need to be increased to a large value quickly upon demand. This situation would benefit from a nonlinear relationship between the controller output and the flow, which is provided by the "quick-opening" characteristic. Both of these characteristics are shown in Figure 16.6, and many other characteristics are commercially available (Hutchinson, 1976).

There are many physical designs of the flow patterns, orifice shape, and valve plug shape that are used to achieve the desired relationship. The specific design selected depends on many factors (Hutchinson, 1976), such as the desire to

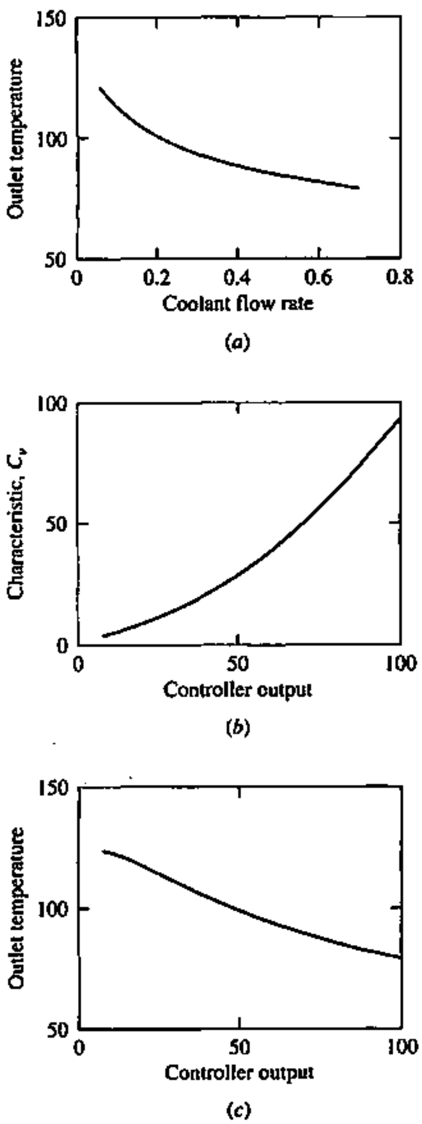


FIGURE 16.10

Summary of nonlinear process behavior and compensating characteristic.

TABLE 16.3

Method for achieving a linear control system by selecting the proper valve characteristic

Goal: A linear relationship [i.e., constant $G_{OL}(s)$] between the controller output and the controlled variable. The valve stem position is assumed to be equal to the controller output.

1. Determine the relationship between the pressure drop and the flow for the specific process system considered, K_{p1}.
2. Determine the relationship between the flow and controlled variable (if not flow rate), K_{p2}.
3. Calculate the C_v based on the results in (1) and (2) so that $C_v K_{p1} K_{p2} =$ constant. This will ensure that the steady-state gain of the process, as "seen" by the controller, is constant.
4. Select the commercial valve with the inherent characteristic, C_v, closest to the function determined in (3).

1. Have tight closing (i.e., no flow) when the controller output is 0% (or 100% for a fail-open valve)
2. Prevent sticking or clogging when the fluid is viscous or is a slurry
3. Accurately control the flow over a specified range
4. Reduce the pressure loss due to the valve, to conserve energy

The reader is cautioned that the selection of the proper control valve requires more information than is provided in this brief introduction. Details of typical valves, along with pictures of the internal details, are available and should be consulted (Hutchinson, 1976; Andrew and Williams, 1979, 1980; Driskell, 1983). Also, engineering standards for sizing calculations and selection are available for many common situations (ISA, 1992).

Finally, it should be noted that a nonlinearity can be added to the controller calculation in place of the nonlinear valve characteristic. Many commercial digital controllers have the facility to introduce a nonlinearity after the control algorithm, in the output processing phase, via a general polynomial. However, the use of the valve characteristic is still the most common means in practice for compensating for simple process gain nonlinearities.

16.6 ■ IMPROVING NONLINEAR PROCESS PERFORMANCE THROUGH CASCADE DESIGN

Other particularly simple nonlinearities can be addressed through cascade design that compensates for nonlinearities in the secondary, resulting in a (nearly) linear primary control system. One example encountered often in the process industries is maintaining two quantities in a desired proportion. An example of blending is shown in Figure 16.5*b*, although the concept applies to other proportions, such as reboiler to feed in a distillation tower or reactant ratio in a chemical reactor. The feedback controller can adjust the set point of the ratio controller as shown in Figure 16.5*b*. This is really another example of feedforward and feedback being combined as a product rather than a sum; thus, Figure 16.5*a* and *b* are alternative solutions to the same control design problem.

A cascade can also provide compensation for nonlinearities in other control designs. An example is shown in Figure 16.11, in which the relationship

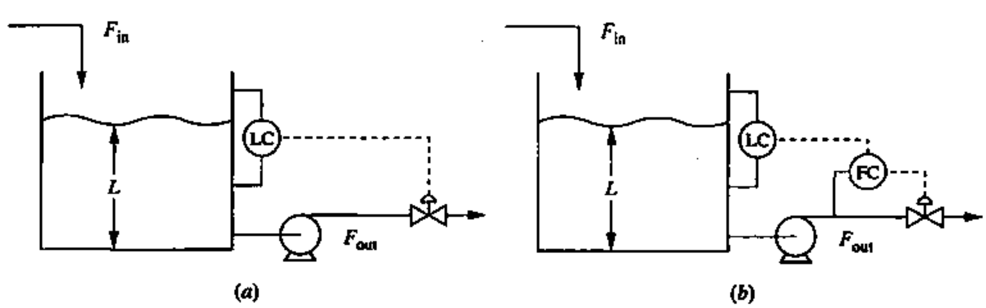


FIGURE 16.11

Example of cascade control applied to linearize the loop.

between the level and the level controller output is desired to be linear. Following the arguments in this section, the valve in Figure 16.11*a* would normally have an equal-percentage characteristic so that the relationship between the controller output and flow would be approximately linear. However, since the flow controller in Figure 16.11*b* is a fast loop, the relationship between the primary controller output and the flow would be linear, regardless of how well the characteristic compensated for other nonlinearities. Thus, the level control system is linearized as a result of the cascade design. Notice that the cascade strategy retains the advantages of improved disturbance response explained in Chapter 14.

16.7 ■ REAL-TIME IMPLEMENTATION ISSUES

Adaptive methods involving real-time calculations are relatively straightforward to implement; however, a few special considerations should be included. Some of the methods for adapting tuning are based on one or more measurements, and should a measurement not represent the true process conditions because of a sensor failure, the resulting tuning constants could be far from the proper values, leading to poor or even unstable performance. Thus, the measurement(s) used in the updating calculations should be checked for validity before being used to calculate the tuning. An example is checking the consistency of a flow measurement with associated flow rates in the process to ensure that a realistic flow is being used to update tuning. In addition, the value of the measured variable used in the correlations, as in equations (16.8), should be limited to the range over which the correlation is valid. This practice serves two purposes:

1. Error due to an unrecognized sensor failure is limited.
2. Extrapolation of a correlation beyond its region of applicability is prevented.

An issue that may not have been apparent in the previous sections is the ever-present need for defining the desired control performance. The tuning correlations must reflect the performance desired; thus, the tuning correlations selected must be based on control objectives consistent with the performance desired in the plant. As will be explained in Parts V and VI, tight control of one variable may degrade the control performance of another, more important variable because of process interaction. Thus, the performance goals of all control loops must be determined considering the overall process performance, which may lead to loose tuning for selected loops.

Also, it would be wise to provide the facility to fine-tune the controller tuning constants while retaining the correlations. One simple method would be to provide an adjustable parameter in equations (16.8). The engineer could adjust the parameter to achieve improved performance at one operating condition, and the parameter would be unchanged for other operating conditions.

16.8 ■ ADDITIONAL TOPICS IN CONTROL LOOP ADAPTATION

All of the methods described in detail in this chapter are based on the assumption that the change in process dynamics can be predicted. This assumption, which leads to the compensating calculations and equipment designs, is not always valid.

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For example, the effect of acid flow on pH (i.e., the shape of a pH curve) can change substantially due to changes in the buffering agents present; the effect of temperature on reactor conversion can depend on the activity of the catalyst. Therefore, there are situations in which deterministic methods are not appropriate. One response to this situation would be to detune the controller substantially and accept the performance degradation. Better performance would be possible with an adaptive method that could "learn" the process dynamics from the real-time system behavior and retune the controller based on the updated knowledge of process dynamics. Two general retuning approaches are used in this situation:

1. Periodic adaptive tuning at the request of a person, which is applicable when the dynamics change infrequently
2. Continuous adaptive retuning, which is applicable when the dynamics change frequently

The analysis of these approaches requires more advanced mathematics than is consistent with the level of this book; however, a few of the methods are introduced in the following paragraphs.

Periodic Retuning Based on Model Identification

In this approach an empirical model identification method is implemented to determine the dynamic model of the process, $G_p(s)G_v(s)G_s(s)$. The model fitting could use one of the methods described in Chapter 6 or other statistically based methods. Based on this model, the method can automatically introduce updated tuning using an appropriate controller tuning method. Note that this method introduces perturbations in the manipulated variable, which will disturb the process, but only when a person requests a retuning.

Periodic Retuning Based on Empirical Identification of the Critical Conditions

The Bode stability criterion highlighted the importance of the feedback system at the critical frequency. The feedback system's stability and controller tuning can be based on the amplitude ratio at the critical frequency, $|G_{OL}(\omega_c)|$. Thus, some methods of adaptive tuning determine the critical conditions empirically. One possible approach would be to automate the Ziegler-Nichols continuous cycling experiment described in Section 10.8, Interpretation IV; however, this approach would introduce large, prolonged disturbances. A more successful approach uses this principle with a relay in place of the controller to determine the same information, with smaller disturbances to the plant (Astrom and Hagglund, 1984).

Continuous Retuning Based on Statistics

It is possible to identify the process dynamics and determine how to modify the tuning without introducing external perturbations, as long as some disturbances occur in the process. Approaches to formulating and solving this problem are given in Astrom and Wittenmark (1989).

Continuous Retuning Based on Rules

Fine tuning of closed-loop systems based on the response to set point changes was discussed in Chapter 9. This concept can be applied to disturbance responses so that external perturbations are not necessary; then the method can be automated to achieve continuous retuning. In one commercial system the control performance is defined by the engineer in terms of (1) controlled-variable damping and overshoot, (2) expected noise levels, and (3) bounds on the controller tuning constants (Bristol, 1977; Kraus and Myron, 1984). The retuning method uses rules to adjust the tuning constants to achieve the desired performance.

16.9 ■ CONCLUSIONS

Modification of an element in the closed-loop system may be needed to attain high-performance feedback control when the feedback dynamics change. The three major steps are given in Table 16.2 for evaluating deterministic approaches for compensating for nonlinearities. The first step is to determine whether the process dynamics change significantly over the range of operation. If a fundamental analytical model is available, the linearized expression can be evaluated through the range of operating variables to determine whether the gain, dead time, and time constants change significantly. If no analytical model is available, several linear models can be determined through empirical identification at various operating conditions. The variability in the tuning and degradation in performance due to the nonlinearities can be determined as explained in Section 16.2. Since control objectives are different from plant to plant, it is not possible to give a generally applicable threshhold for when the nonlinearity is "significant." However, since modelling errors of ±20% are expected in identification, nonlinearities causing model parameter variations of this magnitude or less would normally not be considered significant.

The approaches presented in this chapter are summarized in Table 16.4. The order of presentation is from simplest and most reliable to most complex and challenging to implement. Generally, the engineer will apply the methods in the order presented in the table, proceeding only to the method needed to achieve acceptable performance.

If the variability is significant and it can be predicted based on real-time measurements, an element can be introduced to linearize the control loop by compensating for the nonlinearity. The compensating element can be in any of the three categories of the control calculation: input processing, control algorithm, or output processing. It can also be included in the control equipment, specifically in the final control element.

If the variability in dynamics is significant but cannot be predicted using correlations, one approach is to detune the feedback controller so that it is stable for all dynamics encountered. Naturally, this approach will result in a degradation in performance. Another approach is to modify the tuning of the controller based on some information of the real-time dynamic behavior of the system. Various methods are available, and references are provided.

Finally, the limits of the adapting approach should be recognized. First, a great strength of feedback control is that it does not require a highly accurate model. Thus, reasonable model errors can be tolerated with little degradation in control performance. Second, the adaptations require some time for the method to recogthe change in process behavior and introduce the compensation to the tuning (or other element of the loop). Thus, if the process dynamics are changing with a frequency near the critical frequency of the feedback control loop, an adaptive approach will not be able to introduce the compensation quickly enough. This

TABLE 16.4

Summary of methods to compensate control systems for nonlinearities

Description	Compensation for nonlinearity	Example	Additional effects on control performance
Measurement	Calculation to compensate for nonlinear sensor	Square root of orifice flow meter (Figure 12.5)	Improved accuracy for process monitoring
Final element	Final element selected to compensate for process nonlinearity	Valve characteristic to account for changes in pressure (Figures 16.6 and 16.7)	
Cascade control	Select secondary set point that has linear relationship with primary	Level-flow cascade (Figure 16.11)	Improved response to secondary disturbances
Detune	Determine single set of tuning constants for the range of operating conditions	Three-tank mixing process (Table 16.1, Case A tuning)	Poor performance can result
Gain schedule	Calculate the controller gain based on real-time measurement		
	Multiply feedforward and feedback (where this leads to proper gain scheduling)	Figure 16.5	Feedforward compensation for measured disturbance
Controller tuning	Calculate tuning based on a process model and real-time measurement	Three-tank mixing process [equation (16.8)]	
Occasionally retune	Empirically determine key model characteristics and tune controller according to preselected performance criteria	Relay method of finding critical conditions (Astrom and Hagglund, 1984)	Undesired variation during (infrequent) retuning
Continuously retune	Empirically determine key model characteristics and tune controller according to preselected performance criteria	On-line identification	

limitation also holds when an infrequent change in process dynamics is large abrupt: adaptation may not be able to detect the situation rapidly enough. Finally, the reader is advised to establish the potential improvement using the first entries in Table 16.4 before attempting the substantially more complex approaches in later table entries.

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ADDITIONAL RESOURCES

Computer programs and exercises for several adaptive tuning methods have been prepared by

Roffel, B., P. Vermeer, and P. Chin, *Simulation and Implementation of Self-Tuning Controllers,* Prentice-Hall, Englewood Cliffs, NJ, 1989.

Adaptive tuning of PID and other feedback control algorithms are presented at an advanced level in

Astrom, K., and T. Hagglund, *Automatic Tuning of PID Controllers,* ISA, Research Triangle Park, NC, 1988.

Goodwin, G., and K. Sin, *Adaptive Filtering, Prediction, and Control*, Prentice-Hall, Englewood Cliffs, NJ, 1984.

A particularly challenging situation occurs when the process gain changes *sign* as operating conditions change. An industrial process where this occurs is discussed in

Dumont, G., and Astrom, K., "Wood Chip Refiner Control," *IEEE Cont. Sys.*, *8*, 38–43 (1988).

Some industrial examples of adaptive control are given in the following references.

Piovoso, M., and J. Williams, *Self-Tuning Control of pH*, ISA Paper no. 0-87664-826-x, 1984.

Vermeer, P., B. Roffel, and P. Chin, "An Industrial Application of an Adaptive Algorithm for Overhead Composition Control," *Proc. Auto. Cont. Conf.*, St. Paul, MN, 1987.

Whately, M., and D. Pott, "Adaptive Gain Improves Reactor Control," *Hydro. Proc.*, 75–78 (May 1988).

QUESTIONS

16.1. Consider the three-tank mixing example process, but with the outlet concentration of component A changed to 50 percent in all cases. Recalculate the values in Table 16.1 for the same changes in the flow rate of stream *B*. Compare and comment on the similarities and differences.

16.2. Answer each of the following questions, with a full explanation of your answer.

(*a*) Could closed-loop frequency response, as explained in Section 13.3, be used to determine when feedback controller tuning should be adapted for changes in operating conditions?

(*b*) Review all cascade examples in Section 14.7 and determine whether each results in a (nearly) linear relationship between the secondary and primary. Would the single-loop control (primary to valve) be significantly nonlinear?

(*c*) Review all of the feedforward-feedback control designs in Section 15.7 and for each, recommend how to combine the feedforward and feedback signals (add, multiple, divide, other) to provide the best tuning compensation for the measured disturbances.

(*d*) The discussion and examples in this chapter involved feedback control. Discuss whether there is any advantage to adapting the adjustable parameters in a feedforward controller. If yes, discuss how this could be evaluated and the proper values determined.

16.3. Recalculate the tuning in Table 16.1 using the Ziegler-Nichols closed-loop tuning method. Compare the similarities and differences of the effect of F_B on the tuning for the two tuning methods. Which tuning would you recommend using?

16.4. Based on the information in question 9.10, would you recommend automatic deterministic retuning of the feedback control system? If yes, determine the measured variable and the tuning constants as a function of the measured variable.

16.5. Based on the information in question 10.2, would you recommend automatic deterministic retuning of the feedback control system? If yes, determine the measured variable and the tuning constants as a function of the measured variable.

16.6. You have been given the task of developing a rule-based adaptive tuning method for use with a PID controller. Also, the introduction of any perturbations by the method has been prohibited. Develop a set of rules that can be applied to normal operating data (with disturbances) to improve the performance gradually by adjusting the controller tuning constants. Remember to consider both the controlled and manipulated variables when evaluating performance.

16.7. In some control designs, the location of the sensor can be changed; one method for changing the effective sensor location is switching between sample tap locations that feed an analyzer. In this question we consider the dynamic system described in Example 5.2, Case 1. The original feedback PI controller measured Y4 and adjusted the input to the system; the modified feedback PI controller measures Y3 and adjusts the input to the system. Calculate the tuning for both PI controllers and decide whether the controller tuning should be adjusted when the sensor is switched.

16.8. The stirred-tank heat exchanger in Section 16.5 experiences changes in the feed inlet temperature of 120 to 170°C and in the coolant inlet temperature of 20 to 30°C. These temperature changes occur independently, and the feed flow and temperature set point remain constant at their base-case values. Discuss the need for adapting the feedback PI controller tuning constants and, if necessary, provide correlations for the valve characteristic and tuning as appropriate.

16.9. Design feedforward-feedback control for the chemical reactor in Example 15.1 for input disturbances in both feed composition, A2, and feed flow, F1. Pay particular attention to how the feedforward and feedback signals should be combined. Is there a need for adapting the feedback tuning for disturbances? Can this be achieved in combination with the feedforward control? Is there a need to adapt the feedforward controller parameters? Since you do not have fundamental models for this system, answer this question based on your qualitative understanding of the behavior of the process equipment.

16.10. The behavior of the heat exchanger in the recycle system in Example 5.3 varies due to fouling. Experience has shown that G_{H2} changes within the range of 0.20 to 0.32 about its nominal value of 0.30. Determine whether this change is significant. If so, how could deterministic controller adaptation be implemented?

16.11. Sometimes process equipment has to be removed from service occasionally for maintenance. Consider a multiple-tank mixing process that is basically the mixing tank process in Example 7.2, but modified to have between two and four tanks, depending on the equipment availability. Determine how the feedback controller tuning has to be modified for the situations of two, three, and four tanks in the feedback process. Also, compare the control performance for these three situations.

16.12. Level control is to be added to the draining tank process in Example 3.6. The controller adjusts the opening of a valve in the exit pipe at the base of the tank, and essentially all of the pressure drop in the pipe and valve occurs across the valve. Determine the valve characteristic that will yield a linear relationship between the controller output and the level. The inlet flow varies from 50 to 150 m^3/h.

16.13. In some feedback control systems the manipulated variable can be changed, usually by selecting the position of a switch at the controller output that directs the signal to one of the possible manipulated variables. For the following cases, determine whether the difference in feedback dynamics is significant enough to require changing the tuning depending on the manipulated variable selected for the controller output.

(*a*) For a distillation column, the controlled variable is the light key in the distillate, X_D, and the two manipulated variables are the reflux flow, FM_R, and the vapor boilup, VM_0. For dynamic models, refer to Figure 5.17.

(*b*) For a single, isothermal CSTR, the controlled variable is the effluent reactant concentration, C_A, and the two manipulated variables are the inlet concentration, C_{A0}, and the total feed flow rate, F. For dynamic models, refer to Example 5.5.

(*c*) For an open-top liquid tank, the controlled variable is the liquid level, and the two manipulated variables are the valves in the two outlet pipes. The process is sketched in Figure Q1.9*a*.

16.14. Question 13.1 describes a process with feedback control and changes in operating conditions, (*a*) through (*f*). For each change in operating conditions, determine whether it is necessary to adapt the feedback controller tuning, and if so, how the adaptation could be implemented automatically.

16.15. Consider a series of *three* isothermal CSTRs, each with the physical design parameters of the process in Example 3.5. The base case operating conditions are the same as the example: $F = 0.085$, $C_{A0} = 0.925$, and $k = 0.50$. The composition of reactant A in the third reactor is controlled by adjusting the feed composition, C_{A0}. Determine (*a*) the steady-state operating conditions for this base case, (*b*) the linearized model for the system, and (*c*) PID feedback tuning for this base case system. Then determine whether the controller tuning must be adjusted if the feed flow rate changes from 0.085 to 0.20.