Variable-Structure and Constraint Control

CHAPTER 22

22.1 ■ INTRODUCTION

To this point we have made the assumption that the multivariable process has the same number of manipulated and controlled variables. This situation is often referred to as a square or $n \times n$ system. Square systems are typical, because we consider dynamic behavior and control when designing plants and provide sufficient manipulated variables for at least the most important controlled variables. However, it is often the case that, due to process limitations and overriding control objectives, the number of manipulated and controlled variables are not always equal, and control approaches are needed to address these situations.

In this chapter, situations will be considered in which the number of manipulated variables is greater than or less than the number of controlled variables. When an excess of manipulated variables exists, the controlled variables can be returned to their set points at steady state by many combinations of the steady-state manipulated variables. Thus, the control system should operate the process in the most economical manner, in addition to providing good dynamic performance. When an excess of controlled variables exists, not all controlled variables can be maintained at their set points simultaneously. However, the control system can be designed to maintain the most important controlled variables at their set points.

The branch of process control that addresses these situations is known as *variable-structure* control. In this chapter, methods based on *single-loop control algorithms* are presented that provide the ability to change the input-output pairings of selected loops automatically. These methods are easy to design and simple to use and are therefore widely applied in practice. However, they are normally restricted to cases with limited dimensionality, such as one manipulated and several controlled variables or several manipulated variables and one controlled variable. A method that can address higher-dimensional structures, as well as square problems, is presented in the next chapter.

First, split range control systems are presented for processes with excess manipulated variables. Then, signal select control systems are presented for processes with excess controlled variables. In each section, examples demonstrate typical reasons for variable structure control, along with implementation guidelines. Finally, a few applications of constraint control are provided; these demonstrate the combined application of split range and signal select, along with some frequently used extensions, such as multiple controllers with different set points and valve position controllers.

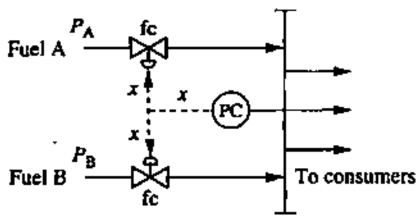

FIGURE 22.1

Split range pressure control.

22.2 ■ SPLIT RANGE CONTROL FOR PROCESSES WITH EXCESS MANIPULATED VARIABLES

The concept of split range control will be introduced through the example process in Figure 22.1. In this process, the flows of gaseous fuels from two sources are adjusted to control the pressure of a header, which is a pipe from which fuel is distributed to many consumers. The flow to the consumers is determined by many independent processes and cannot be adjusted to control the pressure. The following simple model of the (well-mixed) gas header system can be used to evaluate the degrees of freedom:

$$V\frac{dC_A}{dt} = F_A C_{A0} - F_{out} C_A \tag{22.1}$$

$$V\frac{dC_B}{dt} = F_B C_{B0} - F_{out} C_B \tag{22.2}$$

$$F_A = K_A v_A \sqrt{\frac{P_A - P}{\rho_A}} \tag{22.3}$$

$$F_B = K_A v_B \sqrt{\frac{P_B - P}{\rho_B}} \tag{22.4}$$

$$P = \frac{(VC_A + VC_B)RT}{V} \tag{22.5}$$

Variables	External variables	Constants
P	F_{out}	R
F_A	T	V
F_B	P_A	K_A
C_A	P_B	K_B
C_B	C_{A0}	ρ_A
v_A	C_{B0}	ρ_B
v_B		

The model could be improved by including nonlinear valve characteristics and a nonideal gas law, but the model of this resolution is sufficient to demonstrate the degrees-of-freedom analysis. There are 5 equations and 7 variables; thus, the system is not specified. For this system to be specified, values for two input variables, v_A and v_B, should be defined. In this example, the prices of the two fuels are not equal; fuel A has a lower price than fuel B. Therefore, the control system should automatically adjust the valves so that as much of fuel A as possible is consumed before any fuel B is consumed, while providing good control of the pressure.

The split range control system in Figure 22.1 achieves the desired behavior in a simple manner. The pressure in the header is measured and used as the controlled variable to a standard PI feedback control algorithm, which has a single calculated output signal, x. This signal is sent to both control valves, but these valves are calibrated to open or close differently from the standard control valves. To achieve the desired behavior, the controller and valves obey the behavior defined in Table 22.1 and shown in Figure 22.2.

With this modification, the control equations become, for controller output $x < 50\%$,

$$v_A = 2\left\{K_c\left[(P_{sp} - P) + \frac{1}{T_I}\int_0^t (P_{sp} - P)dt'\right] + I\right\} \qquad (22.6)$$

$$v_B = 0.0$$

TABLE 22.1

Typical valve adjustments for split range control

		Percent opening	
Controller output	**Pressure to valve**	**Valve A**	**Valve B**
0–50%	3–9 psig	0–100%	0%
50–100%	9–15 psig	100%	0–100%

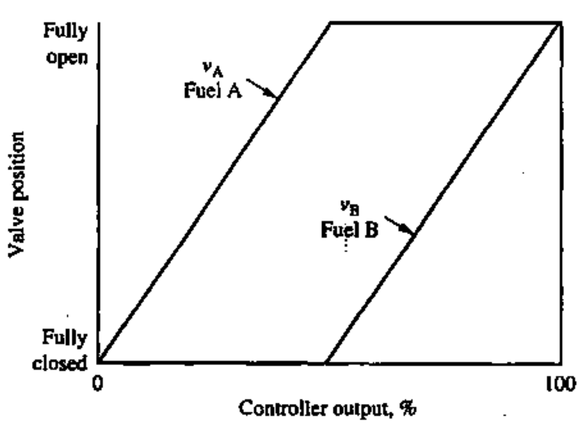

FIGURE 22.2

Fixed ranking of valve adjustments for split range.

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For controller output $x \geq 50\%$,

$$v_A = 100$$
$$v_B = -100 + 2\left[K_c\left((P_{sp} - P) + \frac{1}{T_I}\int_0^t (P_{sp} - P)dt'\right) + I\right] \tag{22.7}$$

Note that either set of two equations introduces no dependent variables and one external variable, P_{sp}, along with the controller tuning constants. The combination of the controller equations, either (22.6) or (22.7), with equations (22.1) through (22.5) results in a system with 7 equations and 7 variables. Thus, the process and control system is completely defined when the pressure set point has been specified.

Split range control is depicted in the process diagram in Figure 22.1. The fixed relationship between the controller output and the position of the two valves is shown in Figure 22.2. As the controller output initially begins to increase from 0%, the valve in the less expensive fuel line opens, while the valve in the more expensive fuel line remains closed. When the controller output reaches 50%, the fuel A valve is fully open, and the fuel B valve is closed. When the controller output continues to increase beyond 50%, the fuel A valve remains fully open, and the fuel B valve opens.

The behavior of the control system is given in Figure 22.3. The initial situation has a low total fuel demand, so that the pressure controller manipulates only the fuel A valve. At time 30, an increase in the fuel consumption occurs; the pressure in the header initially decreases; and the controller output increases. The fuel A valve is adjusted until the pressure is returned to its set point. At time 110 another increase in consumption occurs. The pressure controller responds by increasing its output. In this situation, valve A reaches its limit of 100%; then the fuel B valve is opened until the pressure is returned to its set point. This example demonstrates that the split range controller can smoothly adjust the two valves to maintain the controlled variable at the set point, while minimizing the cost of the fuel consumed.

Several important implementation issues arise in applying split range control. In principle, the concept of split range can be extended to any number of manipulated variables. However, there is a limit on how accurately a control valve can be adjusted. Therefore, split range is normally limited to two, or three at most, manipulated variables. Also, a feedback control system could tend to cycle if it had a "dead zone" in which neither valve is adjusted. To prevent this situation arising from inaccurate valve calibrations, the valves are normally calibrated to have an overlap (e.g., 0 to 55% and 45 to 100%).

Typical behavior can be easily implemented by a simple calibration of standard control valves, which has essentially no cost implication. Recall that the signal to the single-loop control valve is normally 3 to 15 psig, which relates to 0 to 100% of the controller output signal, respectively.

Another important issue is the stability and tuning of the control system. Note that the feedback process dynamics change when the controller output crosses the 50% value, as shown in Figure 22.4. Therefore, the controller tuning should remain constant only if the process dynamics for the two closed-loop paths are the same or similar; that is, if $G_{vA}(s)G_{pA}(s) \approx G_{vB}(s)G_{pB}(s)$. If the closed-loop dynamics are significantly different, the controller tuning should be changed automatically by the control system. The controller tuning could be switched based on the value of the controller output, using one set of tuning constants for controller output values

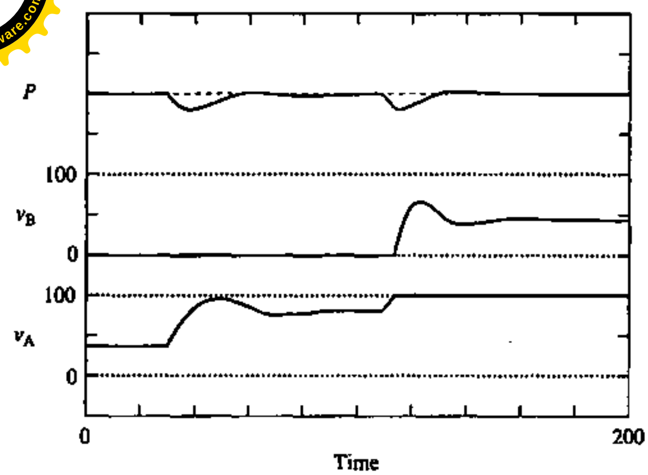
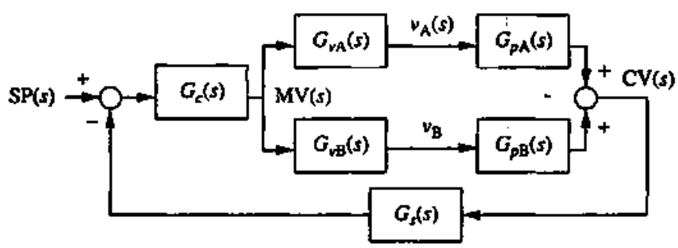


FIGURE 22.3

Dynamic response of split range control system.



Value of MV	Characteristic equation
0–50%	$1 + G_{pA}(s)G_{vA}(s)G_c(s)G_s(s) = 0$
50–100%	$1 + G_{pB}(s)G_{vB}(s)G_c(s)G_s(s) = 0$

FIGURE 22.4

Schematic of split range control.

TABLE 22.2

Split range control criteria

Split range control is possible when

1. There is one controlled and more than one manipulated variable.
2. There is a causal relationship between each manipulated variable and the controlled variable.
3. The proper order of adjusting the manipulated variables adheres to a fixed priority ranking.

of 0 to 50% and another set of tuning for 50 to 100%. This retuning approach is another application of the adaptive tuning method referred to in Section 16.3 as *deterministic modification* of controller tuning.

In conclusion, split range control is widely applied to processes with excess manipulated variables. The general criteria for split range control are summarized in Table 22.2. The feedback controller can use one tuning if the dynamics for all feedback paths are similar. If the dynamics are significantly different, the feedb… controller must be (1) detuned to be stable without excessive oscillations for a… situations or (2) retuned automatically via programmed modification.

22.3 ■ SIGNAL SELECT CONTROL FOR PROCESSES WITH EXCESS CONTROLLED VARIABLES

Often, many control objectives exist for a process, and not all of these can be satisfied simultaneously. As an example, consider the chemical reactor shown in Figure 22.5, which has control objectives to maximize conversion while (1) maintaining the reactant composition in the effluent at or above a value $(C_A)_{min}$ and (2) preventing the reactor temperature from exceeding T_{max}. Each of these control objectives can be satisfied individually by adjusting the cooling medium flow rate. The engineer must determine the relative importance of the control objectives and design a control system that satisfies the priority ranking.

A signal select control strategy to implement this ranking is shown in Figure 22.5. An individual controller is implemented for each measured controlled variable, and the output signals from the two controllers are sent to a signal select element.

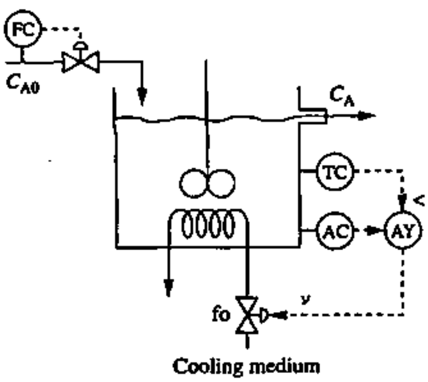


FIGURE 22.5

Example of signal select with two controllers.

> The output of a signal select is either the minimum (low signal select) or maximum (high signal select) value of all inputs to the signal select.

In the example, the proper element is a low signal select, since the largest flow of cooling medium is preferred and the valve is fail-open. (Selecting the lowest signal to the cooling medium valve ensures the largest coolant flow.) The output of the signal select is sent to a control valve, as in this case, or can be sent to the set point of a secondary controller in a cascade system.

Again, the degrees of freedom of the control system should be analyzed. The equations that define the process and the control calculations for the example are as follows:

$$V\frac{dC_A}{dt} = F(C_{A0} - C_A) - Vk_0e^{-E/RT}C_A \tag{22.8}$$

$$V\rho C_p\frac{dT}{dt} = F\rho C_p(T_0 - T) - UA(T - T_c) + (-\Delta H_{rxn})Vk_0e^{-E/RT}C_A \tag{22.9}$$

$$\text{MV}_1 = K_{c1}\left[(C_{Asp} - C_A) + \frac{1}{T_{I1}}\int_0^t (C_{Asp} - C_A)\,dt'\right] + I \tag{22.10}$$

$$\text{MV}_2 = K_{c2}\left[(T_{sp} - T) + \frac{1}{T_{I2}}\int_0^t (T_{sp} - T)\,dt'\right] + I \tag{22.11}$$

$$UA = f(v) \tag{22.12}$$

$$v = \min(\text{MV}_1, \text{MV}_2) \tag{22.13}$$

Variables	External variables	Constants
C_A	C_{A0}	ρ
C_{Asp}	T_0	C_p
T	T_c	V
T_{sp}		
UA		R
MV_1		E
MV_2		k_0
v		$(-\Delta H_{rxn})$
		$K_{c1}, K_{c2}, T_{I1}, T_{I2}$

The system has 6 equations and 8 variables; thus, the system behavior is defined when two variables, C_{Asp} and T_{sp}, have been specified (i.e., are shifted to external variables). To achieve the objectives in this example, the set points must be set to the limiting values for these variables [i.e., $C_{Asp} = (C_A)_{min}$ and $T_{sp} = T_{max}$].

Depending on the operating conditions of the chemical reactor (e.g., feed composition, temperature, concentration of reaction inhibitors), either of the controllers could be selected to manipulate the cooling medium valve. Dynamic responses for two feedback systems are given to demonstrate the effect of the signal select. The initial steady-state conditions of the system result in the outlet composition being at its set point (minimum value) and the temperature being below its set point (maximum value). After a short initial period of steady operation, reaction inhibitor is introduced with the feed, causing the reaction rate to decrease. The first response involves the reactor system with only composition control (and no temperature control), so no signal select exists. As shown in Figure 22.6*a*, the

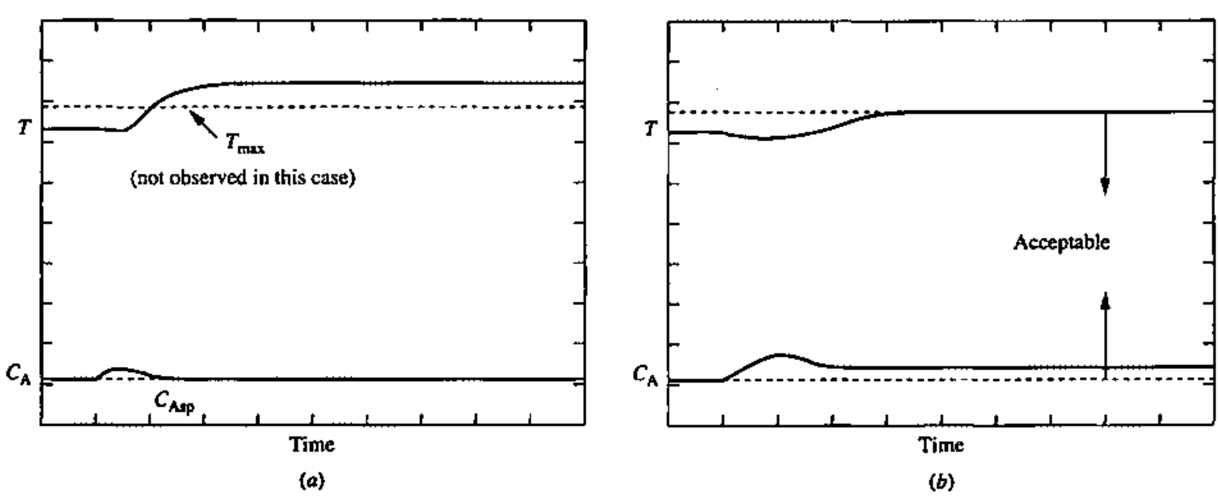


FIGURE 22.6

Reactor disturbance response: (*a*) with only composition control; (*b*) with signal select design.

composition controller reduces the coolant flow to increase the reaction rate. The control system returns the composition to its set point, but it increases the reactor temperature above its maximum value.

The second response involves the same reactor and disturbance but with the signal select control design shown in Figure 22.5. The dynamic response is given in Figure 22.6*b*. Initially, the temperature is below its maximum limit, and the composition of product in the effluent is at its set point. Since the temperature controller is sending a higher signal (to increase the temperature) than the composition controller, the output of the composition controller is initially selected. In the initial response to the disturbance, the coolant is decreased by the concentration controller until the reactor temperature reaches its maximum value: the temperature controller set point. Then, the temperature controller output signal becomes smaller than the output from the composition controller. At the new steady-state conditions, the temperature is at its set point and the composition is above its set point. This situation may not be the most profitable in the short run, but it is the best operation, given the input variables, because it prevents damage to equipment due to extreme temperature. Improvement would require an elimination of the inhibitor in the feed.

The steady-state relationships between the split range manipulated and controlled variables are depicted in Figure 22.7 for the reactor example. Two cases show how the control objectives can be satisfied by adjusting the manipulated variable when the temperature or composition is the limiting factor. When sufficient range exists for the manipulated variable, the best steady-state operation can be achieved by opening the valve the least amount as constrained by the most limiting controlled-variable value. If the manipulated variable does not have sufficient range, the proper value of the output of the signal select is at either its minimum value (more cooling capacity required) or maximum value (zero cooling insufficient, heating required). When the valve saturates, no control system can do better; the equipment design or cooling medium temperature must be changed to satisfy the objectives. Thus, the simple signal select control system always achieves the best (unique) steady-state performance possible for the process design and control objectives.

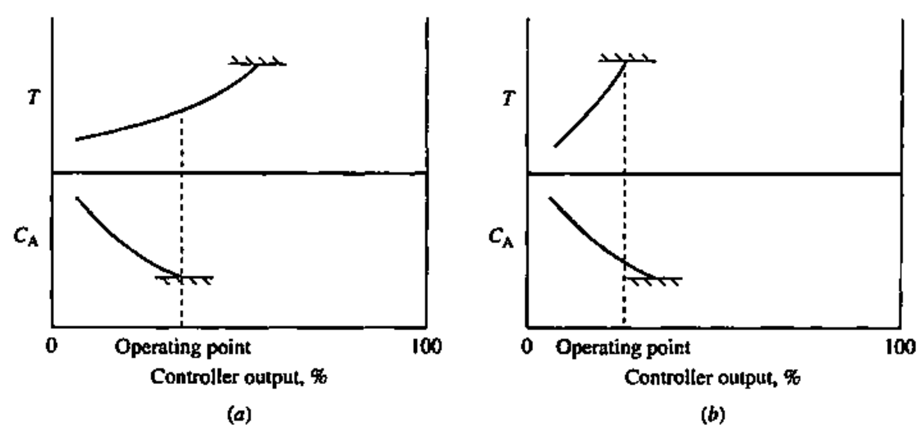


FIGURE 22.7

Determining the operating point for systems with signal select control.

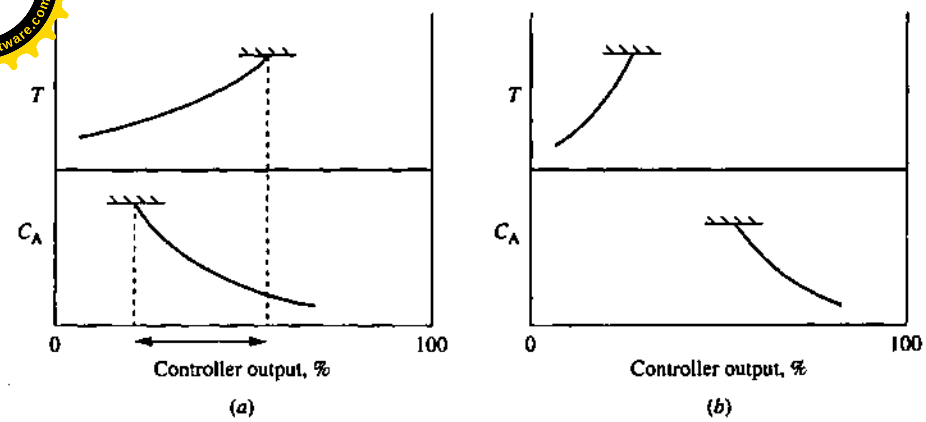


FIGURE 22.8

Systems for which signal select control is not appropriate.

However, signal selects are not appropriate for all cases of multiple controlled variables and one manipulated variable. An example where signal select is not appropriate is the same chemical reactor as in Figure 22.5 with different control objectives: Maintain (1) the effluent composition to be *no greater than* $(C_A)_{max}$ and (2) the temperature below T_{max}. This situation is depicted in Figures 22.8*a* and *b*. In Figure 22.8*a* both limits can be satisfied, but the control objectives are not defined completely enough to determine a unique value of the manipulated variable. In Figure 22.8*b*, no value of the manipulated variable satisfies the objectives. In either case, no unique operating point exists. Therefore, the control system must perform a task more complex than determining a limiting value. It must determine the best or "optimum" operation within acceptable limits (Figure 22.8*a*) or the operation that violates the important limits the least (Figure 22.8*b*). This task cannot be performed by signal selects but can be solved, using additional criteria entered by the engineer, with an optimization calculation. Control algorithms that are capable of performing optimization are introduced in Chapter 26.

The split range elements are designated by the symbols in Figure 22.9. As presented in Appendix A the designation "Y" is used for the second letter inside the symbol for a calculation and the less-than or greater-than symbol to indicate low or high select. An older method that is still used frequently is to write LSS and HSS for low and high signal select, respectively.

The term *signal select* indicates that many different types of signals, not just controller outputs, can be used. As another example, the temperatures along a packed-bed reactor are monitored, and each temperature is to be maintained below its specified value. Two signal select control systems are shown in Figure 22.10*a* and *b*. In Figure 22.10*a* the measurements are input to a high signal select, and the output of the select is used as the *controlled* variable for a single controller, which adjusts the preheat. In Figure 22.10*b* each measurement goes to a separate controller, each controller output goes to the low signal select, and the output of the signal select goes to the *control valve*.

Both designs could succeed in maintaining the highest measured temperature at the set point. One difference is that the design in Figure 22.10*a* has one controller with one set of tuning constants, whereas the design in Figure 22.10*b* has separate

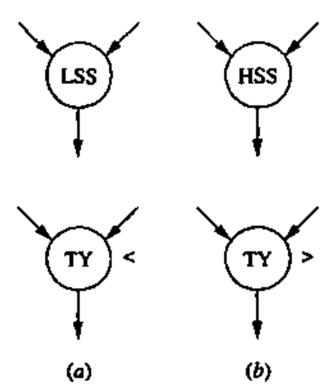
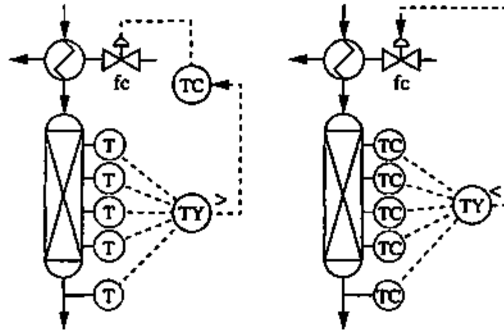


FIGURE 22.9

Symbols for signal selects.



(a) (b)

FIGURE 22.10

Examples of signal select control on:
(a) measurements; (b) controller outputs.

tuning for each controller. The design in Figure 22.10*b* would be preferred if the feedback loop dynamics change with the measurement selected, as they might in this example. If the loop dynamics are essentially the same for all measurements, the design in Figure 22.10*a* would be preferred for its simplicity. Also, the design in Figure 22:10*a* enforces the same set point value for all measured variables, whereas the alternative design in Figure 22.10*b* allows different set points for different locations in the packed bed.

The designs in Figure 22.10 are similar, and often engineers have difficulty selecting between them. The proper selection is based on the recognition that the process dynamics in the feedback loop should be nearly constant (when the controller tuning constants are unchanged). The design in Figure 22.10*b* can have tuning tailored to each measurement and is thus a more general design. Three cases can occur:

1. When every closed-loop system has the same dynamics,

$$\frac{T_1(s)}{v(s)} = \frac{T_2(s)}{v(s)} = \cdots = G_p(s) \tag{22.14}$$

   either design in Figure 22.10 can be used.

2. When the closed-loop systems have the same dynamics except for the steady-state gain,

$$\frac{T_1(s)}{v(s)} = K_1 G_p(s) \qquad \frac{T_2(s)}{v(s)} = K_2 G_p(s) \cdots \frac{T_i(s)}{v(s)} = K_i G_p(s) \tag{22.15}$$

   the design in Figure 22.10*a* can be used if the controller gain is divided by a value K_i to compensate for the feedback process gain of the selected temperature, which would tune the single temperature controller to give the same stability margin:

$$G_{OL}(s) = K_i G_p(s) \frac{K_c}{K_i}\left(1 + \frac{1}{T_I s}\right) \tag{22.16}$$

TABLE 22.3

Signal select criteria

A signal select is possible when

1. There is one manipulated variable and several potential controlled variables.
2. There is a causal relationship between the manipulated variable and each controlled variable.
3. There is a unique, feasible operating point that satisfies all control objectives in the steady state (see Figure 22.7).

3. When the process dynamics are significantly different in each feedback loop, only the design in Figure 22.10*b* can be used for controllers with constant tuning values. (See Chapter 16 for evaluation of significant differences and methods for modifying controller tuning in real time.)

A very important implementation issue in the application of signal selects is the potential for reset (integral) windup in systems like the ones shown in Figures 22.5 and 22.10*b*. While all controller outputs are sent to the signal select, only one is used to determine the valve position; thus, there is only one feedback control system. The outputs from the other controllers do not influence the manipulated variable, and because of their controller integral modes, the outputs from the controllers not selected could wind up (i.e., increase or decrease without limit). Several possible solutions exist to prevent windup, with perhaps the clearest being the application of external feedback, which was introduced in Chapter 12. For signal select control, the value *after* the signal select is used as the external feedback variable for all controllers whose outputs go to the signal select. Such a system will not experience integral windup.

In conclusion, signal selects are widely applied to processes with excess controlled variables. The general approach for using a signal select is summarized in Table 22.3. Controlled variables can be used as inputs to a signal select if the feedback loop dynamics are similar for all controlled variables. The controller outputs should be used in the signal select if the feedback loop dynamics are different.

22.4 ■ APPLICATIONS OF VARIABLE-STRUCTURE METHODS FOR CONSTRAINT CONTROL

The variable-structure control methods introduced in this chapter are based on a fixed ranking of controlled or manipulated variables. As a result, the operating conditions achieved through the control system maintain the process near a limiting or constraining value—for example, the minimum use of the more expensive fuel or the maximum reactor temperature. Control systems that result in process operation near a limit are generally termed *constraint controllers*, and since they often require variable-structure capability, constraint control is often implemented using split range and signal select methods. A few additional examples of constraint control are presented in this section.

Combined Variable-Structure Methods

This first example demonstrates how split range and signal select can be combined in control designs to achieve good control performance. Consider the situation in Figure 22.11, which shows two series processes. The product flow from unit 1 is usually not equal to the feed flow to unit 2; therefore, a large storage tank is located between the two units. One approach for dealing with the differences in flows would be to cool the entire production from unit 1, send it to the storage tank, and heat the feed to unit 2 as it flows at its desired rate from the storage tank. This approach would provide smooth and reliable flow control, but it would be very energy-inefficient.

A more efficient alternative approach would be to provide the maximum allowable direct flow from unit 1 to unit 2. The maximum direct flow between units would be determined by either the availability from unit 1 or the demand for unit 2, with the limiting condition changing as both unit operations change. A control system to maximize the direct flow automatically while always achieving proper level and flow control would be desirable. Such a system is shown in Figure 22.11, where both the level and flow controllers have split range outputs. Both controller outputs are sent to the low signal select, which determines the proper signal to manipulate the direct flow valve, which in this example is the smallest signal, which gives the smallest direct flow rate. Thus, one controller will adjust the direct flow valve, and the other controller will continue to increase its output until it adjusts the flow to the tank (for level control) or flow from the tank (for flow control), as appropriate. The resulting operations for the two situations are summarized in the following table.

	How each valve is adjusted			
Relative flows	**v100**	**v110**	**v200**	**Net flow**
Unit 1 flow > unit 2 flow	By FC	By LC	Closed	To storage
Unit 1 flow < unit 2 flow	By LC	Closed	By FC	From storage

With this control system the plant personnel need only input the set points to the level controller (normally 50 percent of range) and the flow controller (required flow to unit 2). The system automatically adjusts the valves as described to meet the level and flow requirements while minimizing flows to and from storage, thus minimizing energy use.

Multiple Controllers for One Variable

A design involving two separate controllers is sometimes used as an alternative to split range when the use of one manipulated variable is to be strictly minimized, even in short transients. For example, consider the level-flow system in Figure 22.12*a*, in which the level is normally controlled by manipulating the flow to a downstream unit but can be controlled by adjusting the flow to waste, if required. Naturally, the flow to waste is to be minimized. A split range controller is employed in the figure to achieve the control objective.

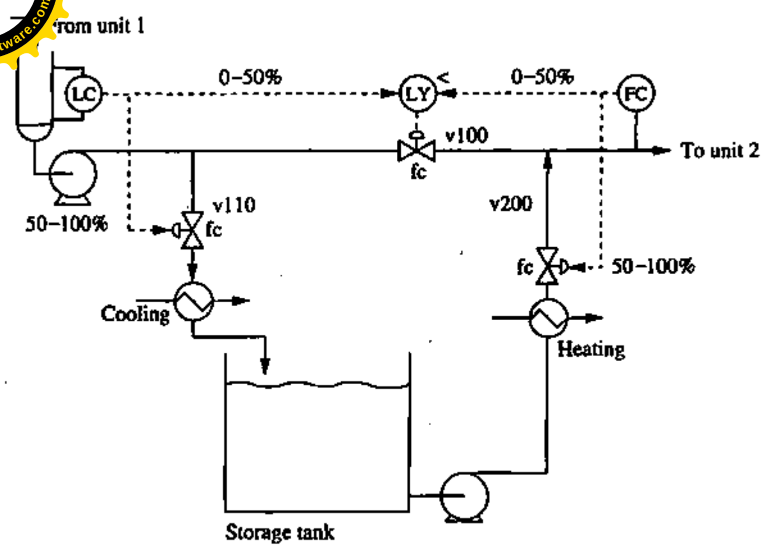
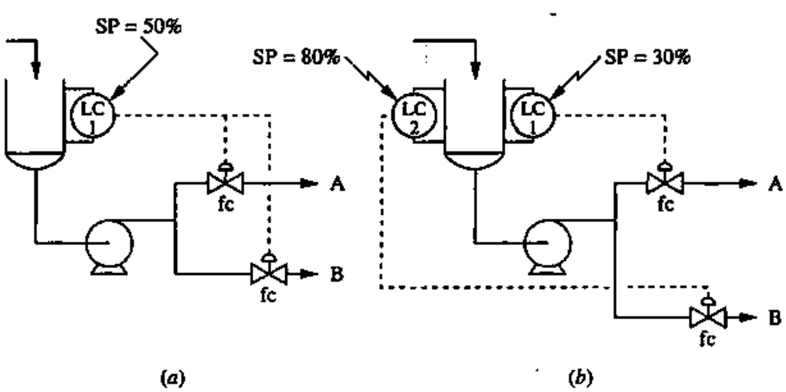


FIGURE 22.11

Example of combined split range and signal select.



A = downstream unit, B = waste

FIGURE 22.12

Alternative approaches to controlling inventory by (*a*) split range; (*b*) two controllers with different set points.

An alternative control system is given in Figure 22.12*b*, which employs two feedback controllers with *different set points*. Under normal conditions, the controller with the set point of 30 percent level (LC-1) adjusts the flow to the downstream unit, and the valve to waste is completely closed. If the flow in becomes large, the valve to the downstream unit is opened completely. If the flow in is still larger than the maximum flow to the downstream unit, the level then increases above 30 percent. If the large flow in remains for a long enough period of time, the level reaches the set point of the alternative controller with a higher set point (LC-2), here shown as 80 percent. When this level is reached, the alternative controller begins to increase the flow to waste.

The two-controller design in Figure 22.12*b* has three advantages. First, it uses the inventory in the vessel, so there is no flow to waste until the flow to downstream unit is at its maximum *and* the level increases to the upper set point. Thus, short-term disturbances in the inlet flow that can be accumulated in the vessel will not result in material being diverted to waste. Second, it has two sensors, valves, and controllers, so failures of elements in both control loops would have to occur before the level could overflow. This increase in reliability would be important if a large safety or economic penalty were incurred for an overflow. Finally, the use of two controllers allows separate tuning for the two feedback loops, although this probably would not be necessary in the example in Figure 22.12. The split range controller has one set point and one set of tuning constants and is preferred, if it achieves the objectives, because of its simplicity.

Valve Position Control

Sometimes a limit is the result of equipment performance, and the approach to the limit is not easily inferred from measured process variables such as flow or temperature. This situation is demonstrated in Figure 22.13, in which the feed rate to a chemical reactor is to be maximized. The reactor temperature must be maintained constant, and the heat exchanger duty is the limiting factor in increasing the feed rate. There is no process variable that indicates how close the process operation is to the limit. One indication that the limit had been *exceeded* would be the reactor temperature remaining below its set point for a long time; however, this indication would be available only after the process had been upset. Thus, this measure of the limit is not normally acceptable.

Another potential indication of the limit is the temperature controller output, which is essentially the value of the heating medium valve position. When this value nears its maximum value of 100 percent, the limitation in heating duty is being approached. This analysis leads to the use of a *valve position controller*

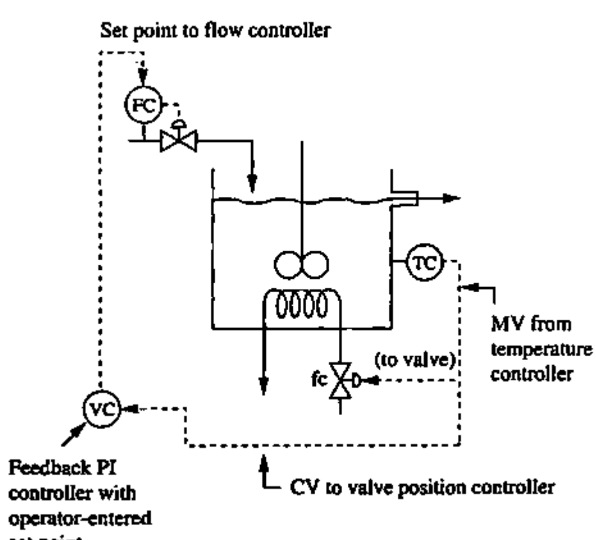


FIGURE 22.13

Example of valve position control.

C), which uses the temperature *controller output* as its controlled variable and adjusts the feed flow controller set point. This is a feedback system and can use a standard proportional-integral algorithm; the set point of the valve position control system is chosen sufficiently far from the limiting value that the valve nearly never reaches a limit during a transient response to an upset. This approach ensures that the temperature control system has the range to respond to high-frequency disturbances and maintain the temperature at its set point. (A typical value might be 90%, but could be lower if the system experiences large temperature disturbances.)

The valve position controller feedback path involves the reactor and temperature control loop. Therefore, the valve position controller must be tuned loosely so as not to upset the temperature controller and to provide smooth, nonoscillatory approach to the constraint. Also, the feedback loop in the valve position controller includes the temperature controller; in other words, there is no causal feedback path without the temperature controller functioning. Thus, the valve position controller represents a loop pairing on a zero relative gain (see Chapter 20), and a monitoring program is recommended to determine whether the temperature controller is functioning and, if not, to switch the valve position controller into manual status.

Plantwide Variable-Structure Control

Sales demands and prices sometimes result in the pleasant circumstance that all of the plant's production can be sold at a profit. In this situation, the control system should be structured to result in the highest production rate possible, consistent with product quality and equipment performance limitations. Since most plants have several possible limiting factors, a variable-structure control system is normally used to monitor all likely limiting factors and adjust the feed rate so that the most restrictive factor closely approaches but does not violate its limiting value.

The situation is shown in Figure 22.14 for a hypothetical plant in which three possible factors could limit the operation: heating medium availability for the

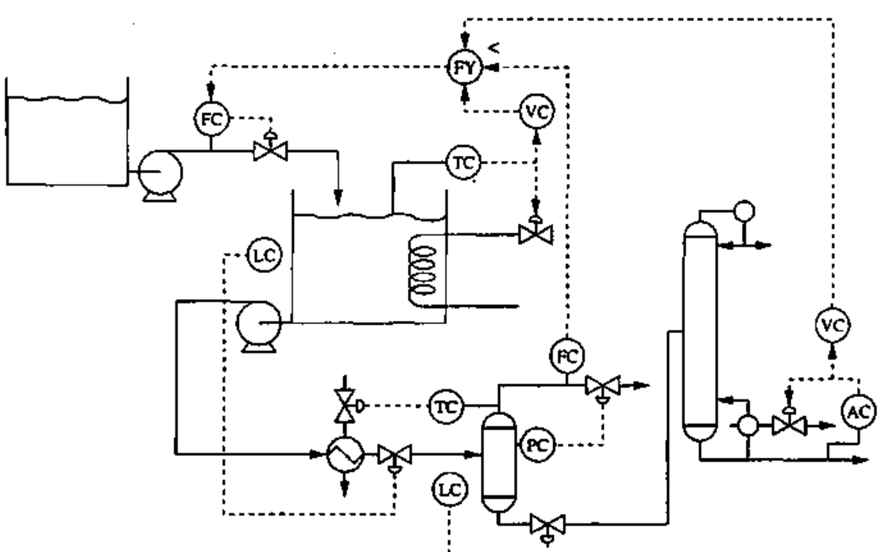


FIGURE 22.14

Example of maximum feed constraint control.

reactor, maximum flow of vapor product from the flash drum, and maximum boiler duty in the distillation tower. The control system monitors all three (i.e., with valve position controllers), uses each in a *feedback* controller, and selects the lowest value of the three controller outputs to adjust the feed flow set point. (Note that many important controllers are not shown in the simplified figure so that the feed maximization can be clearly shown.) Since the feedback processes are relatively slow for these constraint controllers, their set points should not be exactly the limiting values; the set points provide a safety margin from the limits, to account for the likely variability about the set point.

22.5 ■ CONCLUSIONS

To achieve process objectives, engineers design equipment with appropriate capacities, provide measurements and manipulated variables, and design flexible control systems to respond to normal and upset conditions. Variable-structure control often enables the system to satisfy the operating objectives when the numbers of manipulated variables and controlled variables are not equal. Two methods have been presented in this chapter that are applicable to commonly occurring objectives. In split range control a single feedback controller output is sent to more than one final element, and the final elements are calibrated to operate over different ranges of the controller output signal (e.g., 0 to 50% and 50 to 100%). A signal select, on the other hand, is used when there are several controlled variables and one manipulated variable, and the signal select determines the most limiting control objective.

These methods are appropriate for situations in which the best operation resides on a constraint or "frame" of the steady-state operating window, as demonstrated in the following examples. First, consider the fuel pressure split range control in Figure 22.1. Since the flexibility in the system involves the manipulated variables, the operating window in Figure 22.15 has manipulated variables as the coordinates. Any point inside the steady-state window that satisfies $F_A + F_B = \sum F_{consumer}$ is a feasible plant operating point. Clearly, there are infinite combinations of fuel flows that can satisfy the total consumer demand. The best operation is designated by the dashed line, which shows the combination of flows

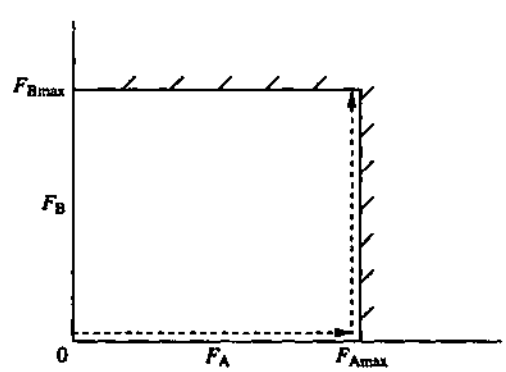


FIGURE 22.15

Operating window for fuel pressure split range control.

the two fuels that satisfies the consumer demand from zero to maximum while also minimizing fuel cost. The split range control system implements this strategy and therefore is appropriate for this example.

Then consider the chemical reactor signal select control system in Figure 22.5. Since the flexibility involves the controlled variables, the operating window in Figure 22.16 has controlled variables as the coordinates. Any point in the window represents feasible plant operation, and there is an infinite number of these points. The best operation, which maximizes conversion subject to the limitations, is designated by the dashed line. The arrows on the dashed line represent the (quasi-steady-state) path followed as the inhibitor disturbance increases. The signal select implements this strategy and therefore is appropriate for this example.

Implicit in the use of the automated variable-structure methods in this chapter is the assumption that the change in structure must be made quickly, when required. If the required structural changes occur very infrequently and need not be made immediately, a simple switch could be used, and the position of the switch could be changed by a human operator. The simpler design using a switch is employed when the structure change is needed infrequently, such as during unit startups.

Variable-structure methods presented in this chapter employ single-loop controller algorithms. In this chapter, only PID controllers have been discussed; however, other algorithms, such as the model predictive controllers in Chapter 19, can be used.

The empirical model identification methods and controller tuning procedures for variable-structure systems are the same as presented previously for single-loop systems. Since the feedback path $G_{OL}(s)$ depends on the status of the variable structure, models and tuning for each feedback path must be determined and fine-tuned individually.

It is important to reiterate that the methods presented in this chapter, while simple and easy to apply, are limited to systems with low dimensionality. It is very difficult to implement a variable-structure system with many controlled *and* manipulated variables by the methods in this chapter. Fortunately, the multivariable control algorithm presented in the next chapter has the capability of practically solving high-dimension variable-structure control systems, as well as complex square control systems.

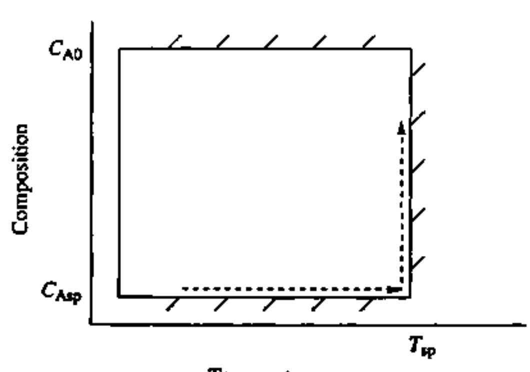


FIGURE 22.16

Operating window for reactor signal select control.

Finally, variable structure is applied widely in the process industries. It e[illegible] a process to operate with a specified (efficient) control pairing for normal opera[illegible] and to maintain acceptable operation as large changes in input variables occur or unusual set points are entered. Thus, the integration of variable-structure control as a component of the control design is essential for proper operation of many process plants.

REFERENCES

MacGregor, J., and T. Harris, "Design of Multivariable Linear-Quadratic Controllers Using Transfer Functions," *AIChE J., 33, 9*, 1481–1495 (1987).

ADDITIONAL RESOURCES

High-frequency measurement noise can influence the performance of signal selects, causing a value that is not the lowest (highest) to be selected by a low (high) signal select. This behavior can prevent a constraint control design from achieving close approach to the best possible operation. This issue is discussed and improvements suggested in the following papers.

Weber, R., and R. Zumwalt, "The Effect of Measurement Noise on Feedback Controllers with Signal Selectors," *AIChE Natl. Meet.*, April 1979.

Giles, R., and L. Gaines, "Integral-Tracking Override Is Better than Output Tracking," *Cont. Eng.*, 63–65 (1978).

Constraint control has been applied to many unit operations. Examples in distillation control are presented in the following papers:

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

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

A slightly different approach for valve position control is presented in

Love, B., "Advanced Technique Improves Two-Valve PRV Control," *Control, V*, 56–60 (1992).

Additional examples of variable-structure control are presented in

Shinskey, F. G., *Controlling Multivariable Processes*, Instrument Society of America, Research Triangle Park, NC, 1981.

> It is very helpful to describe the desired process operating policy in words first and then sketch the behavior in figures similar to Figures 22.2, 22.15, and 22.16. Follow this suggestion when designing controls for the questions in this chapter.

QUESTIONS

22.1. The three-tank mixer problem in Example 7.2 is considered here, with the slight modification that stream *B* is under flow control, with a sensor and valve added to the process. The goal is to maximize the production of material from the third tank. Limitations could be encountered in the flow rates of either stream *A* or *B*.

(*a*) Design at least one control system that would (1) control the product quality to 3% A and (2) maximize the production rate.
(*b*) Estimate the initial tuning for every controller in the design.

22.2. Prepare the detailed equations, with sequence of execution, or a sample digital control program for

(*a*) The split range controller in Figure 22.1
(*b*) The signal select system in Figure 22.5

22.3. Analyze the degrees of freedom based on models of the process and the control calculations for

(*a*) The system in Figure 22.11
(*b*) The system in Figure 22.12, both designs

22.4. Sketch the steady-state operating window and describe the path taken by the process under control for the following systems in response to selected disturbances: (*a*) Figure 22.10 (simplify this to two temperatures), (*b*) Figure 22.11, (*c*) Figure 22.12*a* and *b*, and (*d*) Figure 22.13.

22.5. For the stirred-tank heater system in Figure 22.13,

(*a*) Verify the degrees of freedom from models of the process and control.
(*b*) Specify the control calculations in analog or digital using external reset to prevent integral windup. Indicate the external variable clearly.
(*c*) Since the design includes a pairing on a zero relative gain, describe the monitoring program required. Clearly indicate the variables monitored and the actions taken when specific situations are encountered.

22.6. There are situations with excess manipulated variables in which neither variable should normally have a zero value. For example, consider the process in Figure 22.1, but with a contract that requires the plant to pay for a specified amount of fuel B (say 30% of valve opening), whether it is consumed or not. Design a control strategy that controls the pressure and provides good steady-state economic performance for this process.

22.7. Anti-reset windup is a very important aspect of successful control implementation. For the system in Figure 22.5,

(*a*) Sketch a block diagram in a continuous (analog) implementation of PI controllers with external feedback and clearly show the measurement used for the external variable.
(*b*) Provide the equations or sample program for digital control calculations, including all control elements, and an alternative method for anti-reset windup that does not have to use the external feedback principle.

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22.8. Discuss how the two proportional-integral controllers in Figure 22.[illegible] be replaced with single-loop predictive controllers, either IMC or Sm[illegible] predictors. In the discussion, provide all process and control equations, analyze the degrees of freedom, and explain how to prevent integral windup.

22.9. MacGregor and Harris (1987) describe a process that is shown schematically in Figure Q22.9. A moist film is dried using two sources of heat: an expensive electrical IR heater, which has a rapid effect on the moisture in the material, and a less costly steam heater, which has a slower response on the moisture in the material. Design a control system to provide tight control of the moisture and to minimize energy costs.

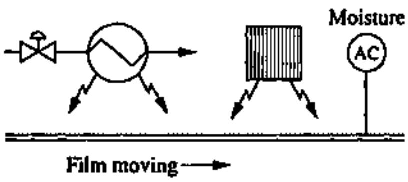

FIGURE Q22.9

22.10. Plantwide throughput maximization is certainly a good concept, but the slow dynamics between the downstream constraints and the manipulated feed flow rates could lead to extreme violations of the constraints as disturbances occur. An approach to prevent these violations is to include extra controllers that adjust manipulated variables that are "close" to the controlled variables (and have fast dynamics to the controlled variable). These *override controllers* prevent large, long constraint violations during the time required for the manipulation of the feed flow to affect the limiting plant variable. Apply this approach to the process in Figure 22.14 to prevent violations of the maximum vapor flow from the flash and the maximum light key in the bottom product of the distillation column.

22.11. For a typical level process, as in Figure 18.1, design two control systems to ensure a minimum flow through the pump. (Some process equipment changes might be required.) Discuss the merits and demerits of each and recommend one for application.

22.12. Discuss the control objectives and control design in Figure 2.2.

22.13. Using the methods described in this chapter, design a control system to maximize the production rate of vapor from the flash drum in Figure 13.19. You may add sensors but may not add valves or otherwise change the process equipment.

22.14. Discuss the steps necessary to identify linear dynamic models empirically, determine initial tuning, and fine-tune all controllers for the systems in Figures 22.1, 22.5, 22.10*a* and *b*, and 22.11.

22.15. The control design in Figure 22.11 has a deficiency, because the controllers experience a "gap" when switching between manipulated valves. Explain how this gap occurs and propose a design modification that eliminates this gap while retaining the good aspects of the original design.

22.16. For the following processes, design a variable-structure control system with a sketch, select feedback algorithms and modes, and estimate all initial tuning constants.

(*a*) The concentration of A in the effluent of the second reactor of the series chemical reactors in Example I.2 is to be controlled. The preferred method is to adjust the flow of reactant A, F_A. When this variable saturates, the flow of solvent, F_s, is adjusted.

(*b*) The bottoms composition in the distillation tower in Example 20.2 is to be controlled. The preferred choice is to manipulate the reboiler duty, but if this saturates, the reflux can be adjusted. Consider two cases: (1) the distillate composition is free to vary: (2) the distillate composition is controlled by adjusting the reflux when possible, but the bottoms purity is of overriding importance.

(*c*) The temperature of the stirred-tank heat exchanger in Example 8.5 is to be controlled. The preferred choice is adjusting the coolant flow rate, but if this saturates, the feed flow rate can be adjusted.

(*d*) The reactor outlet temperature in question 21.4 is to be controlled. The preferred manipulated variable is the set point of the quench temperature controller. If the quench temperature controller set point reaches its limit, the quench flow should be manipulated. Design a control system to satisfy the objectives, sketch the design on Figure Q21.4, and determine the outlet temperature controller tuning for all possible situations.

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