

Overview of Control System Design

General Requirements

1. *Safety.* It is imperative that industrial plants operate safely so as to promote the well-being of people and equipment within the plant and in the nearby communities. Thus, plant safety is always the most important control objective and is the subject of Section 10.5.
2. *Environmental Regulations.* Industrial plants must comply with environmental regulations concerning the discharge of gases, liquids, and solids beyond the plant boundaries.
3. *Product Specifications and Production Rate.* In order to be profitable, a plant must make products that meet specifications concerning product quality and production rate.

4. *Economic Plant Operation.* It is an economic reality that the plant operation over long periods of time must be profitable. Thus, the control objectives must be consistent with the economic objectives.
5. *Stable Plant Operation.* The control system should facilitate smooth, stable plant operation without excessive oscillation in key process variables. Thus, it is desirable to have smooth, rapid set-point changes and rapid recovery from plant disturbances such as changes in feed composition.

Steps in Control System Design

After the control objectives have been formulated, the control system can be designed. The design procedure consists of three main steps:

1. Select controlled, manipulated, and measured variables.
2. Choose the control strategy (multiloop control vs. multivariable control) and the control structure (e.g., pairing of controlled and manipulated variables).
3. Specify controller settings.

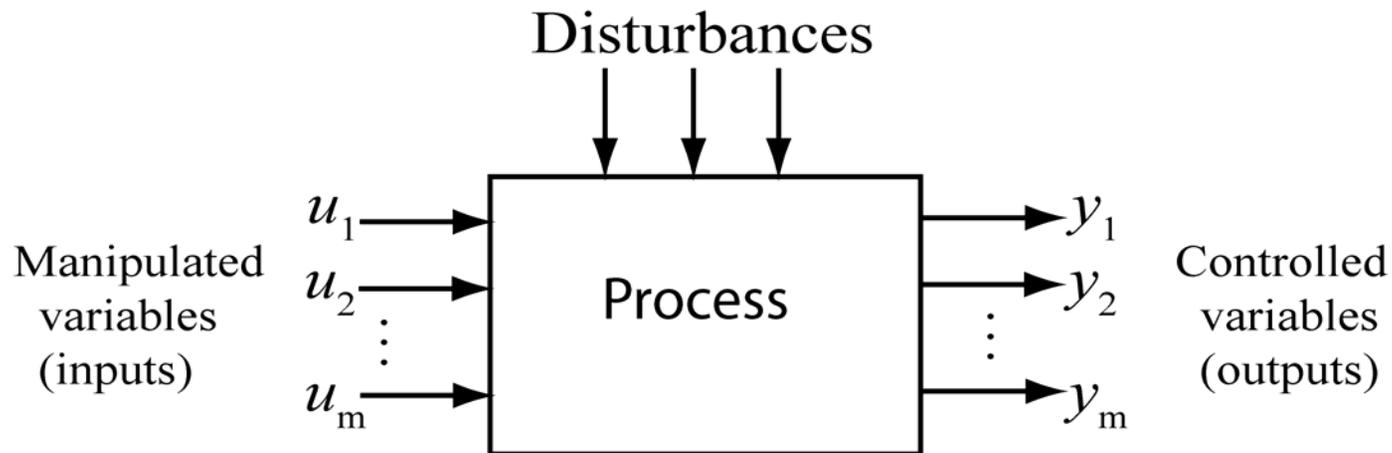
Control Strategies

- **Multiloop Control:**

Each output variable is controlled using a single input variable.

- **Multivariable Control:**

Each output variable is controlled using more than one input variable.

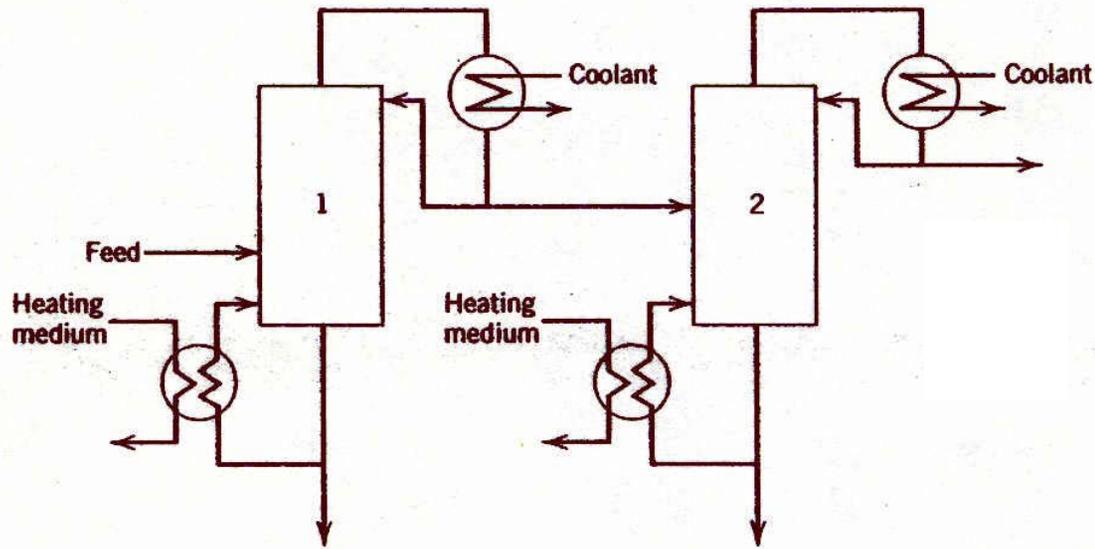


10.2 THE INFLUENCE OF PROCESS DESIGN ON PROCESS CONTROL

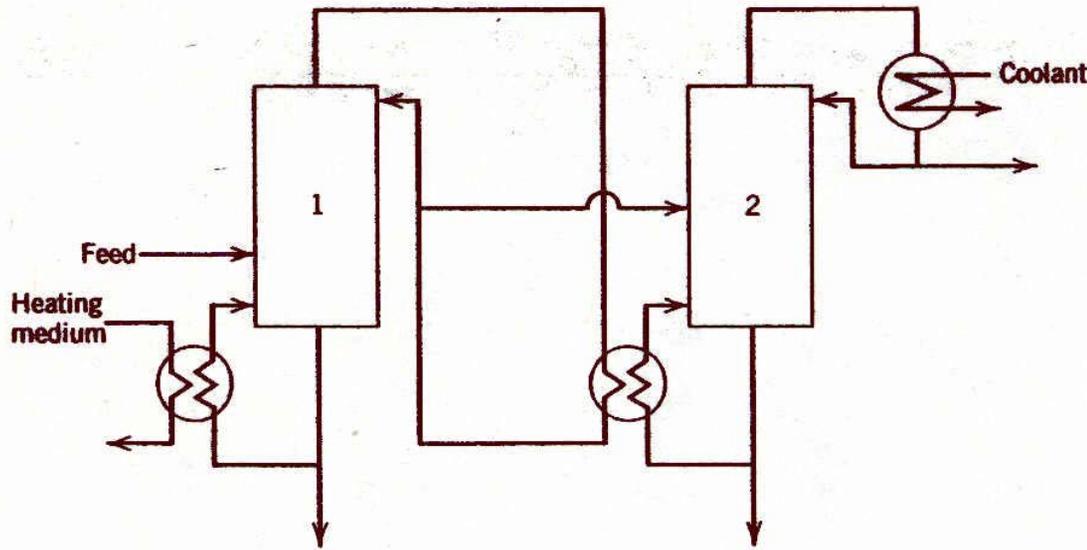
- Traditionally, process design and control system design have been separate engineering activities.
- Thus in the traditional approach, control system design is not initiated until after the plant design is well underway and major pieces of equipment may even have been ordered.
- This approach has serious limitations because the plant design determines the process dynamic characteristics, as well as the operability of the plant.
- In extreme situations, the plant may be uncontrollable even though the process design appears satisfactory from a steady-state point of view.

10.2 THE INFLUENCE OF PROCESS DESIGN ON PROCESS CONTROL (continued)

- A more desirable approach is to consider process dynamics and control issues early in the plant design.
- This interaction between design and control has become especially important for modern processing plants, which tend to have a large degree of material and energy integration and tight performance specifications.
- As Hughart and Kominek (1977) have noted: "The control system engineer can make a major contribution to a project by advising the project team on how process design will influence the process dynamics and the control structure."
- The interaction of the process design and control system design teams is considered in Chapter 23.
- Next, we consider an example of *heat integration*.

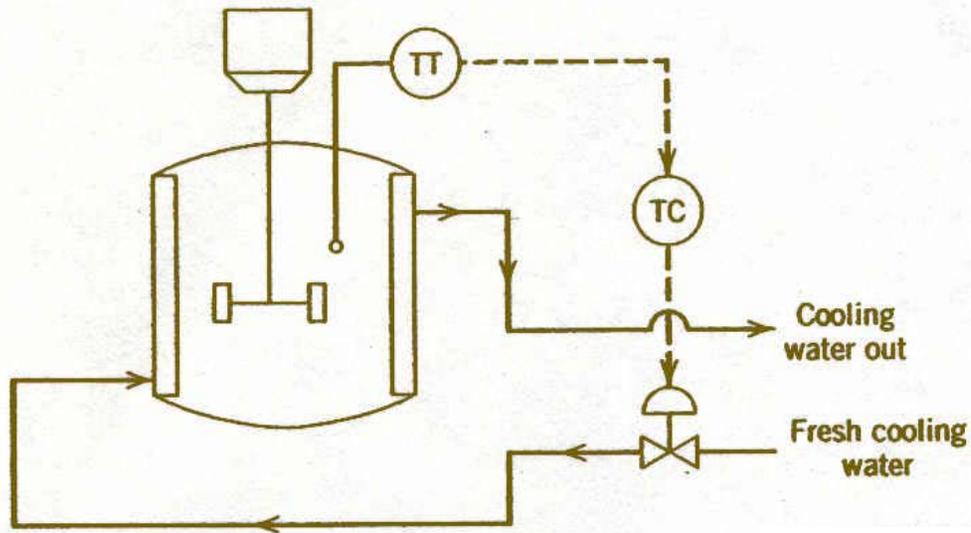


(a) Without heat integration

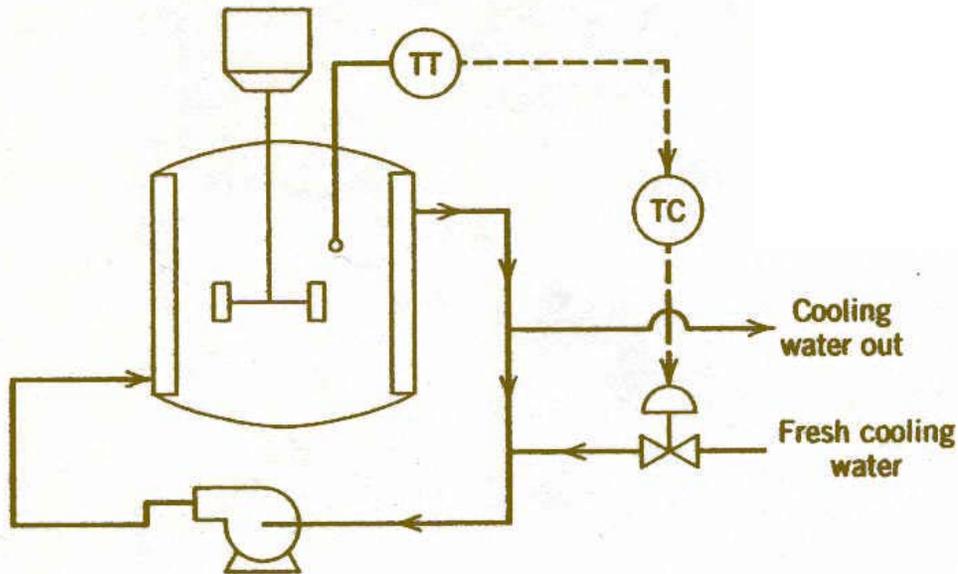


(b) With heat integration

Figure 10.1
Two
distillation
column
configurations.



(a) Original temperature control scheme



(b) Temperature control scheme with coolant recirculation pump

Figure 10.3
Batch reactor
with two
temperature
control
strategies.

10.3 Degrees of Freedom for Process Control

- The important concept of degrees of freedom was introduced in Section 2.3, in connection with process modeling.
- The degrees of freedom N_F is the number of process variables that must be specified in order to be able to determine the remaining process variables.
- If a dynamic model of the process is available, N_F can be determined from a relation that was introduced in Chapter 2,

$$N_F = N_V - N_E \quad (10-1)$$

where N_V is the total number of process variables, and N_E is the number of independent equations.

For process control applications, it is very important to determine the maximum number of process variables that can be independently controlled, that is, to determine the *control degrees of freedom*, N_{FC} :

Definition. The control degrees of freedom, N_{FC} , is the number of process variables (e.g., temperatures, levels, flow rates, compositions) that can be independently controlled.

- In order to make a clear distinction between N_F and N_{FC} , we will refer to N_F as the *model degrees of freedom* and N_{FC} as the *control degrees of freedom*.
- Note that N_F and N_{FC} are related by the following equation,

$$N_F = N_{FC} + N_D \quad (10-2)$$

where N_D is the number of disturbance variables (i.e., input variables that cannot be manipulated.)

General Rule. For many practical control problems, the control degrees of freedom N_{FC} is equal to the number of independent material and energy streams that can be manipulated.

Example 10.2

Determine N_F and N_{FC} for the steam-heated, stirred-tank system modeled by Eqs. 2-44 – 2.46 in Chapter 2. Assume that only the steam pressure P_s can be manipulated.

Solution

In order to calculate N_F from Eq. 10-1, we need to determine N_V and N_E . The dynamic model in Eqs. 2-44 to 2.46 contains three equations ($N_E = 3$) and six process variables ($N_V = 6$): T_s , P_s , w , T_i , T , and T_w . Thus, $N_F = 6 - 3 = 3$.

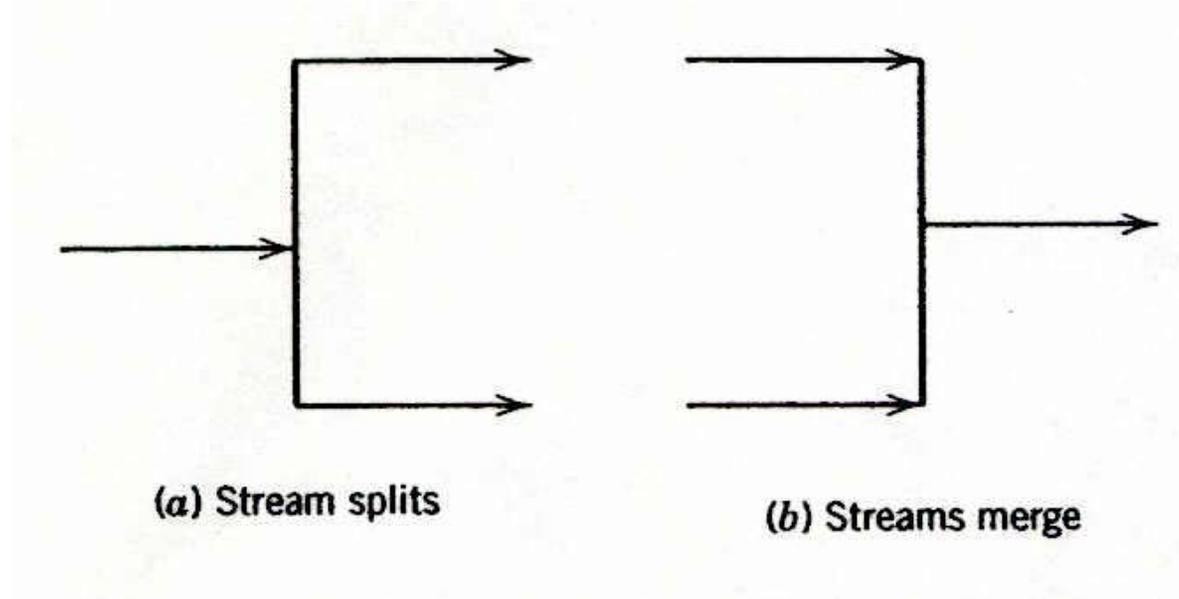


Figure 10.4 Two examples where all three process streams cannot be manipulated independently.

Stirred-Tank Heating Process

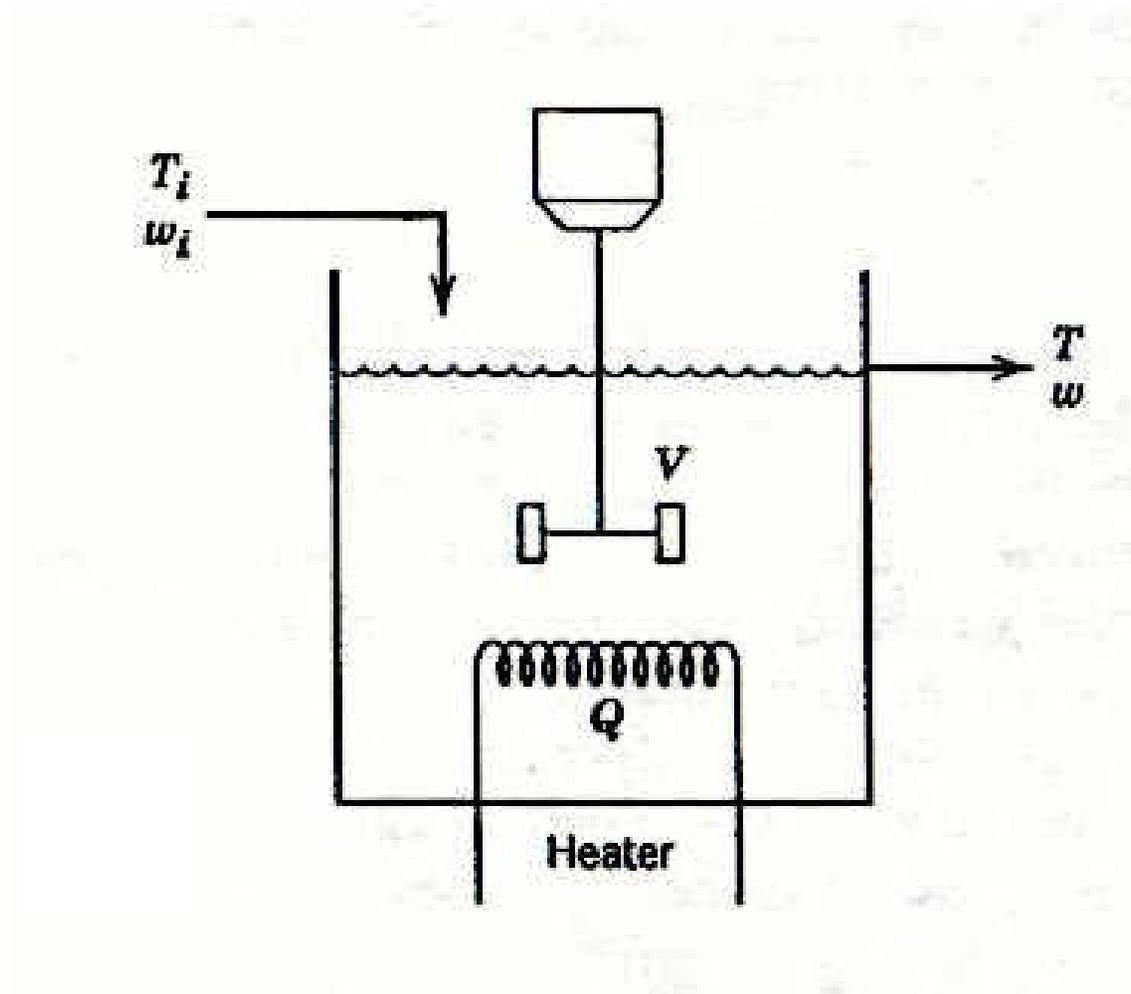


Figure 2.3 Stirred-tank heating process with constant holdup, V .

- If the feed temperature T_i and mass flow rate w are considered to be disturbance variables, $N_D = 2$ and thus $N_{FC} = 1$ from Eq. (10-2).
- It would be reasonable to use this single degree of freedom to control temperature T by manipulating steam pressure, P_s .

Example 10.4

The blending system in Fig. 10.6 has a bypass stream that allows a fraction f of inlet stream w_2 to bypass the stirred tank. It is proposed that product composition x be controlled by adjusting f via the control valve. Analyze the feasibility of this control scheme by considering its steady-state and dynamic characteristics.

In your analysis, assume that x_1 is the principal disturbance and that x_2 , w_1 , and w_2 are constant. Variations in the volume of liquid in the tank can be neglected because $w_2 \ll w_1$.

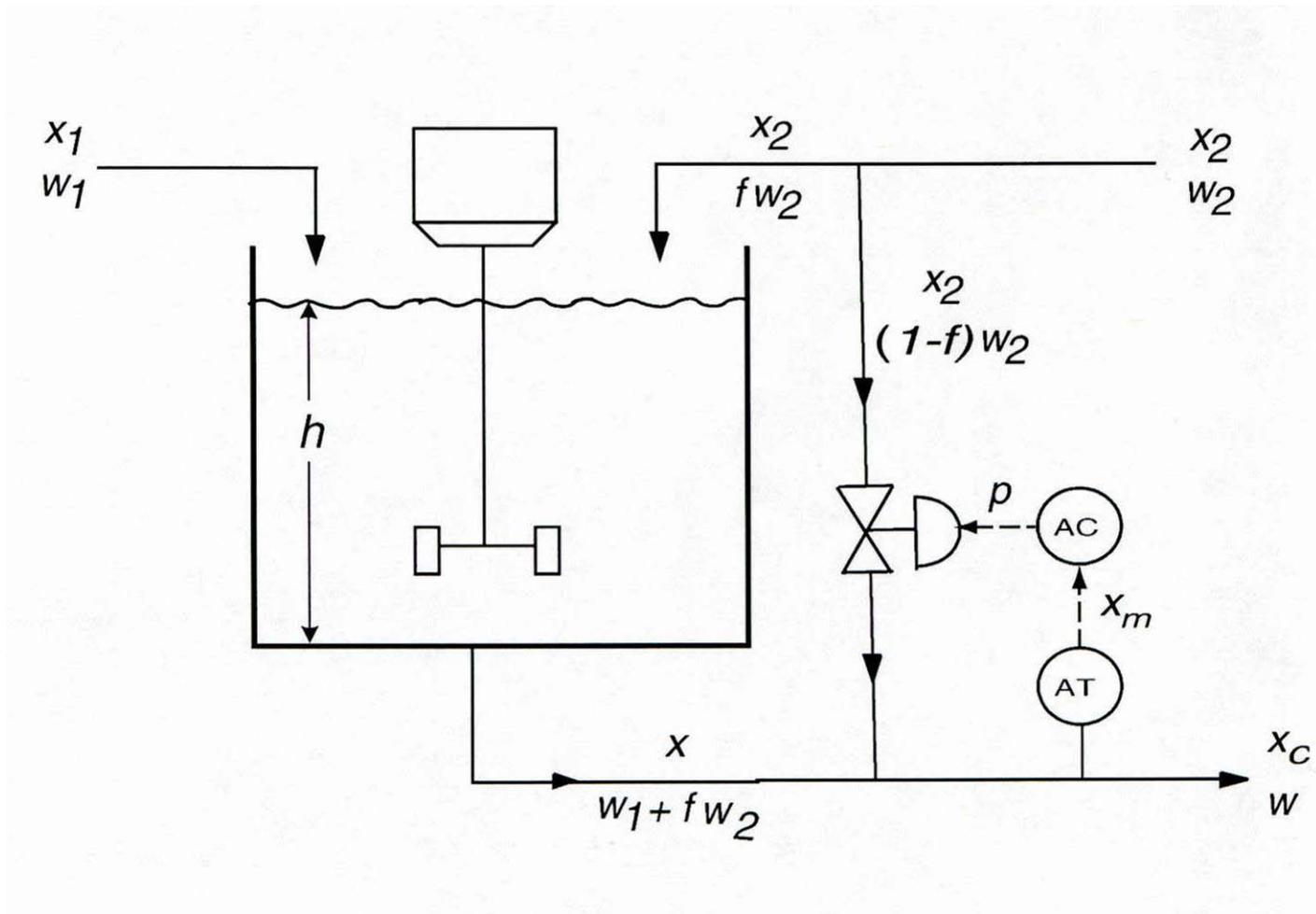


Figure 10.6. Blending system with bypass line.

Solution

- The dynamic characteristics of the proposed control scheme are quite favorable because the product composition x responds rapidly to a change in the bypass flow rate.
- In order to evaluate the steady-state characteristics, consider a component balance over the entire system:

$$w_1 \bar{x}_1 + w_2 x_2 = w \bar{x} \quad (10-3)$$

Solving for the controlled variable gives,

$$\bar{x} = \frac{w_1 \bar{x}_1 + w_2 x_2}{w} \quad (10-4)$$

- Thus \bar{x} depends on the value of the disturbance variable \bar{x}_1 and four constants (w_1 , w_2 , x_2 , and w).
- But it does not depend on the bypass function, f .

- Thus, it is *not* possible to compensate for sustained disturbances in x_1 by adjusting f .
- For this reason, the proposed control scheme is not feasible.
- Because f does not appear in (10-4), the steady-state gain between x and f is zero. Thus, although the bypass flow rate can be adjusted, it does not provide a control degree of freedom.
- However, if w_2 could also be adjusted, then manipulating both f and w_2 could produce excellent control of the product composition.

Effect of Feedback Control

- Next we consider the effect of feedback control on the control degrees of freedom.
- In general, adding a feedback controller (e.g., PI or PID) assigns a control degree of freedom because a manipulated variable is adjusted by the controller.
- However, if the controller set point is continually adjusted by a higher-level (or supervisory) control system, then neither N_F nor N_{FC} change.
- To illustrate this point, consider the feedback control law for a standard PI controller:

$$u(t) = \bar{u} + K_c \left[e(t) + \frac{1}{\tau_1} \int_0^t e(\tau) d\tau \right] \quad (10-5)$$

where $e(t) = y_{sp}(t) - y(t)$ and y_{sp} is the set point. We consider two cases:

Case 1. *The set point is constant, or only adjusted manually on an infrequent basis.*

- For this situation, y_{sp} is considered to be a parameter instead of a variable.
- Introduction of the control law adds one equation but no new variables because u and y are already included in the process model.
- Thus, N_E increases by one, N_V is unchanged, and Eqs. 10-1 and 10-2 indicate that N_F and N_{FC} decrease by one.

Case 2. *The set point is adjusted frequently by a higher level controller.*

- The set point is now considered to be a variable. Consequently, the introduction of the control law adds one new equation and one new variable, y_{sp} .
- Equations 10-1 and 10-2 indicate that N_F and N_{FC} do not change.
- The importance of this conclusion will be more apparent when cascade control is considered in Chapter 16.

Selection of Controlled Variables

Guideline 1.

All variables that are not self-regulating must be controlled.

Guideline 2.

Choose output variables that must be kept within equipment and operating constraints (e.g., temperatures, pressures, and compositions).

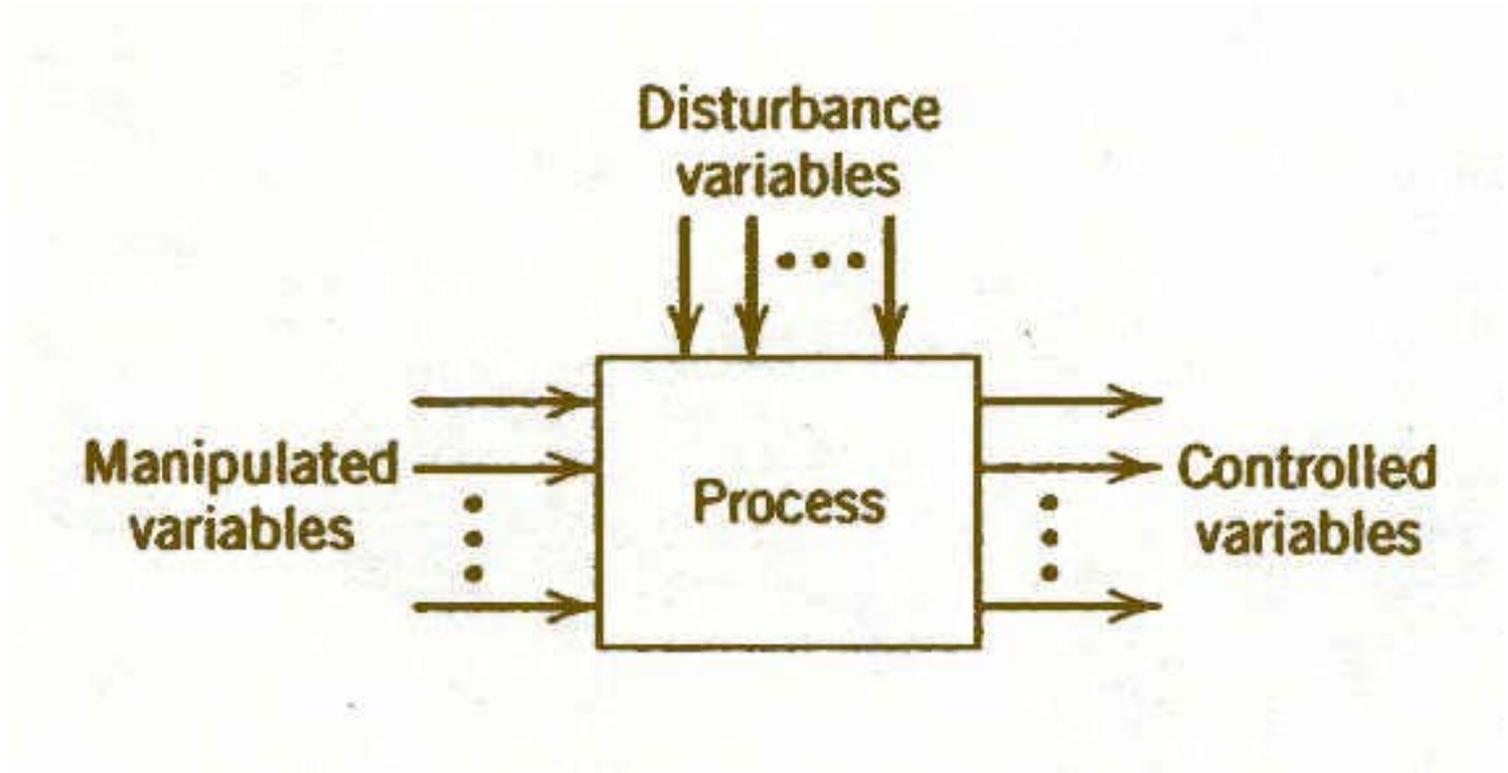


Figure 10.7 General representation of a control problem.

Guideline 3.

Select output variables that are a direct measure of product quality (e.g., composition, refractive index) or that strongly affect it (e.g., temperature or pressure).

Guideline 4.

Choose output variables that seriously interact with other controlled variables.

Guideline 5.

Choose output variables that have favorable dynamic and static characteristics.

Selection of Manipulated Variables

Guideline 6.

Select inputs that have large effects on controlled variables.

Guideline 7.

Choose inputs that rapidly affect the controlled variables.

Guideline 8.

The manipulated variables should affect the controlled variables directly rather than indirectly.

Guideline 9.

Avoid recycling of disturbances.

Selection of Measured Variables

Guideline 10.

Reliable, accurate measurements are essential for good control.

Guideline 11.

Select measurement points that have an adequate degree of sensitivity.

Guideline 12.

Select measurement points that minimize time delays and time constants

10.5 Process Safety and Process Control

- Process safety has been a primary concern of the process industries for decades.
- But in recent years, safety issues have received increased attention for several reasons that include increased public awareness of potential risks, stricter legal requirements, and the increased complexity of modern industrial plants.

Overview of Process Safety

Process safety is considered at various stages in the lifetime of a process:

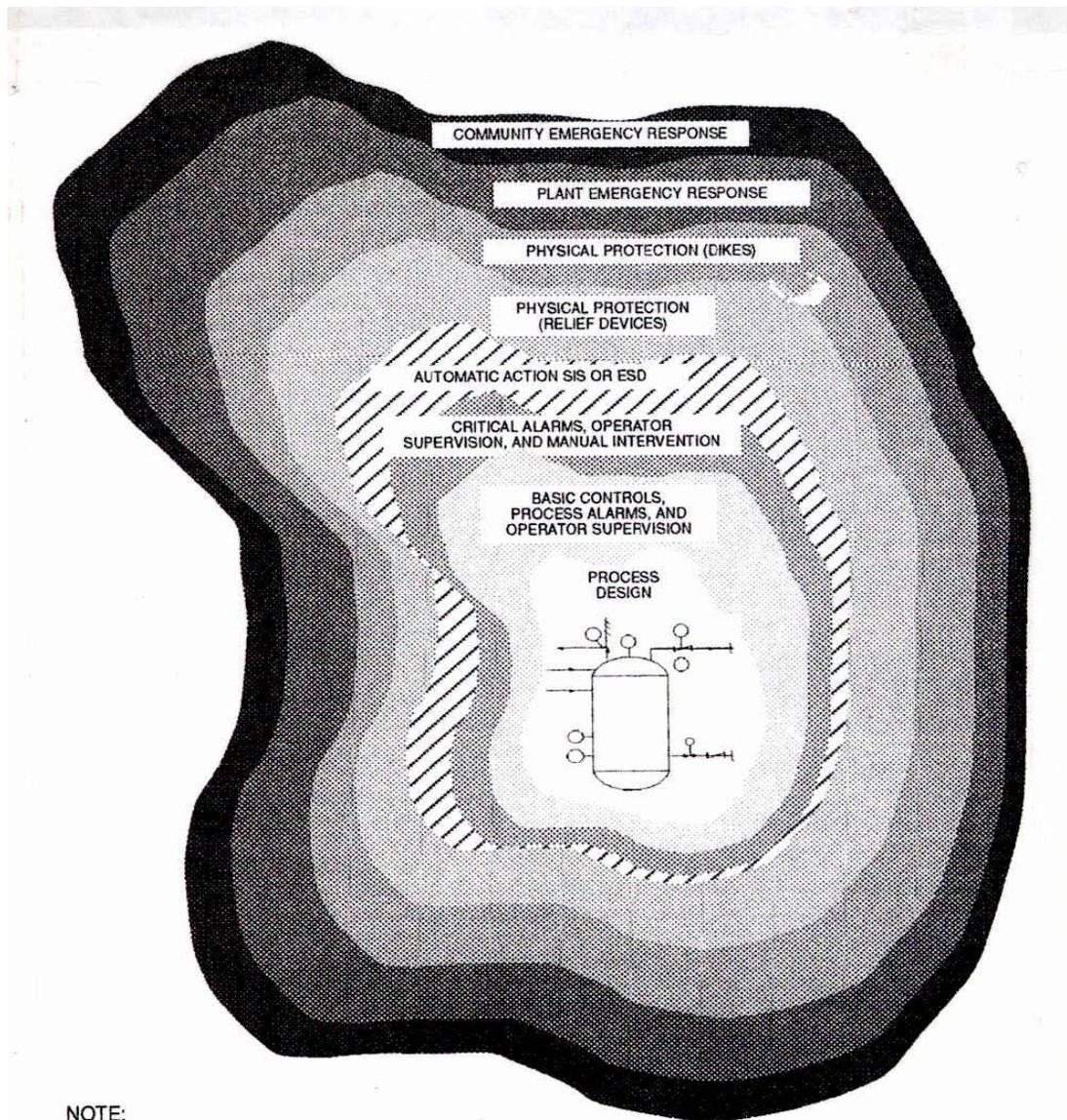
1. An initial safety analysis is performed during the preliminary process design.

2. A very thorough safety review is conducted during the final stage of the process design using techniques such as *hazard and operability* (HAZOP) studies, failure mode and effect analysis, and fault tree analysis.
3. After plant operation begins, HAZOP studies are conducted on a periodic basis in order to identify and eliminate potential hazards.
4. Many companies require that any proposed plant change or change in operating conditions require formal approval via a *Management of Change* process that considers the potential impact of the change on the safety, environment, and health of the workers and the nearby communities. Proposed changes may require governmental approval, as occurs for the U.S. pharmaceutical industry, for example.

5. After a serious accident or plant “incident”, a thorough review is conducted to determine its cause and to assess responsibility.

Multiple Protection Layers

- In modern chemical plants, process safety relies on the principle of *multiple protection layers* (AIChE, 1993b; ISA, 1996). A typical configuration is shown in Figure 10.11.
- Each layer of protection consists of a grouping of equipment and/or human actions. The protection layers are shown in the order of activation that occurs as a plant incident develops.
- In the inner layer, the process design itself provides the first level of protection.



NOTE:

Protection layers for a typical process are shown in the order of activation expected as a hazardous condition is approached.

ESD – Emergency Shutdown
SIS – Safety Interlock System

Figure 10.11. Typical layers of protection in a modern chemical plant (CCPS 1993).

- The next two layers consist of the *basic process control system* (BPCS) augmented with two levels of alarms and operator supervision or intervention.
- An alarm indicates that a measurement has exceeded its specified limits and may require operator action.
- The fourth layer consists of a *safety interlock system* (SIS) that is also referred to as a *safety instrumented system* or as an *emergency shutdown* (ESD) system.
- The SIS automatically takes corrective action when the process and BPCS layers are unable to handle an emergency. For example, the SIS could automatically turn off the reactant pumps after a high temperature alarm occurs for a chemical reactor.

- Relief devices such as rupture discs and relief valves provide physical protection by venting a gas or vapor if over-pressurization occurs.
- As a last resort, dikes are located around process units and storage tanks to contain liquid spills.
- Emergency response plans are used to address emergency situations and to inform the community.

Types of Alarms

Type 1 Alarm: Equipment status alarm. Indicates equipment status, for example, whether a pump is on or off, or whether a motor is running or stopped.

Type 2 Alarm: Abnormal measurement alarm. Indicates that a measurement is outside of specified limits.

Type 3 Alarm: An alarm switch without its own sensor. These alarms are directly activated by the process, rather than by a sensor signal. Type 3 alarms are used for situations where it is not necessary to know the actual value of the process variable, only whether it is above (or below) a specified limit.

Type 4 Alarm: An alarm switch with its own sensor. A type 4 alarm system has its own sensor that serves as a backup in case the regular sensor fails.

Type 5 Alarm: Automatic Shutdown or Startup System. These important and widely used systems are described in the next section on Safety Interlock Systems.

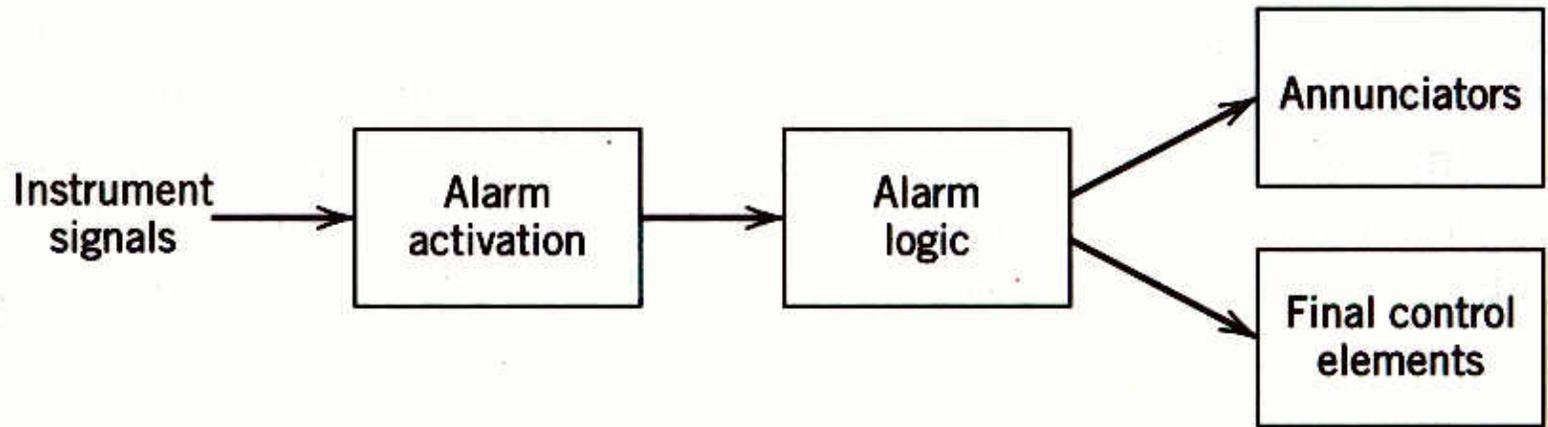


Fig. 10.12 A general block diagram for an alarm system.

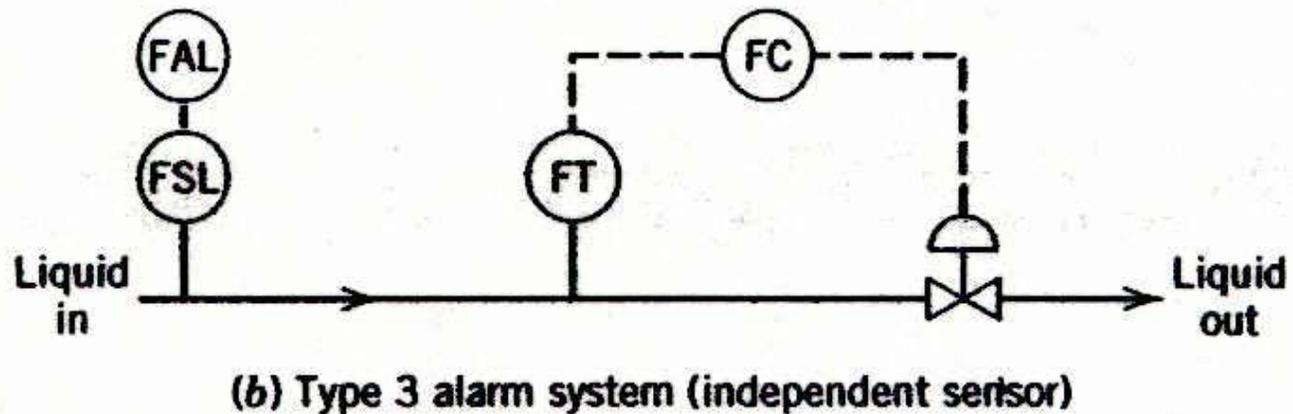
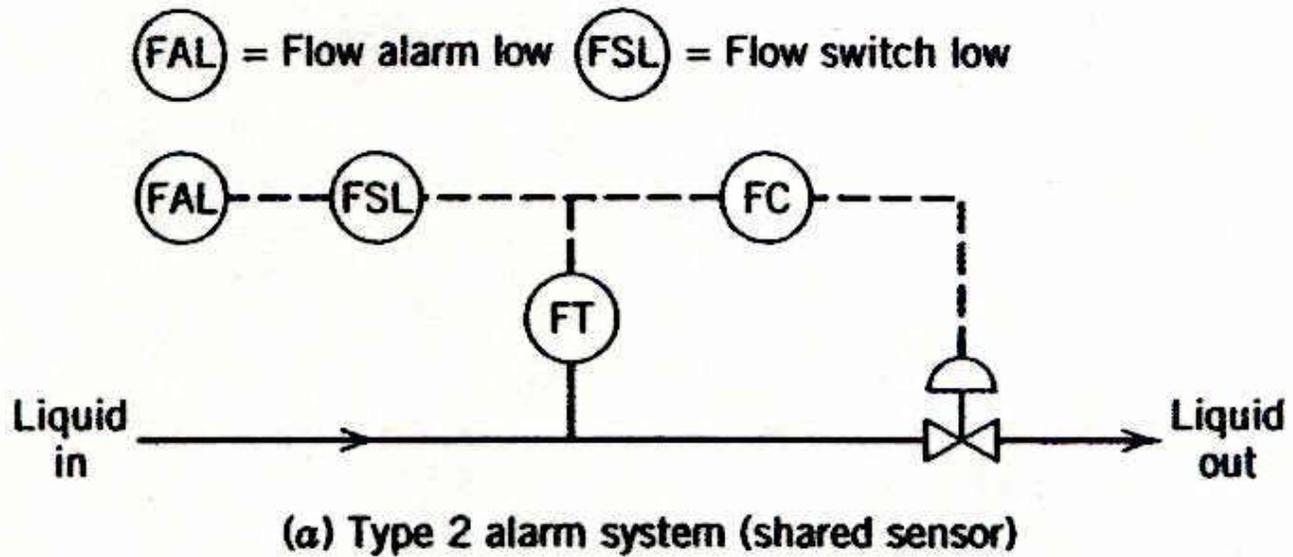
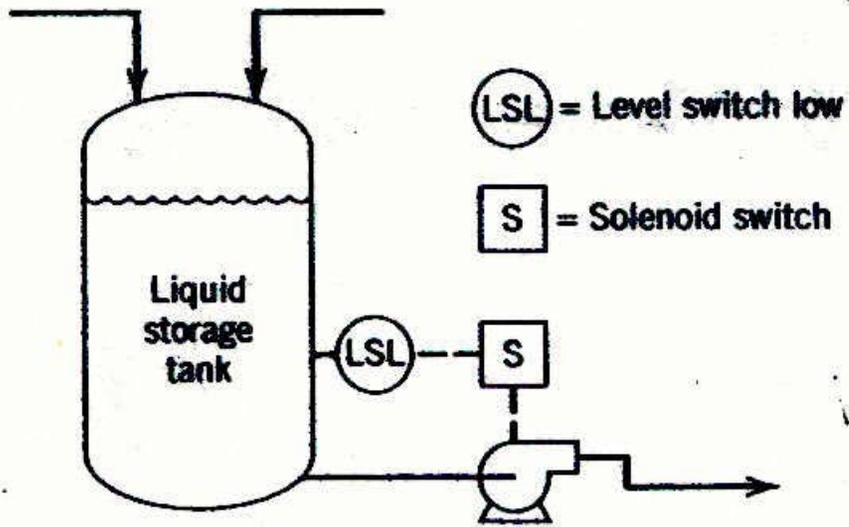
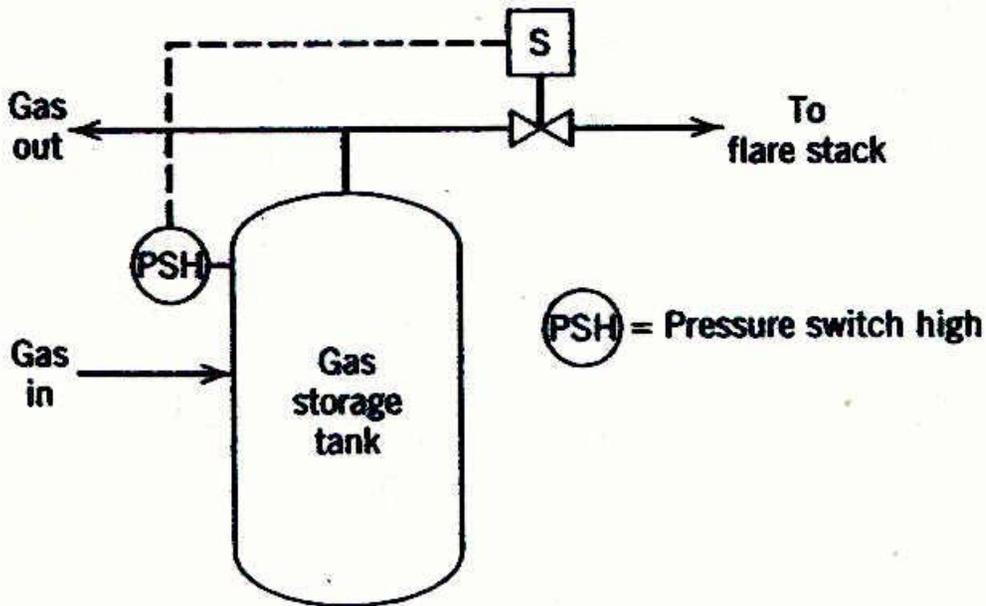


Fig. 10.13 Two flow alarm configurations.



(a) Low-level interlock



(b) High-pressure interlock

Fig. 10.14 Two interlock configurations.

Safety Interlock System (SIS)

- The SIS in Figure 10.11 serves as an emergency back-up system for the BPCS.
- The SIS automatically starts when a critical process variable exceeds specified alarm limits that define the allowable operating region.
- Its initiation results in a drastic action such as starting or stopping a pump or shutting down a process unit.
- Consequently, it is used only as a last resort to prevent injury to people or equipment.

- It is very important that the SIS function *independently* of the BPCS; otherwise, emergency protection will be unavailable during periods when the BPCS is not operating (e.g., due to a malfunction or power failure).
- *Thus, the SIS should be physically separated from the BPCS (AIChE, 1993b) and have its own sensors and actuators.*

A Final Thought...

As Rinard (1990) has poignantly noted, “The regulatory control system affects the size of your paycheck; the safety control system affects whether or not you will be around to collect it.”