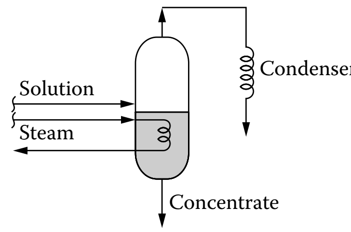
8.23 Evaporator Controls

T. J. MYRON (1972, 1985) **B. G. LIPTÁK** (1995, 2005)



Flow sheet symbol

INTRODUCTION

Evaporation is used in a number of unit operations. For example, one of the important steps in the operation of a chiller is the evaporation of the refrigerant, which usually occurs at low pressures. The controls of evaporators serving such heat pumps are described in Sections 8.12 and 8.13. If evaporation occurs at higher pressures, we refer to that unit process as vaporization or reboiling and the controls of those processes are described in Sections 8.19, 8.20, 8.21, and 8.27. The evaporation controls that are discussed in this section are for water-based near-atmospheric processes.

Evaporation is one of the oldest unit operations, dating back to the Middle Ages, when any available energy source was used to concentrate thin brine solutions in open tanks. As archaic as this technique is, it is still widely used during the early spring in northern New England when maple sugar sap is tapped and concentrated in open-fired pans.

Solar evaporation and bubbling hot gases through a solution are other examples of concentrating solutions. For our purposes, evaporation will be limited to concentrating aqueous solutions in a closed vessel or group of vessels in which the concentrated solution is the desired product and indirect heating (usually steam) is the energy source. Occasionally, the water vapor generated in the evaporator is the product of interest, such as in desalinization or in the production of boiler feed water. In other cases neither vapor nor concentrated discharge has any market value, as in nuclear wastes.

Evaporators can be arranged in forward-feed, reverse-feed, or parallel-feed configurations, with each stage being heated by the vapors of the previous stage. Evaporation is an energy-intensive process. Its efficiency can be improved by increasing the number of evaporators in series (effects), and some of the energy can be recovered by vapor recompression.

The product concentration can be measured by a variety of analyzers, including density, conductivity, refractive index, percent solids, turbidity, and boiling or freezing point analyzers, which are all discussed in Chapter 8 of the first volume of this handbook, *Process Measurement and Analysis*.

Evaporator Terminology

SINGLE-EFFECT EVAPORATION Single-effect evaporation occurs when a dilute solution is contacted only once with a heat source to produce a concentrated solution and an essentially pure water vapor discharge. The operation is shown schematically in Figure 8.23a.

MULTIPLE-EFFECT EVAPORATION Multiple-effect evaporations use the vapor generated in one effect as the energy source to an adjacent effect (Figure 8.23b). Double- and triple-effect evaporators are the most common; however, six-effect evaporation can be found in the paper industry, where kraft liquor is concentrated, and as many as 20 effects can be found in desalinization plants.

BOILING-POINT RISE This term expresses the difference (usually in °F) between the boiling point of a constant composition solution and the boiling point of pure water at the same pressure. For example, pure water boils at 212°F (100°C) at 1 atmosphere, and a 35% sodium hydroxide solution boils at about 250°F (121°C) at 1 atmosphere. The boiling-point

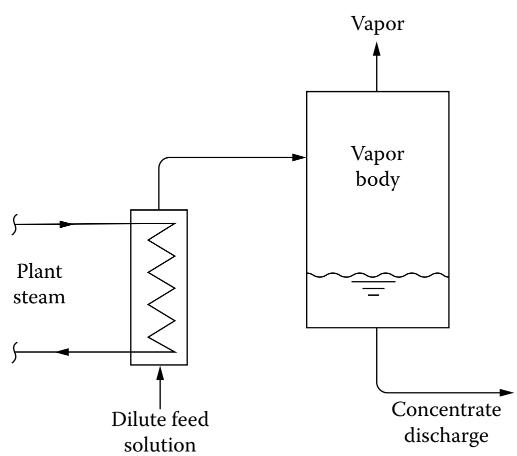


FIG. 8.23a
Single-effect evaporator.

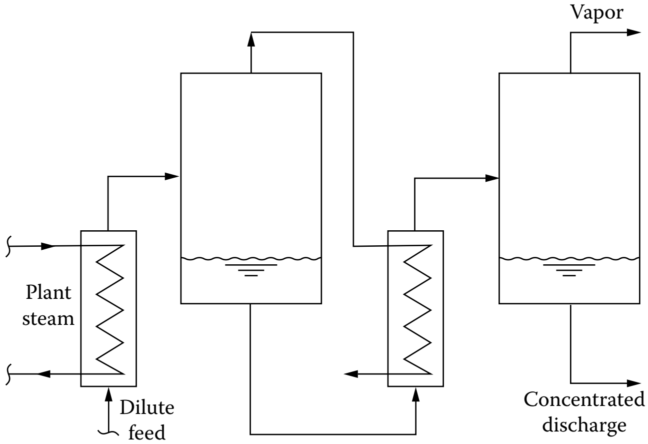


FIG. 8.23b
Multiple-effect evaporator.

rise is therefore 38°F (21°C). Figure 8.23c illustrates the features of a Dühring plot in which the boiling point of a given composition solution is plotted as a function of the boiling point of pure water.

ECONOMY This term is a measure of steam use and is expressed in pounds of vapor produced per pound of steam supplied to the evaporator train. For a well-designed evaporator system the economy will be about 10% less than the number of effects; thus, for a triple-effect evaporator the economy will be roughly 2.7.

CAPACITY The capacity for an evaporator is measured in terms of its evaporating capability, viz., pounds of vapor produced per unit time. The steam requirements for an evaporating train may be determined by dividing the capacity by the economy.

CO-CURRENT OPERATION The feed and steam follow parallel paths through the evaporator train.

COUNTERCURRENT OPERATION The feed and steam enter the evaporator train at opposite ends.

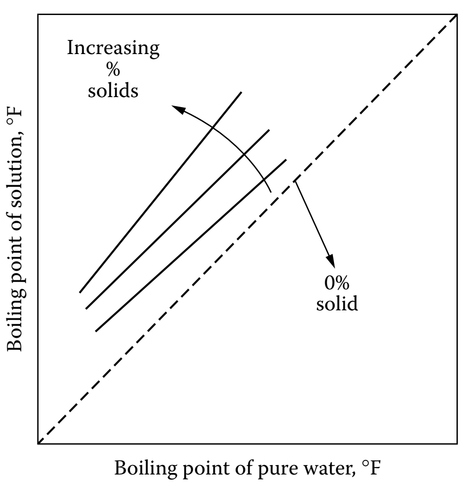


FIG. 8.23c
Dühring plot of boiling point rise.

EVAPORATOR MODELING

The Steady-State Model

Development of the steady-state model for an evaporator involves material and energy balances. A relationship between the feed density and percentage of solids is also required and is specific for a given process, whereas the material and energy balances are applicable to all evaporator processes. Figure 8.23d illustrates the double-effect evaporator from the standpoint of a material balance.

where

- W_o = feed rate in lb (kg) per unit time
- V_o = feed flow in gpm or lpm
- V_1 = vapor flow from Effect I in lb (kg) per unit time
- X_o = weight fraction solids in feed
- W_1 = liquid flow rate leaving Effect I in lb (kg) per unit time
- X_1 = weight fraction solids in W_1
- V_2 = vapor flow from Effect II in lb (kg) per unit time
- W_p = concentrated liquid product flow in lb (kg) per unit time
- X_p = weight fraction of solids in product (the controlled variable)

Overall balance in Effect I:

$$W_o = V_1 + W_1 \tag{8.23(1)}$$

Overall balance in Effect II:

$$W_1 = V_2 + W_p \tag{8.23(2)}$$

Solid balance in Effect I:

$$W_oX_o = W_1X_1 \tag{8.23(3)}$$

Solid balance in Effect II:

$$W_1X_1 = W_pX_p \tag{8.23(4)}$$

Substituting Equation 8.23(2) in Equation 8.23(1):

$$W_o = V_1 + V_2 + W_p \tag{8.23(5)}$$

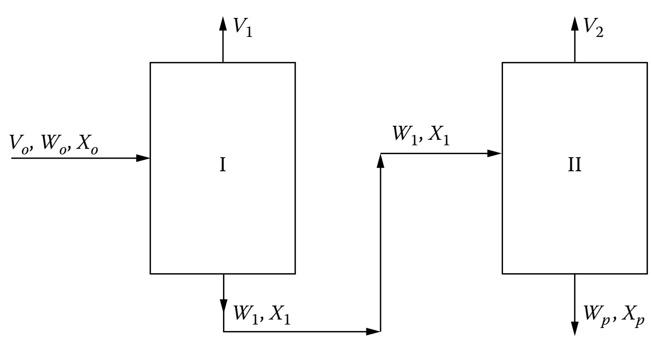


FIG. 8.23d
The material balance of a double-effect evaporator.

Combining Equations 8.23(3) and 8.23(4):

$$W_oX_o = W_pX_p \qquad \textbf{8.23(6)}$$

Solving Equation 8.23(6) for W_p and substituting in Equation 8.23(5) gives

$$W_o = V_1 + V_2 + \frac{W_oX_o}{X_p} \qquad \textbf{8.23(7)}$$

The W_o term of Equation 8.23(7) can be written in terms of volumetric flow (gallons per unit time) times density or specific gravity as follows:

$$W_o = V_oD_o = V_oD_wS_o \qquad \textbf{8.23(8)}$$

where
- V_o = volumetric feed rate in gallons per unit time
- D_o = fed density in lb per gallon
- D_w = nominal density of water, 8.33 lb per gallon (1 kg/l)
- S_o = specific gravity of feed

Substituting for W_o from Equation 8.23(8) in Equation 8.23(7) and combining terms,

$$V_oD_wS_o\left(1-\frac{X_o}{X_p}\right) = V_1 + V_2 = V_t \qquad \textbf{8.23(9)}$$

where V_t = total vapor flow in lb per unit time.

The total vapor flow (V_t) is proportional to the energy supplied to the train (plant steam) and the proportionality constant is the economy (E) of the system, i.e.,

$$V_t = W_sE \qquad \textbf{8.23(10)}$$

where
- W_s = steam flow in lbs steam per unit time
- E = economy in lbs vapor per lb steam

Substituting for V_t in Equation 8.23(9):

$$V_oD_wS_o\left(1-\frac{X_o}{X_p}\right) = W_sE \qquad \textbf{8.23(11)}$$

Equation 8.23(11) is the steady-state model of the process and includes all of the load variables (V_o and X_o), the manipulated variable (W_s), and the controlled variable (X_p). At this point the W_sE portion of Equation 8.23(11) may be modified to include heat losses from the system and to include the fact that the feed may be subcooled. These are two forms of heat losses, because in either case a portion of the steam supplied is for purposes other than producing vapor. Typical values of effective heat losses vary from 3 to 5%. If, for example, a 5% heat loss is assumed, Equation 8.23(11) becomes

$$V_oD_wS_o\left(1-\frac{X_o}{X_p}\right) = 0.95W_sE \qquad \textbf{8.23(12)}$$

TABLE 8.23e
Weight Fraction and Specific Gravity Relationship

Solids Weight Fraction (X_o)	*Specific Gravity (S_o)*
0.08	1.0297
0.16	1.0633
0.24	1.0982

For an in-depth discussion relating to methods of computing the economy (E) of a particular evaporator system, see References 2 and 3.

The $S_o(1 - X_o/X_p)$ portion of Equation 8.23(11) is a function of the feed density, $f(D_o)$, i.e.,

$$f(D_o) = S_o\left(1-\frac{X_o}{X_p}\right) \qquad \textbf{8.23(13)}$$

For each feed material a relationship between the density of the feed material and its solids weight fraction has usually been empirically determined by the plant or is available in the literature. See Reference 4 where the density equals percentage of solids relationship of 70 inorganic compounds is available.

Assume, for example, that a feed material (to be concentrated) has the solids–specific-gravity relationship shown in Table 8.23e.

If this feed material were to be concentrated so as to produce a product having a weight fraction of 50% ($X_p = 0.50$), the $f(D_o)$ relationship of Equation 8.23(13) could be generated as shown in Table 8.23f.

This body of data is plotted in Figure 8.23g. In all the cases investigated, the $f(D_o) = S_o$ relationship is a straight line having an intercept of 1.0, 1.0.

The $f(D_o)$ relationship can then be written in terms of the equation of a straight line: $y = mx + b$, i.e.,

$$f(D_o) = 1.0 + m(S_o - 1.0) \qquad \textbf{8.23(14)}$$

where m = slope of line.

TABLE 8.23f
Density to Weight Fraction Relationship

Solids Weight Fraction (X_o)	*Specific Gravity (S_o)*	$\left(1-\frac{X_o}{X_p}\right)$	$S_o\left(1-\frac{X_o}{X_p}\right) = f(D_0)$
0	1.0000	1.0	1.0000
0.08	1.0297	0.840	0.865
0.16	1.0633	0.680	0.723
0.24	1.0982	0.520	0.571

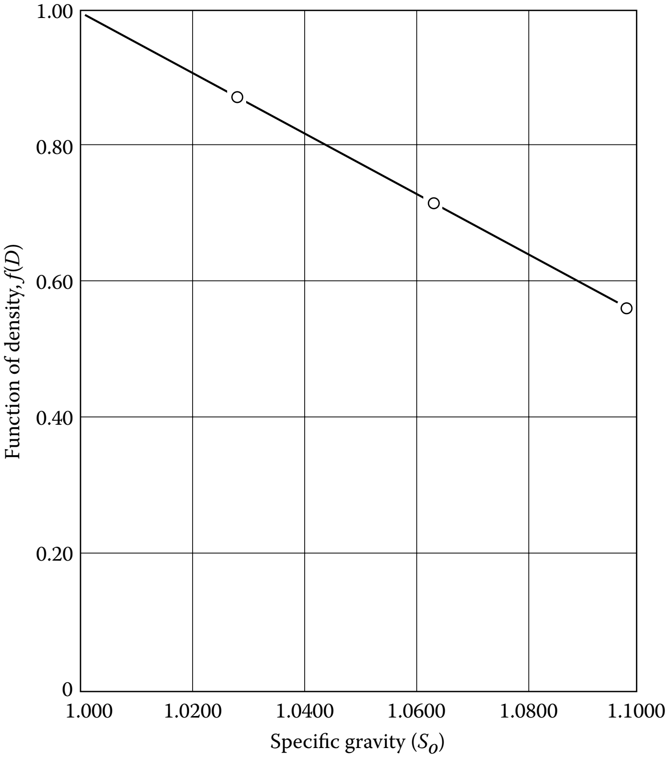


FIG. 8.23g
The straight-line relationship between density and specific gravity.

Using the data of Table 8.23f the value of m is determined as

$$m = \frac{1.000 - 0.571}{1.0000 - 1.0982} = -4.37 \qquad \mathbf{8.23(15)}$$

therefore,

$$f(D_o) = 1.0 - 4.37(S_o - 1.0) \qquad \mathbf{8.23(16)}$$

Substituting Equation 8.23(16) into Equation 8.23(11) and solving for the manipulated variable, steam flow, W_s:

$$W_s = \frac{V_o D_w f(D)}{E} \qquad \mathbf{8.23(17)}$$

Therefore, W_s can be expressed as

$$W_s = \frac{V_o[1 - m(D-1)]}{E} \qquad \mathbf{8.23(18)}$$

If feed rate is manipulated in response to a variable steam flow and feed density, Equation 8.23(18) becomes

$$V_o = \frac{W_s E}{1 - m(D-1)} \qquad \mathbf{8.23(19)}$$

Scaling and Normalizing

With the equation for the steady-state model defined, it can now be scaled and the digital or analog instrumentation specified. Scaling of computing instruments or software is necessary to ensure compatibility with input and output signals and is accomplished most effectively by normalizing, i.e., assigning values from 0 to 1.0 to all inputs and outputs. The procedure involves (1) writing the engineering equation to be solved, (2) writing a normalized equation for each variable in the engineering equation, and (3) substituting the normalized equivalent of each term in (2) into (1).

The first step of the procedure has already been done, because Equation 8.23(17) has already been written. To illustrate, let $V_o = 0$ to 600 gph (0 to 2.3 m^3/h), $W_s = 0$ to 2500 lb/hr (0 to 1125 kg/h), S_o = 1.000 to 1.1000, E = 1.8 lb vapor per lb of steam (1.8 kg vapor per kg of steam), and D_w = 8.33 lb/gallon (1 kg/l). The scaled equations for each input are

$$V_o = 600\, V_o' \qquad \mathbf{8.23(20)}$$

$$W_s = 2500\, W_s' \qquad \mathbf{8.23(21)}$$

$$S_o = 1.0000 + 0.10000 S_o' \qquad \mathbf{8.23(22)}$$

where
V_o' = volumetric flow transmitter output, 0–1.0 or 0–100%
W_s' = steam flow transmitter output, 0–1.0 or 0–100%
S_o' = specific gravity transmitter output, 0–1.0 or 0–100%

The values of D_w and E need not be scaled, because they are constants. Because $f(D_o)$ is already on a 0 to 1.0 basis, the $f(D_o)$ term for the sake of completeness can be written

$$f(D_o) = 1.0 f(D_o)' \qquad \mathbf{8.23(23)}$$

Operating on the $f(D_o)$ equation first, Equations 8.23(23) and 8.23(22) are substituted into Equation 8.23(16):

$$(D_o)' = 1.0 - 4.37 \times (1.0000 + 0.1000 S_o - 1.0) \qquad \mathbf{8.23(24)}$$

$$(D_o)' = 1.0 - 4.37 S_o' \qquad \mathbf{8.23(25)}$$

Substituting Equations 8.23(20), 8.23(21), and 8.23(25) into Equation 8.23(17) as well as the values of E and D_w,

$$2500 W_s' = \frac{600(V_o')8.33(1.0 - 0.437 S_o')}{1.8} \qquad \mathbf{8.23(26)}$$

$$\begin{aligned} W_s' &= 1.11 V_o'(1.0 - 0.437 S_o') \\ &= 1.11 V_o' f(D_o)' \end{aligned} \qquad \mathbf{8.23(27)}$$

or if feed flow is the manipulated variable:

$$V_o' = \frac{W_s'}{1.11(1 - 0.437 S_o')} \qquad \mathbf{8.23(28)}$$

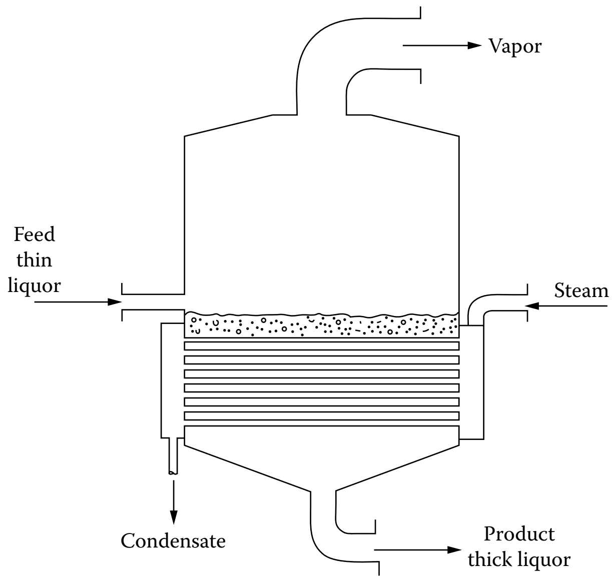


FIG. 8.23h
Horizontal-tube evaporator.

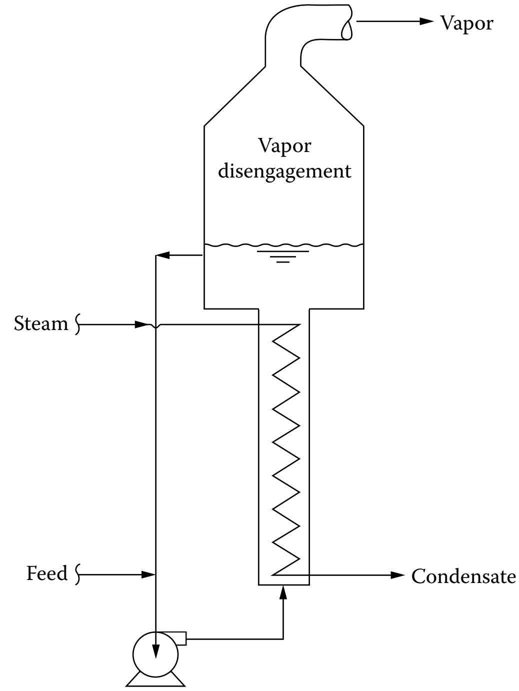


FIG. 8.23i
Forced-circulation evaporator.

EVAPORATOR DESIGNS

Six types of evaporators are used for most applications, and for the most part the length and orientation of the heating surfaces determine the name of the evaporator.

Horizontal-tube evaporators (Figure 8.23h) were among the earliest types. Today, they are limited to preparation of boiler feed water and, in special construction (at high cost), for small-volume evaporation of severely scaling liquids, such as hard water. In their standard form they are not suited to scaling or salting liquids and are best used in applications requiring low throughputs.

Forced-circulation evaporators (Figure 8.23i) are the most popular and have the widest applicability. Circulation of the liquor past the heating surfaces is assured by a pump, and consequently these evaporators are frequently external to the flash chamber so that actual boiling does not occur in the tubes, thus preventing salting and erosion. The external tube bundle also lends itself to easier cleaning and repair than the integral heater shown in the figure. Disadvantages include high cost, high residence time, and high operating costs due to the power requirements of the pump.

A short-tube vertical evaporator (Figure 8.23j) is common in the sugar industry for concentrating cane sugar juice. Liquor circulation through the heating element (tube bundle) is by natural circulation (thermal convection).

Because in the short-tube vertical evaporator the mother liquor flows through the tubes, they are much easier to clean than those shown in Figure 8.23h, in which the liquor is outside the tubes. Thus, this evaporator is suitable for mildly scaling applications in which low cost is important and cleaning or descaling must be conveniently handled. Level control is important — if the level drops below the tube ends, excessive scaling results. Ordinarily, the feed rate is controlled by evaporator level to keep the tubes full. The disadvantage of high residence time in the evaporator is compensated for by the low cost of the unit for a given evaporator load.

A long-tube vertical evaporator, or rising film concentrator (RFC), shown in Figure 8.23k, is in common use today,

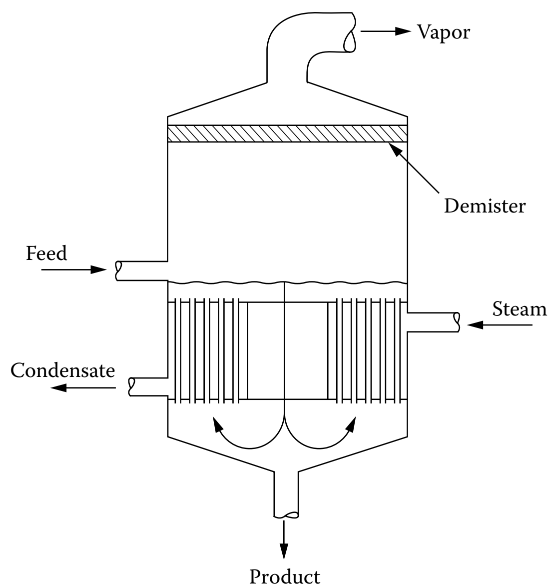


FIG. 8.23j
Short-tube vertical evaporator.

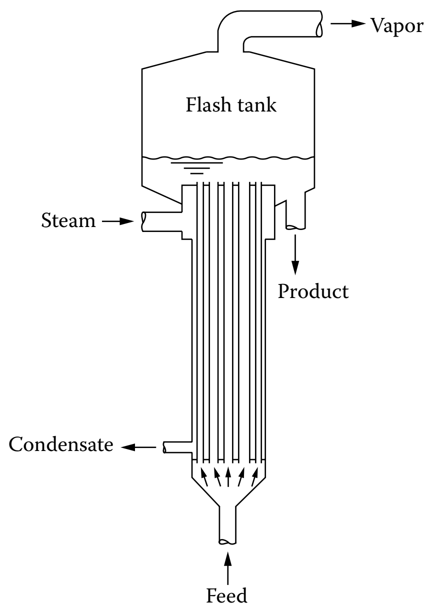


FIG. 8.23k
Long-tube vertical evaporator.

because its cost per unit of capacity is low. Typical applications include concentrating black liquor in the pulp and paper industry and corn syrup in the food industry.

Most of these evaporators are of the single-pass variety, with little or no internal recirculation. Thus, residence time is minimized. Level control is important in maintaining the liquid seal in the flash tank. The units are sensitive to changes in operating conditions, which is why many of them are difficult to control. They offer low cost per pound of water evaporated and have low holdup times but tend to be tall (20 to 50 ft, or 6 to 15 m), requiring more head room than other types.

A falling-film evaporator (Figure 8.23l) is commonly used with heat-sensitive materials. Physically, the evaporator looks like a long-tube vertical evaporator, except that the feed material descends by gravity along the inside of the heated tubes, which have large inside diameters (2 to 10 in., or 50 to 250 mm).

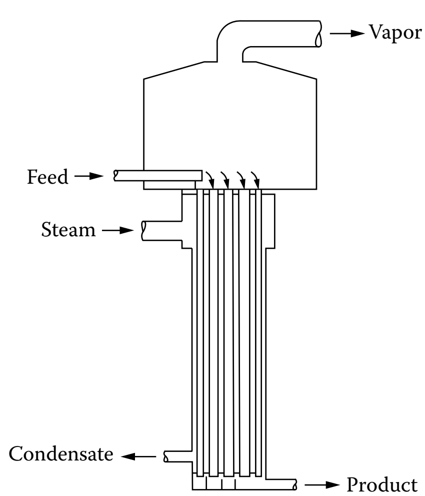


FIG. 8.23l
Falling-film vertical evaporator.

An agitated-film evaporator, like the falling-film evaporator, is commonly used for heat-sensitive and highly viscous materials. It consists of a single large-diameter tube with the material to be concentrated falling in a film down the inside, where a mechanical wiper spreads the film over the inside surface of the tube. Thus, a large heat-transfer coefficient can be obtained, particularly with highly viscous materials.

EVAPORATOR CONTROLS

In the following paragraphs the load variable will be assumed to be the flow rate and concentration of the feed stream. A later paragraph, Other Control Loops, will discuss the control of other variables, such as steam enthalpy, material balance, and absolute pressure controls.

The control systems to be considered in achieving final product concentration include (1) feedback, (2) cascade, and (3) feedforward, (4) auto-select, and (5) advanced controls. For ease of illustration, a double-effect, co-current flow evaporator will be used. Extension to more or fewer effects will not change the basic control system configuration.

The choice of the control system should be based on the needs and characteristics of the process. Evaporators as a process class tend to be capacious (mass and energy storage capability) and have significant dead time (30 sec or greater). If the major process loads (feed rate and feed density) are reasonably constant and the only corrections required are for variations in heat losses or tube fouling, feedback control will suffice.

If steam flow varies because of demands elsewhere in the plant, a cascade configuration will probably be the proper choice. If, however, the major load variables change rapidly and frequently, it is strongly suggested that feedforward in conjunction with feedback be considered.

Selective control can be superimposed on either control configuration, which will stay inactive until a limit is reached (such as running out of steam), at which point control is transferred to keep the operation (the demand for steam) under that limit. Advanced controls usually combine all the previous features and add to it a multivariable model-based predictive capability.

Feedback Control

A typical feedback control system (Figure 8.23m) consists of measuring the product concentration with a density sensor and controlling the amount of steam to the first effect by a three-mode controller. The internal material balance is maintained by level control on each effect. (A brief description of the various methods of measuring product density will be found at the end of this section; additional discussion regarding

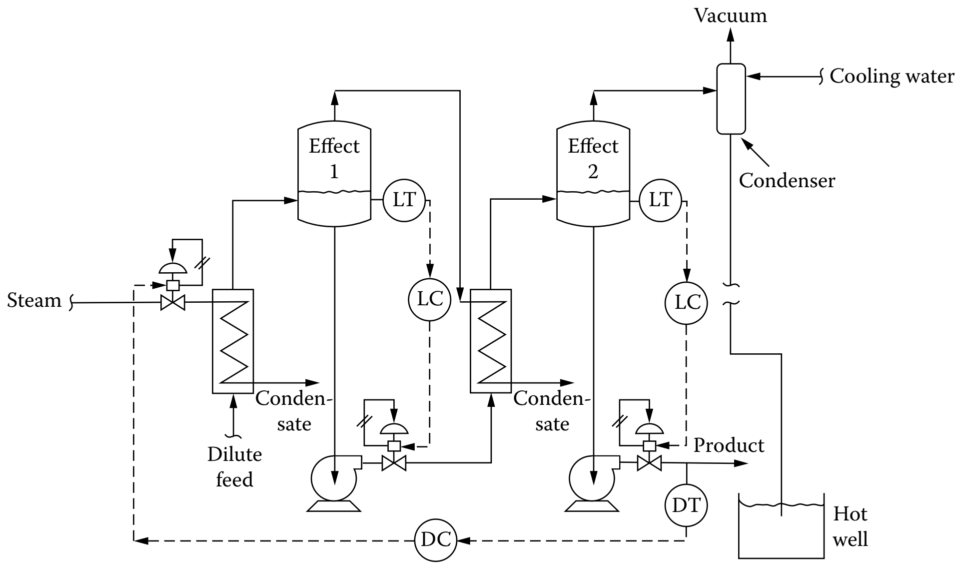


FIG. 8.23m
In a simple feedback control system the steam flow is directly throttled to keep the density of the product constant.

density measurement can be found in Chapter 6 of Volume 1 of this handbook, *Process Measurement and Analysis*.)

Ritter and associates[1] have modeled the control system configuration shown in Figure 8.23m and have found it to be very stable. They investigated other combinations of controlled and manipulated variables, which were less effective.

Cascade Control A typical cascade control system is illustrated in Figure 8.23n. This control system, like the feedback loop in Figure 8.23m, measures the product density and adjusts the heat input. The adjustment in this instance, however, is through a flow loop that is being set in cascade from the final density controller, an arrangement that is particularly effective when steam flow variations (outside of the evaporator) are frequent. It should be noted that with this arrangement the valve positioner is not required and can actually degrade the performance of the flow control loop. (For more information on cascade control see Section 2.6 in Chapter 2.)

Selective Control The selective control scheme shown in Figure 8.23o uses the previously described cascade configuration and adds to it the feature of protection against running

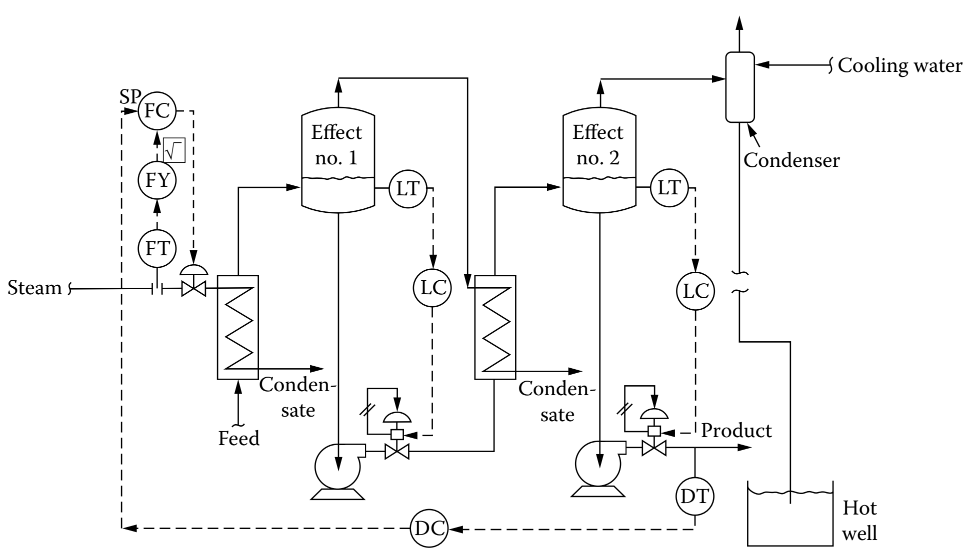


FIG. 8.23n
In the cascade version of a feedback control system the steam flow is controlled and the flow controller set point is adjusted to keep the density of the product constant.

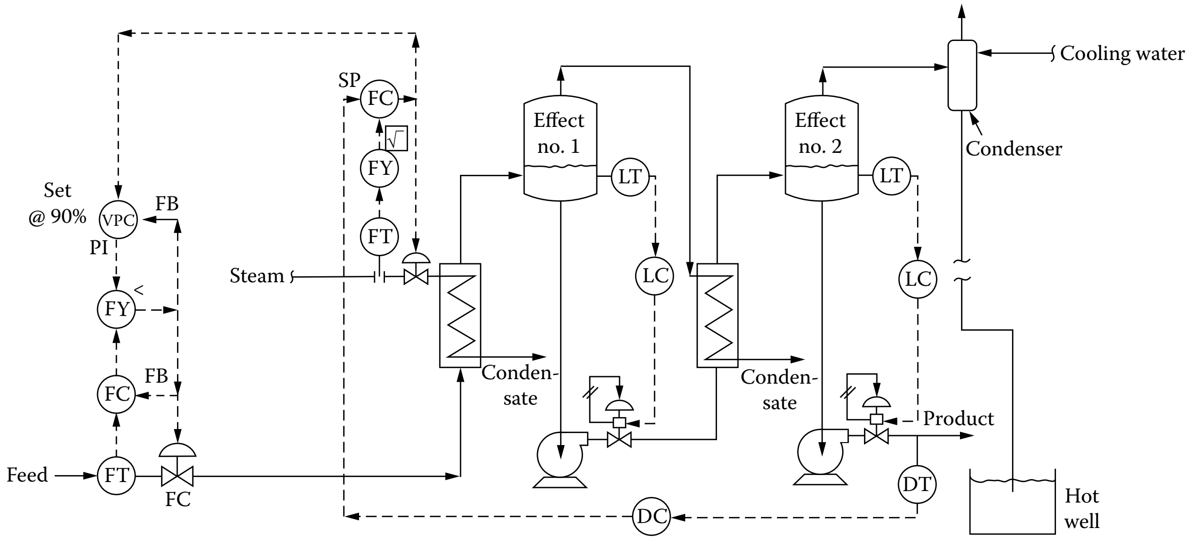


FIG. 8.23o
Valve position controller (VPC) cuts back the feed flow and thereby protects the system from running out of steam. The selective control scheme is protected from reset windup by external feedbacks.

out of steam. This feature is provided by the valve position controller (VPC), whose measurement is the steam valve opening and whose set point is about 90%. Therefore, as the feed gets too diluted and therefore requires more steam to concentrate, the steam valve opens. When it reaches 90% lift (which on an equal-percentage valve corresponds to 70% flow), the VPC output signal drops below that of the feed flow controller (FC), and therefore the low signal selector (FY) transfers the control of the feed valve to the VPC. During the period while the feed is diluted due to some upset, the VPC sets the feed flow rate to a value that corresponds to the allowable maximum opening of the steam valve.

The VPC is a PI controller with the integral mode dominating (as is the case in most valve position controllers). This allows it to operate smoothly even if the measurement signal (the steam valve opening) is noisy, as its output signal responds not so much to the error, but to the total area under the past error curve. Both controllers (FC and VPC) are provided with external feedback taken from the output of the low selector FY. This way the controller that is in control will have its output signal and its feedback signal at the same values and therefore will operate as a normal PI controller.

The controller that is not selected will operate as a proportional-only controller with a bias, where the bias is the external feedback signal. This guarantees that at the time of switchover the two outputs will be identical, and therefore the switchover will occur bumplessly (at the time of switchover the error in the idle controller has just reached zero, and therefore its output equals its external feedback signal).

Feedforward Control

In most evaporator applications the control of product density is constantly affected by variations in feed rate and feed density to the evaporator. In order to counter these load variations, the manipulated variable (steam flow) must attain a new operating level. In the pure feedback or cascade arrangements this new level was achieved by trial and error as performed by the feedback (final density) controller.

A control system able to react to these load variations when they occur (feed rate and feed density) rather than wait for them to pass through the process before initiating a corrective action would be ideal. This technique is termed *feedforward control*. (For an in-depth discussion of feedforward control, refer to Section 2.9 in Chapter 2.)

Figure 8.23p illustrates in block diagram form the features of a feedforward system. Feedforward controls utilize

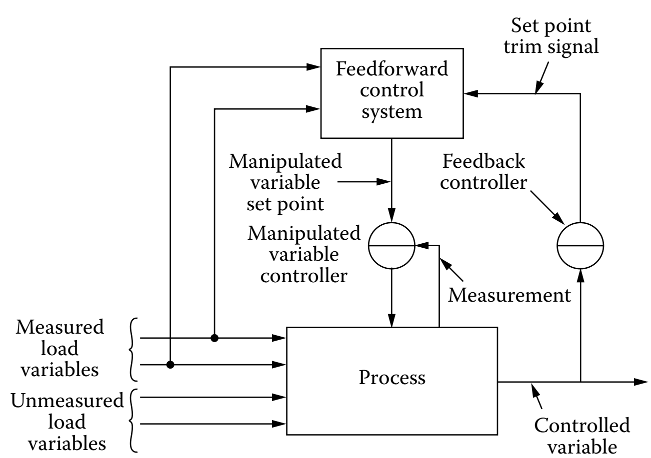


FIG. 8.23p
Feedforward system.

a simplified model of the controlled process. There are two types or classes of load variations—measured and unmeasured. The measured load signals are inputs to the feedforward control system, where they serve to compute the set point of the manipulated variable control loop as a function of the measured load variables.

The unmeasured load variables pass through the process, undetected by the feedforward system (disregarded in this simplified model), and cause an upset in the controlled variable. The output of the feedback loop then trims the calculated value of the set point to the correct operating level. In the limit, feedforward control would be capable of perfect control if all load variables could be defined, measured, and incorporated in the forward set point computation. This is what multivariable model predictive control provides, as will be discussed later.

At a practical level, then, the load variables are classified as either major or minor, and the effort is directed at developing a feedforward model that incorporates the major load variables, the manipulated variables, and the controlled variable. Such a relationship is termed the steady-state model of the process. Minor load variables are usually very slow to materialize and are hard to measure.

In terms of evaporators, minor load variations might be heat losses and tube fouling. Load variables such as these are usually easily handled by a feedback loop. The purpose of the feedback loop is to trim the forward calculation to compensate for the minor or unmeasured load variations. Without this feature the controlled variable would go off set point.

In addition to the two ingredients of a feedforward control system that have already been discussed, the steady-state model and feedback trim, there is a third ingredient. This third ingredient of a successful feedforward system application is dynamic compensation.

Dynamic compensation is required when a change in one of the major process loads also requires a change in the operating level of the manipulated variable. If the load and the manipulated variables enter the process at different locations, there usually will be an imbalance or inequality between the effects of the load and the manipulated variables on the controlled variable; i.e.,

$$\frac{\Delta \text{ controlled variable}}{\Delta \text{ load variable}} \neq \frac{\Delta \text{ controlled variable}}{\Delta \text{ manipulated variable}} \qquad \mathbf{8.23(29)}$$

This imbalance manifests itself as a *transient* excursion of the controlled variable from set point. If the forward calculation is accurate, the controlled variable returns to set point once the new steady-state operating level is reached.

In terms of a co-current flow evaporator, an increase in feed rate will call for an increase in steam flow. Assuming that the level controls on each effect are properly tuned, the increased feed rate will rapidly appear at the end of the train, while the effect of the increased steam flow will take longer, because it has to overcome the thermal inertia of the process. Therefore, an increase in feed flow results in a transient decrease of the controlled variable (density), because the load variable passes through the process faster than the manipulated variable. This behavior is shown in Figure 8.23q.

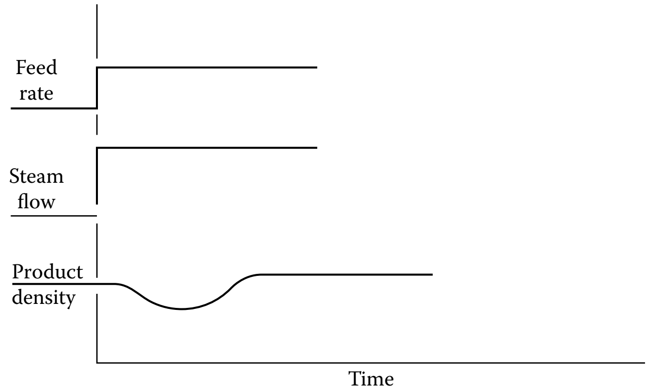


FIG. 8.23q
Load variable faster than manipulated variable.

The same sequence is seen in Figure 8.23r, except that this figure illustrates the case when the manipulated variable passes through the process faster than the load variable. Such behavior may occur in a countercurrent evaporator operation. This dynamic imbalance is normally corrected by inserting a dynamic element (lag, lead-lag, or a combination thereof) in at least one of the load measurements to the feedforward control system. Usually, dynamic compensation of that major load variable that can change in the severest manner (a step change) is all that is required. For evaporators this is usually the feed flow rate to the evaporator.

Feed density changes, although frequent, are usually more gradual, and the inclusion of a dynamic element for this variable is usually not warranted. In summary, the three ingredients of a well-designed feedforward system are (1) an accurate steady-state model, (2) properly set dynamic compensation, and (3) a correctly designed feedback trim, which corrects for the inaccuracy of the steady-state model.

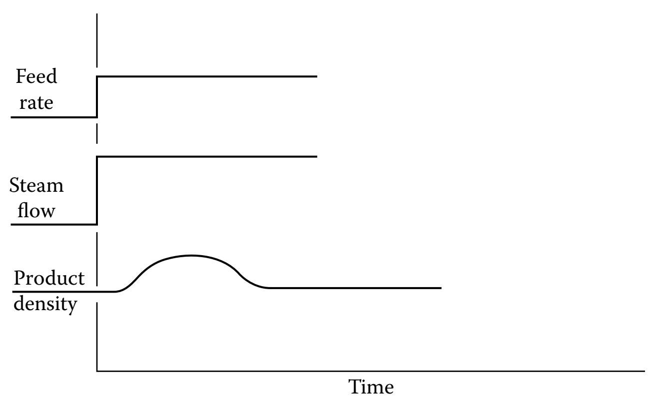


FIG. 8.23r
Manipulated variable faster than load variable.

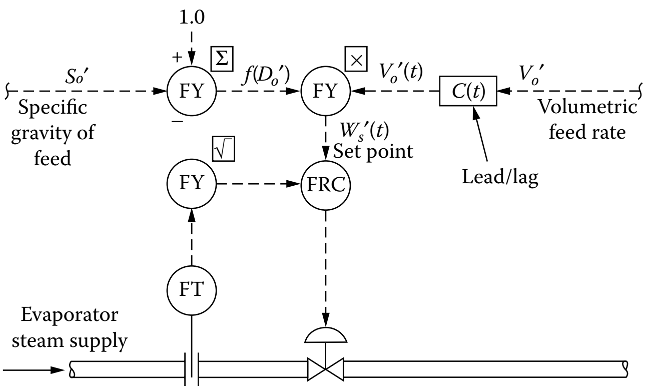


FIG. 8.23s
Implementation of the feedforward control loop for the case when the manipulated variable is steam flow, which is described in Equation 8.23(27), with first-order lag added for dynamic compensation.

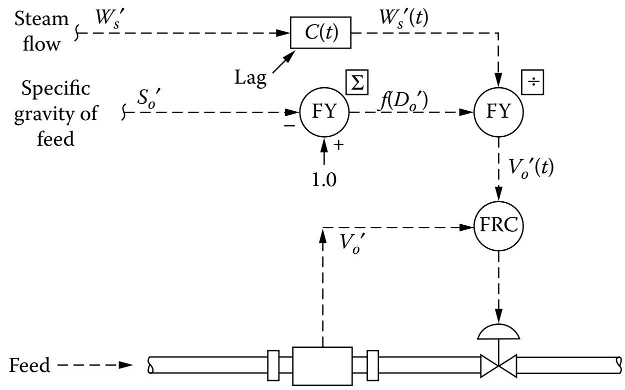


FIG. 8.23t
Implementation of the feedforward control loop for the case when the manipulated variable is feed flow, which is described in Equation 8.23(28), with first-order lag added for dynamic compensation.

Dynamic Compensation The dynamics of the co-current evaporator, in which steam is the manipulated variable, requires that a lead-lag dynamic element be incorporated in the system to compensate for the dynamic imbalance between feed rate and steam flow.

In the example, it was arbitrarily assumed that steam flow is the manipulated variable resulting in Equation 8.23(17). In some applications evaporators are run on waste steam, in which case the feed rate is proportionally adjusted to the available steam (Figure 8.23o), which makes feed the manipulated variable and steam the load variable.

Solving Equation 8.23(17) for feed rate,

$$V_o' = \frac{W_s E}{D_w(D)} \qquad \textbf{8.23(30)}$$

In this arrangement the dynamics do not change, but the manipulated variable advances through the process faster than the load variable, which requires a dynamic element having first-order lag characteristics. The instrument arrangement for the case where the manipulated variable is the steam supply was shown in Figure 8.23s.

The configuration where the manipulated variable is the feed flow is shown in Figure 8.23t and is based on Equation 8.23(28).

Feedback Trim of the Feedforward Loop As a general rule, feedback trim is incorporated into the control system *at the point at which the set point of the controlled variable appears*. For the evaporator the set point is the slope of the $f(D_o)$ relationship (Figure 8.23g). If the weight fraction of solids in the product (X_p) changes, the slope of the line changes too.

To this point the slope of the line (value of m) was assumed to be a constant (0.437), which value is incorporated into the summing relay or amplifier in Figures 8.23s and 8.23t. The instrumentation was scaled to make one grade of product, e.g., 50% solids. If a more or less concentrated product were desired, the gain term would have to be changed manually. In order to increase the flexibility of the control system, in Figure 8.23u a multiplier and a final product density control loop are added.

The controller output is now variable not only to permit changing the concentration of the product (slope adjust) but also to adjust the steam flow set point to compensate for the minor load variations, which up to this point were not considered.

The feedback trim for the case when steam is the manipulated variable is shown in Figure 8.23u. For the configuration where the manipulated variable is the feed flow, the product density controller will trim the value of the slope m, but in most other respects the feedforward-feedback configuration

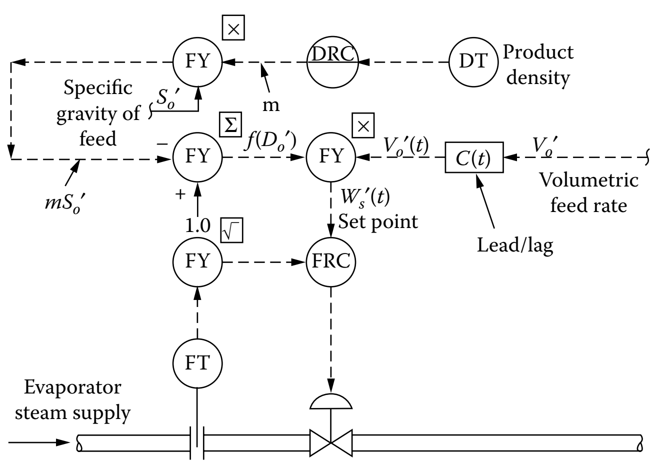


FIG. 8.23u
Feedforward control loop with feedback trimming of the slope m is shown for the case where steam is the manipulated variable.

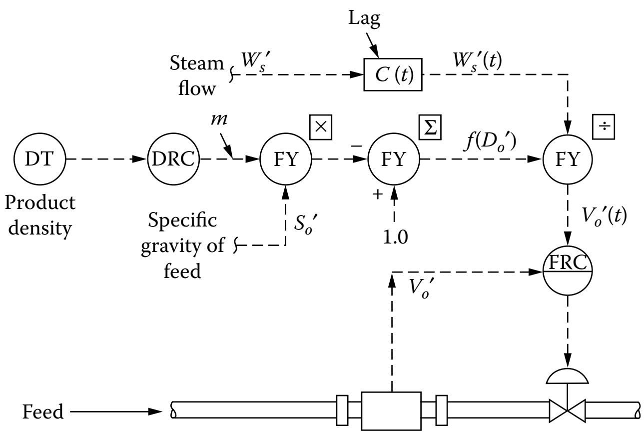


FIG. 8.23v
When feed flow is the manipulated variable, the feedforward signal based on steam flow is lagged while the product density controller trims the slope m in a feedback manner.

will be shown as in Figure 8.23v. The one exception is that the lead-lag dynamics is replaced with a simple lag.

Other Load Variables

Until this point it was assumed that the only load variables in an evaporation process are the feed flow rate and density. This is not the case, and in the next paragraphs, some of the other load variables and the controls they require will be considered.

Steam Enthalpy So far we have discussed only load changes related to feed density and feed flow rate and configured control loops to control the evaporator in response to these variations. One additional load variable that, if allowed to pass through the process, would upset the controlled variable (product density) is steam enthalpy.

In some applications the steam supply may be carefully controlled so that its energy content is uniform. In other applications substantial variations in steam enthalpy may occur. In order to correct for this, one must consider the factors that influence the energy content of the steam and design a control system that will protect the process from the changes in this load variable.

For saturated steam the energy content per unit weight is a function of the absolute pressure of the steam. If the flow of steam to the process is measured with an orifice meter, the mass flow of the steam is

$$W_s = K_1\sqrt{\frac{h}{v}} \qquad \textbf{8.23(31)}$$

where

W_s = steam flow in lb/hr (kg/hr)
h = differential head measurement in ft (m)
v = specific volume in cu. ft per lbm (m^3/kg)
K_1 = orifice coefficient dependent on the physical characteristics of the orifice

Therefore, the total energy to the system is

$$Q = W_s H_s \qquad \textbf{8.23(32)}$$

where

Q = energy to system per BTU per hour (J/hr)
W_s = steam flow in lb/hr (kg/hr)
H_s = heat of condensation in BTU per lb, or J/kg (enthalpy of saturated vapor minus enthalpy of saturated liquid)

Substituting Equation 8.23(31) into Equation 8.23(32) gives

$$Q = K_1\sqrt{\frac{(H_s)^2}{v}h} \qquad \textbf{8.23(33)}$$

For any particular application the steam pressure will vary around a normal operating pressure. To demonstrate the design of a control system to compensate for variations in the energy input into the process, assume that the steam pressure varies between 18 and 22 PSIA (124 and 152 kPa), with a normal operating value of 19.7 PSIA (135.9 kPa). The pressure transmitter has a range of 0–25 PSIA (0–172.5 kPa). These values of pressure variation and operating pressure are typical for a number of evaporator operations.

The value of the $(H_s)^2/v$ term appearing in Equation 8.23(33) will vary, depending on the steam pressure. Over a reasonably narrow range of pressures, the value of $(H_s)^2/v$ can be approximated by a straight line with the general form of

$$(H_s)^2/v = bP + a \qquad \textbf{8.23(34)}$$

where

P = absolute pressure in PSIA (Pa)
a and b = constants

Table 8.23w shows the typical values of H_s, v, $[(H_s)^2/v]'$, and P and P' selected for the specified range of pressures and for a case in which the steam is condensed at 1.5 PSIG.

$$\left(H_s^2/v\right) \text{ normal} = \frac{(970.7)^2}{20.4} = 46,208$$

Rewriting Equation 8.23(34) in scaled form:

$$\left(\frac{(H_s)^2}{v}\right)' = bP' + a \qquad \textbf{8.23(35)}$$

The designer can either linearize the data—using any two points from Table 8.23w—or can use a least squares computation to find the best straight line.

TABLE 8.23w
Specific Volume-Enthalpy Data

(P) Steam Supply Pressure PSIA (kPa)	(H_s) Heat of Condensation BTU/lbm† (MJ/ka)	(v) Specific Volume ft^3/lbm (m^3/kg)	$((H_s)^2/\upsilon) = \frac{(H_s)^2/\upsilon}{((H_s)^2/\upsilon)_{normal}}$	$P' = \frac{P}{P_{max}} = \frac{P}{25}$
18 (124.2)	969.1 (225.4)	22.2 (1.38)	0.916	0.720
19 (131.1)	970.2 (225.7)	22.1 (1.37)	0.965	0.760
19.7* (135.9)	970.7 (225.8)	20.4 (1.26)	1.00	0.788
20 (138)	971.2 (225.9)	20.1 (1.25)	1.02	0.800
21 (144.9)	972.1 (226.1)	19.2 (1.19)	1.07	0.840
22 (151.8)	973.0 (226.3)	18.4 (1.14)	1.11	0.880

* Normal operation.
† Assuming that the steam is condensed at 1.5 PSIG.

In a least square computation the following values of *a* and *b* constants of Equation 8.23(35) were obtained:

$$\left(\frac{(H_s)^2}{v}\right)' = bP' + a \qquad \textbf{8.23(36)}$$

where
$a = 0.085$
$b = 1.161$

Squaring Equation 8.23(33) and substituting in Equation 8.23(36):

$$(Q^2)' = (1.161\ P' + 0.085)hk_1 \qquad \textbf{8.23(37)}$$

The parenthetical portion of Equation 8.23(37) can be rewritten so as to make the sum of the two coefficients equal 1.0, which simplifies its implementation using conventional analog hardware. This is done by multiplying and dividing each term in the parentheses by the sum of the two coefficients.

$$(Q^2)' = 1.246\left(\frac{1.161P'}{1.246} + \frac{0.085}{1.246}\right)hk_1 \qquad \textbf{8.23(38)}$$

$$(Q^2)' = 1.246(0.932P' + 0.068)hk_1 \qquad \textbf{8.23(39)}$$

The instrumentation to implement Equation 8.23(39) is shown in Figure 8.23x.

Internal Material Balance The feedforward system described earlier imposes an external material balance as well as an internal material balance on the process. The internal balance is maintained by liquid level control on the discharge of each effect.

Analysis of the performance of level loops indicates that a narrow proportional band (<10%) can achieve stable control. However, because of the resonant nature of the level loop it can cause the process to oscillate at its natural frequency. Therefore, in most installations a much lower controller gain must be used (proportional bands 50 to 100%).[5]

Because of such wider proportional bands, the addition of the integral mode is required to help maintain the set point. A valve positioner and booster relay are also recommended to overcome the usual limit cycle characteristics of an integrating process and the nonlinear nature of valve hysteresis (Figure 8.23y).

Absolute Pressure The heat to evaporate water from the feed material is directly related to the boiling pressure of the material. In most multiple-effect evaporation each effect is held at a pressure less than atmospheric in order to keep boiling points below 212°F (100°C). The lowest pressure is in the effect closest to the condenser, with pressures increasing slightly in each effect away from the condenser.

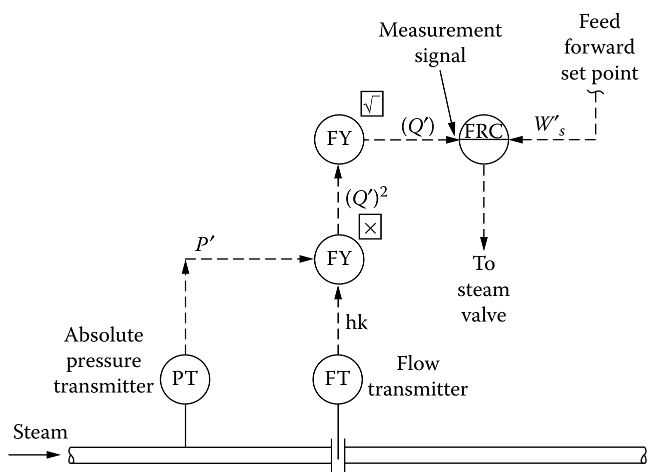


FIG. 8.23x
This control configuration compensates for variations in steam enthalpy by determining the heat input in terms of BTU/hr, based on the calculation required by Equation 8.23(39).

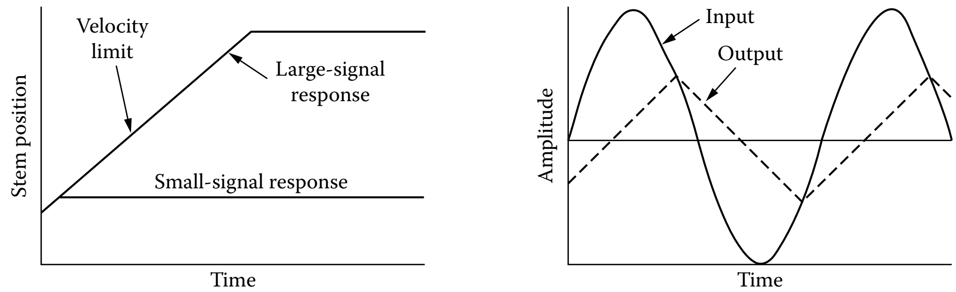


FIG. 8.23y
If a valve actuator cannot move as fast as the speed at which the control signal changes (velocity limited), a step change in the control signal (left) will result in a delayed straight-line response, while a sine wave in the control signal (right) will result in cycling.

The three possible methods of controlling the absolute pressure are (1) controlling the flow of water to the condenser, (2) bleeding air into the system with the water valve wide open, and (3) locating the control valve in the vapor draw-off line, manually setting the water flow rate and air bleeding as necessary.

Method 3 requires an extraordinarily large valve, because the vapor line may be 24 or 30 in. (600 or 750 mm) in diameter. Method 2 is uneconomical, because the expense of pumping the water offsets the savings realized by using a smaller valve on the air line. Method 1 represents the best compromise between cost and controllability, and therefore it is preferred (Figure 8.23z). (For more on condenser pressure control see Section 8.29.)

Auto-Select Controls

In many processes the final product is the result of a two-step operation. The first step produces an intermediate product that serves as the feed to a final concentrator. The aim is to ensure that the process is run at the maximum throughput consistent with the process limitations, an example of which is shown in Figure 8.23aa.

In this two-step evaporation process, three limitations are considered: (1) the steam availability to the intermediate concentrator can be reduced as a result of demands in other parts of the plant; (2) the steam availability to the final concentrator can be reduced as a result of demands in other parts of the plant; and (3) the final concentrator can accept feed only at or below a certain rate, and it is desired to run this part of the process at the rate set by its limits.

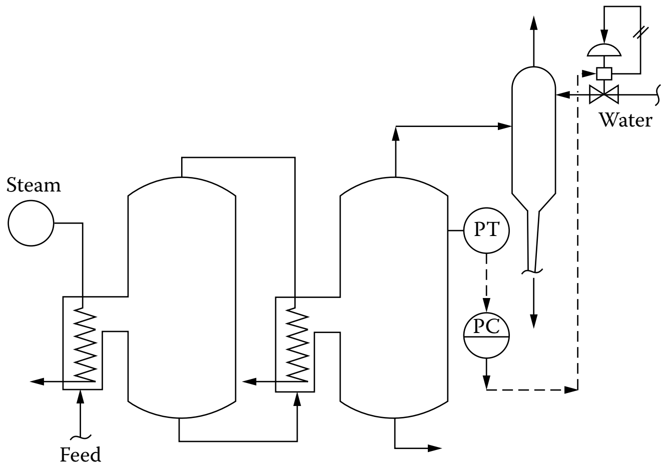


FIG. 8.23z
The recommended method of controlling the absolute pressure of evaporation is to throttle the water flow to the condenser.

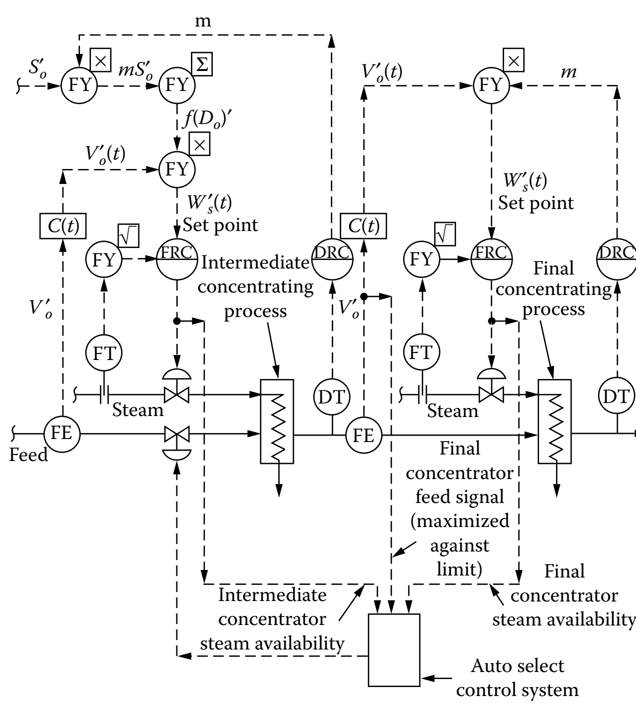


FIG. 8.23aa
This control system includes auto-select limits that consider the maximum feed rate to the final concentrator and the availability of steam in the plant.

Each part of the process has its own feedforward-feedback system. The intermediate concentrating process has the feedforward system shown in Figure 8.23u. The final concentrating process does not require feed density compensation, so that it is necessary only to establish the ratio of steam to feed and to adjust the ratio by feedback. The feed rate to the final concentrator is sent to the auto-select system that manipulates the feed to the intermediate process so as to maintain the final concentrator feed maximized at its set point limit.

Should either steam supply become deficient, the associated FRC output is selected to adjust the feed to the intermediate process. Thus, the process is always run at the maximum permissible rate consistent with the process limitation(s). (For more on auto-select control systems, see the explanation given in connection with Figure 8.23o.)

Trimming Controls

The system illustrated in Figure 8.23bb combines a number of advanced control features. These features include the selective control loop that was described in Figure 8.23o. This loop utilizes the valve position controller (VPC-4) to overrule the feed flow controller (FC-1) if the feed rate exceeds steam availability and requires the steam valve to open beyond 90%.

The feed flow rate is set by a nonlinear level controller (LC-12) on the feed tank. This tank provides surge capacity between the upstream equipment and the evaporator system. LC-12 is nonlinear, meaning that its output signal, which is the set point of FC-1, remains constant as long as the level in the tank is between 40 and 60%. This allows the evaporator loading to be stable, while production surges are averaged out in the feed tank. If the level gets above 60% or below 40%, the feed flow set point is slowly changed to match the new production rate.

The steam flow is measured on the basis of its heat content (enthalpy) as was explained in connection with Figure 8.23x. The set point of the steam flow controller (FRC-2) is obtained on a feedforward basis from the feed flow (FT-1).

Co-current evaporators usually respond faster to changes in feed flow than to steam flow. Therefore, the dynamic compensator in the feedforward loop (FY-11) is a dominant lead in Figure 8.23bb, just as it was in Figures 8.23s and 8.23u. If feed flow was manipulated in response to steam flow variations, the loop would require a dominant lag. (Countercurrent evaporators might respond faster to steam than to feed flow changes and, therefore, require different dynamic compensators.) A lead-lag compensator (FY-11) is usually provided with all feedforward systems and is adjusted as needed in the field.

The feedback trim in Figure 8.23bb is based on product density (DRC-6) and is configured similarly to the loop in Figure 8.23u. The steam flow set point is obtained in accordance with Equation 8.23(18), where E is the economy of the evaporator (the mass of vapor produced per unit mass of steam used).

If the product quality response to the changes in steam flow to the first effect (FRC-2) is too slow, the design engineer might speed up that response by adding a small trim heater to the last effect. Product quality responds faster to this trim heater, but it is more costly to add steam at this point, because the steam is used in only one effect, instead of passing through all.

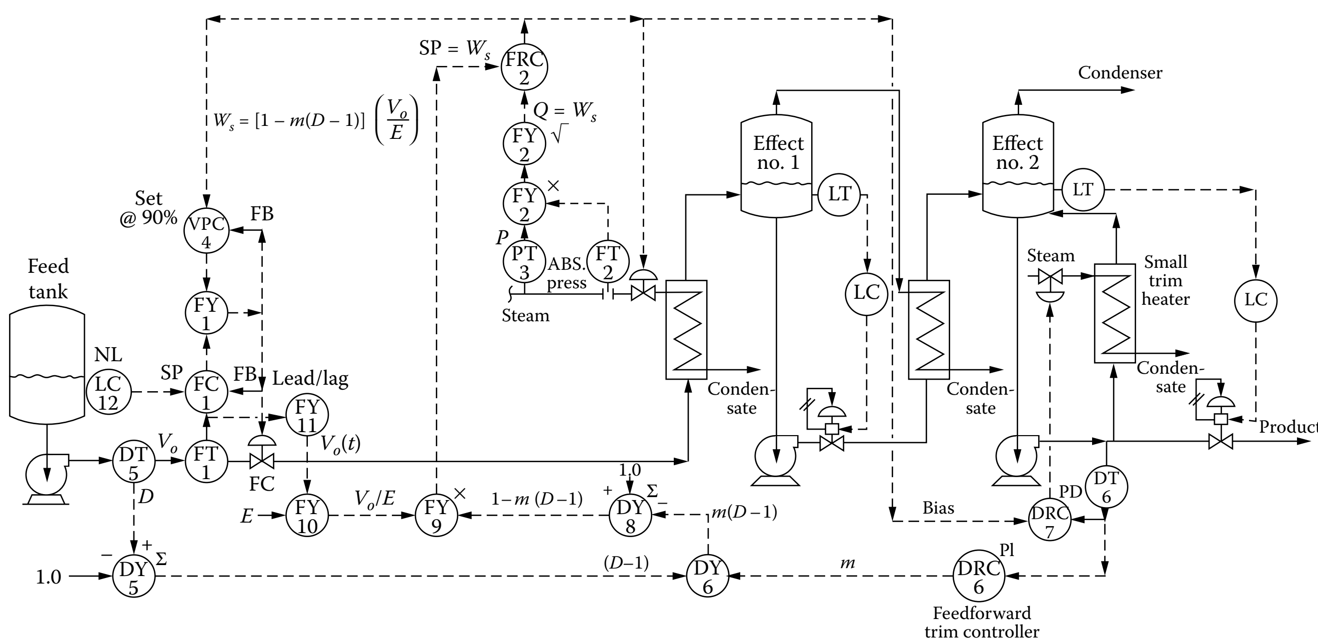


FIG. 8.23bb
Combined control system showing the discussed advanced control features plus controls of a trim heater on the last effect.

When a trim heater is used, as shown in Figure 8.23bb, the product density control is split between a proportional and integral controller (DRC-6), which controls the main steam flow, and a faster, proportional and derivative controller (DRC-7), which throttles the steam to the trim heater.

With feedforward control systems like the one shown in Figure 8.23bb, the control response is fast enough without a trim heater. Yet, if its better response can be economically justified, its controls should be coordinated with the primary steam controls. For this reason DRC-7 is biased by the main steam valve opening, so that they both move in the same direction.

Similar to the use of trim heaters, it is also possible to introduce a small stream of feed directly into the last effect. The use of this technique also improves the response of the system to variations in product quality but is also inefficient, as that stream of feed is exposed only to that one effect instead of to all. Trim controls were used prior to the introduction of feedforward controls, and in many cases the use of trim controls has been discontinued after the installation of feedforward controls.

PRODUCT DENSITY MEASUREMENT

Perhaps one of the most controversial issues in any evaporator control scheme is the method used to measure the product density. Common methods include (1) temperature difference, boiling-point rise; (2) conductivity; (3) differential pressure; (4) gamma gauge; (5) U-tube densitometer; (6) buoyancy float; (7) refractive index; and (8) oscillating Coriolis (see Chapter 6 in the first volume of this handbook).

Each method has its strengths and weaknesses (Table 8.23cc). In all cases, however, care must be taken to select a representative measurement location to eliminate

TABLE 8.23cc
Orientation Table for Density Sensors

	Applicable to					*Design Pressure and Temperature Limitations*				
LIQUID Density Sensor Design	*Clean Process Streams*	*Slurry Service*	*Viscous or Polymer Streams*	*Minimum Span Based on Water SG = 1.0*	*Inaccuracy in % of Spain or SG Units*	*PSIG/°F*	*Bars/°C*	*Temperature Compensation Available*	*Direct Local Indicator*	*Transmitter*
Angular Position Type	✓			0.1	0.5%	1000/500	69/260	N.S.		✓
Ball Type	✓			Digital	0.01 SG	600/160	41/71	N.S.	✓	✓
Capacitance Type	✓	✓	✓	0.1	1%	500/160	34.5/71	✓		✓
Displacement Type Buoyant Force Displacer	✓			0.005	1%	1500/850	130/472	N.S.		✓
Chain Balance Float	✓			0.005	1–3%	500/450	34/232	✓	✓	✓
Electromagnetic Suspension	✓			0.01	0.5–1%	200/350	14/177	✓		✓
Hydrometers	✓			0.05	1%	100/200	7/93	✓	✓	✓
Hydrostatic Head Type	✓	✓	✓	0.05	0.2–1%	5000/350	345/177	N.S.	✓	✓
Oscillating Coriolis	✓	✓	✓	0.1	0.02 SG or better	5000/800	345/426	✓		✓
Radiation Type	✓	✓	✓	0.05	1%	Unlimited	Unlimited	✓		✓
Sonic/Ultrasonic	✓	✓	✓	0.2	1–5%	1000/390	69/199	✓		✓
Twin Tube	✓	✓		Digital	0.0001	1440/356	100/180	✓	✓	✓
Vibrating Fork Type	✓	L	L	0.02	0.001 SG	3000/392	207/200	✓		✓
Vibrating Plate Type (also for gases) (currently not manufactured)	✓	L	L	0.1	0.2%	1440/200	100/95			✓
Vibrating Spool Type (also for gases)	✓	L	L	0.3	0.001 SG	725/300	50/149	✓		✓
Vibrating U.Tube Type	✓			0.05	0.00005–0.005 SG	2900/500	200/260	✓		✓
Weight of Fixed Volume Type	✓	✓	✓	0.05	1%	2400/500	165/260	✓	✓	✓

N.S.: Nonstandard
L: Limited

entrained air bubbles or excessive vibration, and the instrument must be mounted in an accessible location for cleaning and calibration. The relative location of the product density transmitter with respect to the final effect should also be considered. Long runs of process piping for transporting the product from the last effect to the density transmitter increase dead time, which in turn reduces the effectiveness of the control loop.

Boiling-Point Rise

Perhaps the most difficult and controversial method of product density measurement is by temperature difference or boiling-point rise. Dühring's rule states that a linear relationship exists between the boiling point of a solution and the boiling point of pure water at the same pressure. Thus, the temperature difference between the boiling point of the solution in an evaporator and the boiling point of water at the same pressure is a direct measurement of the concentration of the solution. Two problems in making this measurement are location of the temperature bulbs and controls of absolute pressure.

The temperature bulbs must be located so that the measured values are truly representative of the actual conditions. Ideally, the bulb measuring liquor temperature should be just at the surface of the boiling liquid. This location can change, unfortunately, if the operator decides to use more or less liquor in a particular effect. Many operators install the liquor bulb near the bottom of the pan, where it will always be covered, thus creating an error due to *head effects*, which must be compensated for in the calibration.

The vapor temperature bulb is installed in a condensing chamber in the vapor line. Hot condensate flashes over the bulb at an equilibrium temperature dictated by the pressure in the system. This temperature minus the liquid boiling temperature (compensated for head effects) is the temperature difference reflecting product concentration.

Changes in absolute pressure of the system alter not only the boiling point of the liquor but also the flashing temperature of the condensate in the condensing chamber. Unfortunately, the latter effect occurs much more rapidly than the former, resulting in transient errors in the system that may take a long time to resolve. Therefore, it is imperative that absolute pressure be controlled closely if temperature difference is to be a successful measure of product density. These systems are more effective when control of water rate to the condenser rather than an air-bleed system is used.

Conductivity

Electrolytic conductivity is a convenient measurement to use in relationships between specific conductance and product quality (concentration), such as in a caustic evaporator. For the conductivity of some aqueous liquids, refer to Figure 8.23dd. Problem areas include location of the conductivity cell so that product is not stagnant but is flowing past the electrodes; temperature limitations on the cell; cell plugging; and temperature compensation for variations in product temperature.

Differential Pressure

Measuring density by differential pressure is a frequently used technique. The flanged, extended diaphragm differential pressure transmitters are preferred for direct connection to the process (Figure 8.23ee); otherwise, lead lines to the transmitter could become plugged by process material solidifying in the lines. Differential pressure transmitters are more

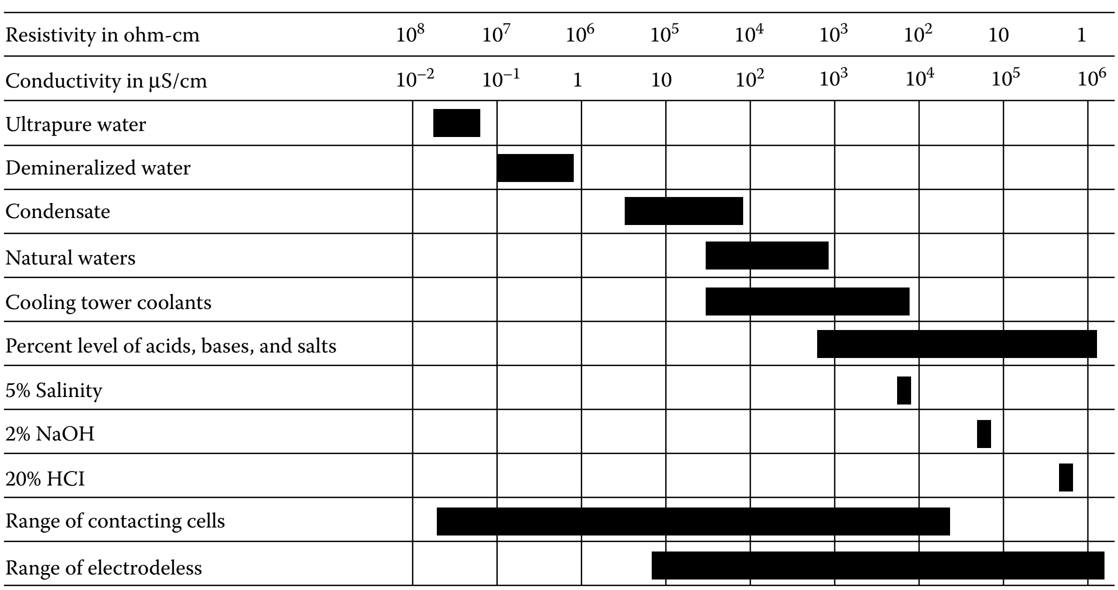


FIG. 8.23dd
Resistivity/conductivity spectrum of aqueous electrolytes. (From Light, T. S., Chemtech, August 1990, pp. 4960–4501.)

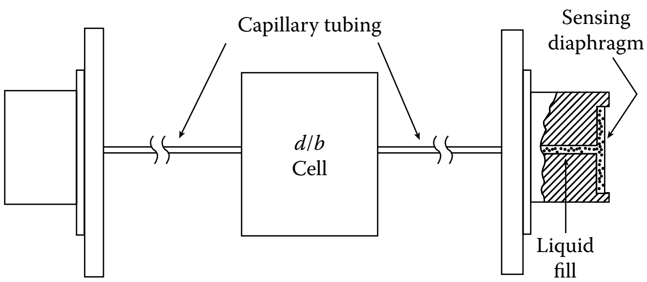


FIG. 8.23ee
d/p cell with extended chemical seal elements.

frequently used on feed density than on final product density measurements.

Gamma Gauge

This measurement is popular in the food industry because the measuring and sensing elements are not in contact with the process. It is very sensitive and not subject to plugging. Periodic calibration may be required because of decay due to the half-life of the source materials. Occasionally, air is entrained, especially in extremely viscous solutions. Therefore, the best sensor location is a flooded low point in the process piping.

U-Tube Densitometer

The U-tube densitometer, a beam-balance device, is also a final product density sensor. Solids can settle out in the measuring tube, causing calibration shifts or plugging.

Buoyancy Float

Primarily used for feed density detection, the buoyancy float can also be applied to product density if a suitable mounting location near the evaporator can be found. Because flow will affect the measurement, the float must be located where the fluid is almost stagnant or where flow can be controlled (by recycle) and its effects zeroed out. A Teflon-coated float helps reduce drag effects.

Oscillating Coriolis

When a Coriolis mass flowmeter (see Section 2.12 in Chapter 2 in the first volume of this handbook) is used for the measurement of feed flow, that same instrument can also provide a density reading for the feed. The error in these measurements is typically between 0.2 and 0.5% SG (0.002 to 0.005 g/cc).

CONCLUSIONS

The process of evaporation is well understood and easily modeled on the basis of mass and energy balance equations. For this reason, *white box*-type (see Section 2.10 in Chapter 2) multivariable models can be used to implement their advanced process control[3] (APC). When using APC-based unit operations controllers, the evaporators can be optimized to reach the criteria of maximized production, minimized energy consumption, or other goals.

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