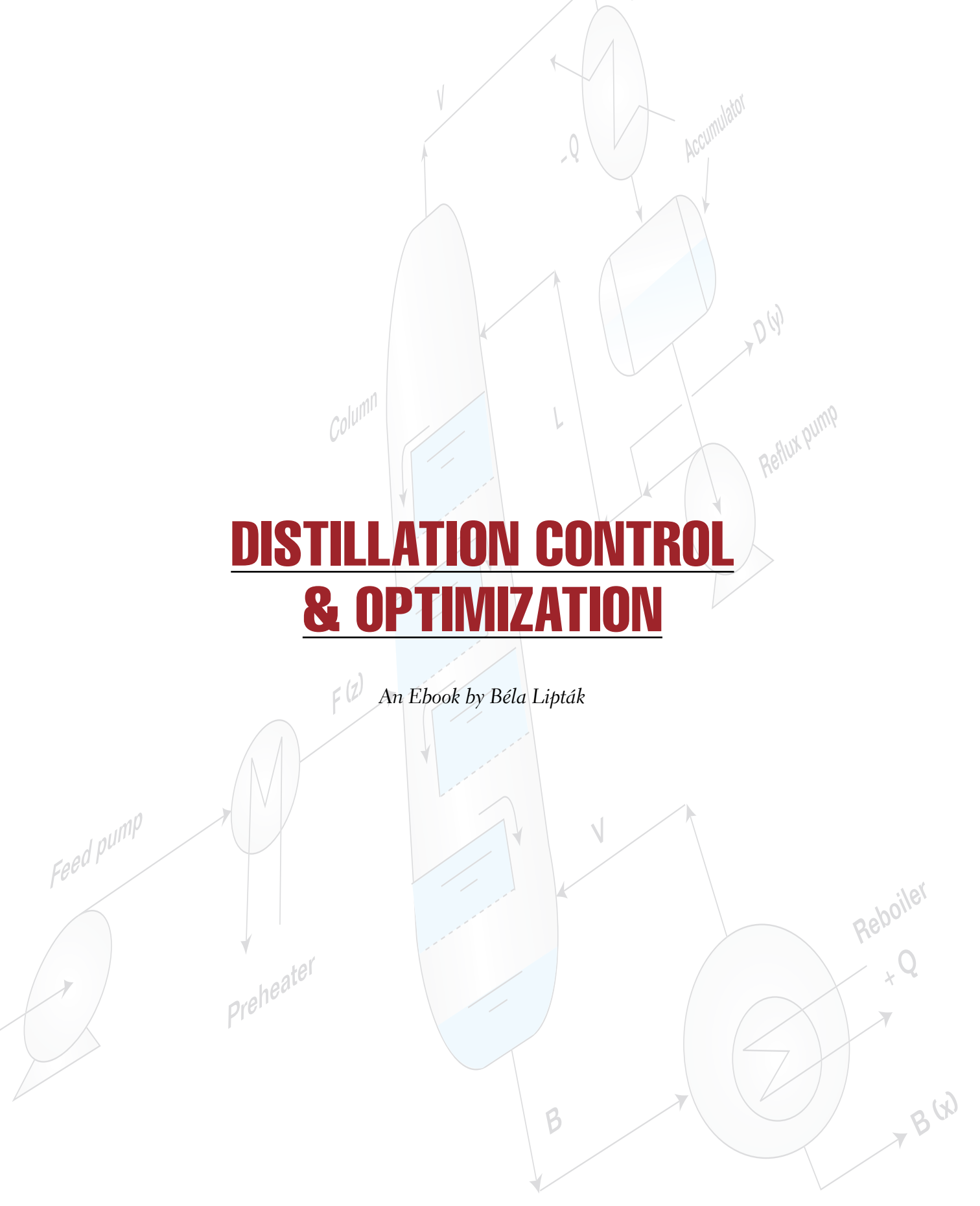


DISTILLATION CONTROL & OPTIMIZATION

An Ebook by Béla Lipták

Giants of the automation industry are few. Béla Lipták is one. From his beginnings in the Hungarian resistance movement to his volumes of work on automation and control, he has proven himself to be one of the great thinkers and authorities in the industry. In this ebook, *Distillation Control & Optimization*, he provides a concise, yet complete analysis of distillation control, showing "the potential for savings through better control and optimization."

In a practical, intuitive way, Lipták emphasizes reliable stable control strategies aimed at optimization of process performance. He claims that these optimization techniques can improve "productivity and profitability by 25%." I believe him. Distillation is arguably the most complex unit operation in most processing plants. Although volumes have been written on distillation control, most columns are not optimized. Instrument and process engineers will find a practical, readable treatment of the subject that can be applied quickly and effectively.

Like other publications from ControlGlobal.com, the *Distillation Control & Optimization* ebook is effective without being laborious. It provides focus on the important aspects of the subject for the time-pressed engineer who is responsible for distillation optimization. We are proud and honored to assist in bringing this important work to the automation community.

Martin Berutti

MYNAH Technologies

Distillation Control and Optimization – Part 1

Distillation Control and Optimization – Part 2

Distillation Control and Optimization – Part 3

Part 1

During the global transition from the present oil-based economy to one based on clean and inexhaustible energy, the importance of distillation will increase. Ethanol will need to be distilled, and new oil refineries will need to be built. It is hoped that the designers of these new refineries will benefit from this discussion, which describes the state of the art in the control and optimization of distillation processes.

Distillation is a common, energy-intensive method of separation in the petroleum, chemical, food, pulp and paper and pharmaceutical industries. Globally, more than 80 million barrels of crude oil are refined daily. In the U.S., 146 refineries operate, employing over 65,000 people and producing a total value that exceeds $151 billion. But no new refinery has been built in the U.S. since 1976, and the technology in use today is not much different from that used on the first distillation columns in the 19th century.

The thermal energy requirements of distillation are enormous. The thermodynamic efficiency of distillation processes is less than 10%. The amount of energy used for distillation is approximately 8% of the total energy used in the industrial sector of the U.S. Refineries spend 50% to 60% of their operating costs (excluding capital costs and depreciation) on energy, while the chemical industry spends only 30% to 40%. This difference shows the saving potential of implementing better control and optimization of the distillation process. *My main reason for writing this ebook is to show how these increases in efficiency and savings can be accomplished by the application of state-of-the-art process controls.*

I will first describe the distillation process, its components and dynamics. Next I will discuss the traditional PID-based control system configurations, which are still used in the majority of the operating refineries. I will conclude with a review of state-of-the-art distillation optimization strategies, including multivariable, model-based and artificial neural network-based control systems, where the process variables are used only as constraints, and the production rate and efficiency are continuously optimized. I hope to show that advanced controls can cut the operating costs and internal energy consumption of refineries and other distillation processes by 25%.

THE PROCESS

Distillation is the most frequently used separation process. It separates the components of a mixture on the basis of their boiling points and on the difference in the compositions of the liquids and their vapors. The product purity of a distillation process is maintained by the manipulation of the material and energy balances. Difficulties in maintaining that purity arise because of dead times, nonlinearities and variable interactions.

Distillation can be performed either as a batch or a continuous operation. The main difference between the two is that in continuous distillation, the feed concentration is relatively constant, while in batch, the concentration of the light components drops and that of the heavy components rises as distillation progresses.

Another basic difference between distillation operations is in the handling of the heat removed by the condenser at the top of the column. The more common approach is to waste that heat by rejecting it into the cooling water. In this case, "pay heat" must be used at the bottom of the column in the reboiler. Figure 1 (p. 5) illustrates this configuration and identifies its main components.

Because a large part of the total operating cost is in providing the heat required at the reboiler in some distillation systems, the heat content of the bottom product is used to preheat the feed to the column.

FIGURE 1.

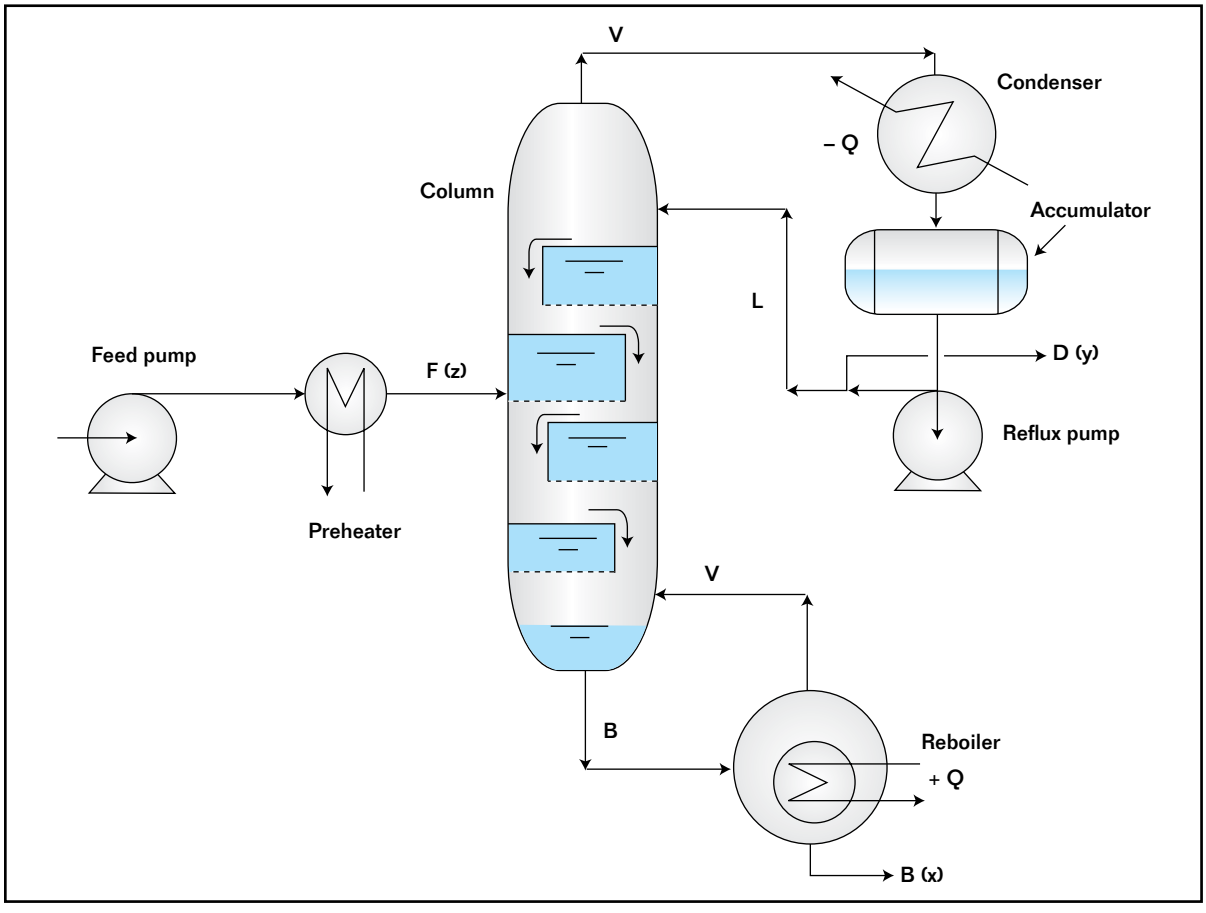


Illustration of a tray-type distillation tower, where (without accumulation), the material balance is F = D + B and D = V – L. The mole fractions of the light key component in the bottoms, distillate and feed are identified as x, y and z. For binary separation, S = (y[1-x])/(x[1-y]).

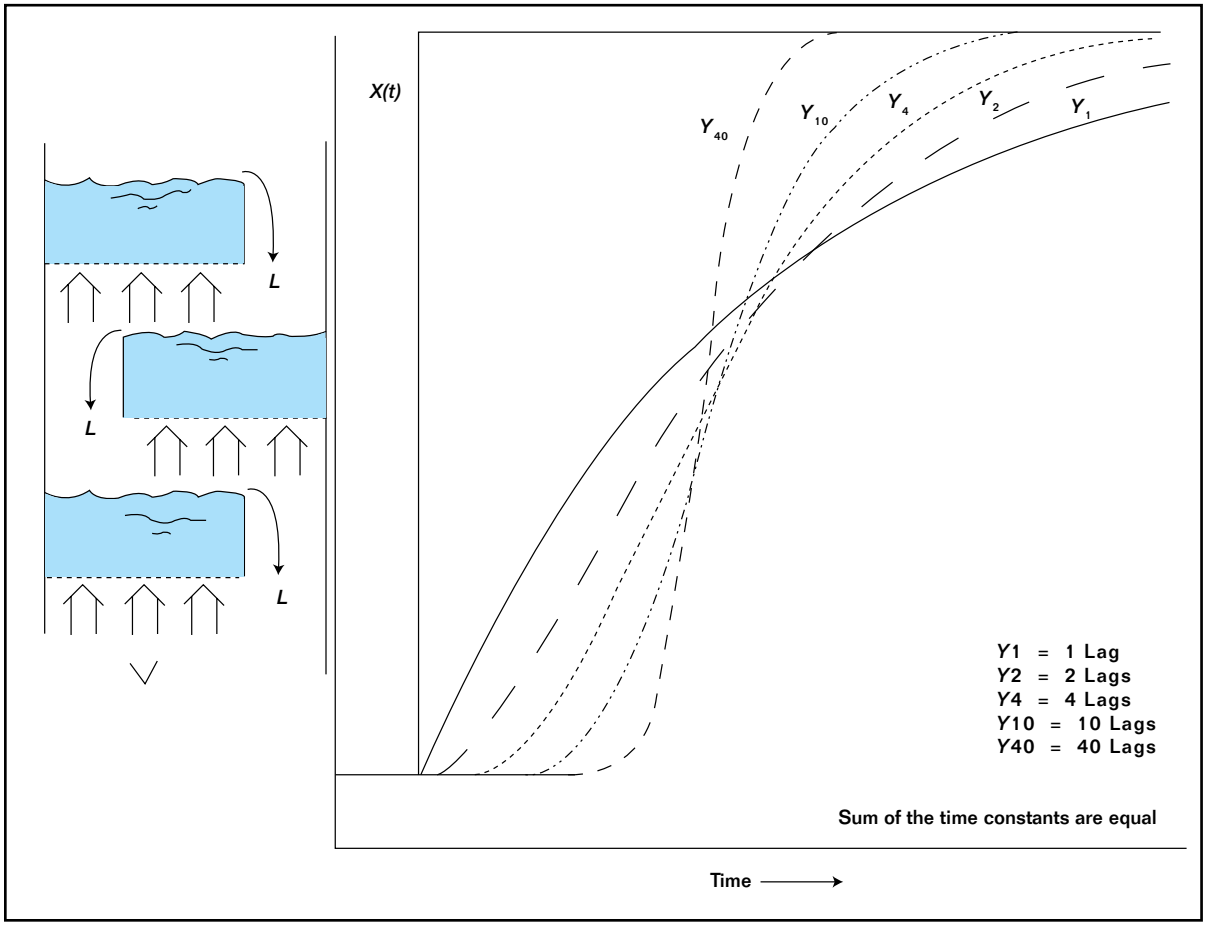
The other option is to recycle the heat removed at the condenser by a heat pump (compressor). In this configuration, as the vapors from the column (V) are condensed, the heat from the condenser is used to vaporize a working fluid. These vapors are at the low pressure of the suction side of the compressor (heat pump). When the working fluid vapors are compressed, and these high-pressure (and temperature) vapors in the reboiler contact the bottoms liquid from the column, they condense, and their heat of condensation serves to vaporize the liquid from the column bottoms (Figure 13, left, p. 21). While vapor recompression is energy-efficient, it is not used very frequently.

The Column

The main distillation equipment is the *column, tower or fractionator.* It has two purposes: First, it separates a feed into a vapor portion that ascends the column and a liquid portion that descends; second, it achieves intimate mixing between the two counter-current flowing phases. The purpose of the mixing is to get an effective transfer of the more volatile components into the ascending vapor and a corresponding transfer of the less volatile components into the descending liquid

The separation of phases is accomplished by the differences in vapor pressures, with the lighter vapor rising to the top of the column and the heavier liquid flowing to the bottom. The portion of the column above the feed is called the *rectifying* section and below the feed, the *stripping* section. The intimate mixing is obtained by either filling the column with lumps of an inert material (*packing*) or by the use a number of horizontal plates, or *trays*, which cause the ascending vapor to bubble through the descending liquid (Figure 2, left, p. 6).

FIGURE 2.



Left: The contact between liquid and vapor is made intimate as the vapors ascend through the liquids held on each tray as the liquid descends. The dynamics of a multiple-tray column can be approximated as a second-order lag, plus dead time. Right: The responses of the distillate composition (y) of 1st-, 2nd-, 4th-, 10th- and 40th-order processes are shown when a unit step change in bottoms composition (x) occurs. A 40-tray column is a 40th-order process.

Generally trays work better in applications requiring high flow, such as those encountered in high pressure distillation columns—depropanizers, debutanizers, xylene purification columns and the like. Packing works best at lower flow parameters because the low pressure drop of structured packing makes it very attractive for use in vacuum columns or ethylbenzene recycle columns of styrene plants.

The influence of plate efficiency in the operation of the distillation tower becomes important in the control of the overhead composition. Because plate efficiencies increase with increased vapor velocities, the influence of the reflux-to-feed ratio on overhead composition becomes a nonlinear relationship.

Column dynamics are a function of the number of trays, because the liquid on each tray must overflow its weir and work its way down the column; therefore, a change in composition will not be seen at the bottom of the tower until some time has passed. These lags are cumulative as the liquid passes each tray on its way down the column. Thus, a 30-tray column could be approximated by 30 first-order exponential lags in a series of approximately the same time constant. The effect of increasing the number of lags in series is to increase the apparent dead time and increase the response-curve slope. Thus, the liquid traffic within the distillation process is often approximated by a second-order lag, plus dead time (Figure 2, right).

Column Variables and Their Pairing

Controlled variables include product compositions, column temperatures and pressure, and tower and accumulator levels. Manipulated variables include reflux, coolant, heating medium and product flows. Load and disturbance variables include feed-flow rate, feed composition, steam-header pressure, feed enthalpy, environmental conditions (e.g., rain, barometric pressure and ambient temperature) and coolant temperature.

The general guidelines for pairing manipulated variables with controlled variables are as follows:

- Manipulate the stream that has the greatest influence on the controlled variable.
- Manipulate the smaller if two streams have the same effect on the controlled variable.
- Manipulate the stream that is more nearly linear with the controlled variable.
- Manipulate the stream that is least sensitive to ambient conditions.
- Manipulate the stream least likely to cause interaction.

In a binary distillation process, the number of independent variables is eleven, and the number of defining equations is two. Therefore, the number of degrees of freedom is nine. Consequently, the maximum theoretical number of automatic controllers that can be used on a binary distillation process is nine, but usually only five are controlled.

These variables are the compositions of the bottom and top products (x and y), the levels in the column base and accumulator, and the column pressure.

The manipulated variables that can be assigned to control these are the distillate (D), bottoms (B) and reflux (L) flows, the vapor boil-up (V set by heat input Q_B), heat removal (Q_T) and the ratios of L/D or V/B. These five single loops can theoretically be configured in 120 different combinations, and selecting the right one is a prerequisite to stability and efficiency.

Column pressure almost always is controlled by heat removal (Q_T). This loop closes the heat balance around the column, while the levels are controlled to close its material balance. Therefore, the key task is the assignment of the manipulated variables to the composition controllers. No matter how we make that selection, these two loops will interact. A change in one will upset the other because whenever the openings of their control valves change, the material and heat balance of the column will also change.

Therefore, the most important decision in designing the distillation controls is to assign the least-interacting manipulated variables to the composition control loops. The tool used in making that selection is the relative gain (RG) calculation.

Assigning Variables, Relative Gain Calculations

The RGs of the two composition loops are calculated as the ratio of their open-loop gains when the other loop is open (in manual,) divided by their open-loop gains when the other loop is closed (in automatic). The open-loop gain can be measured by manually changing the valve opening by, say, 1% and reading the resulting percentage change in the controlled variable when the new steady state is reached. The higher the ratio of controlled-to-manipulated variable response, the higher will be the open-loop gain and, therefore, when the loop is closed (placed in automatic), the lower the controller gain (wider proportional band) has to be to obtain stable control.

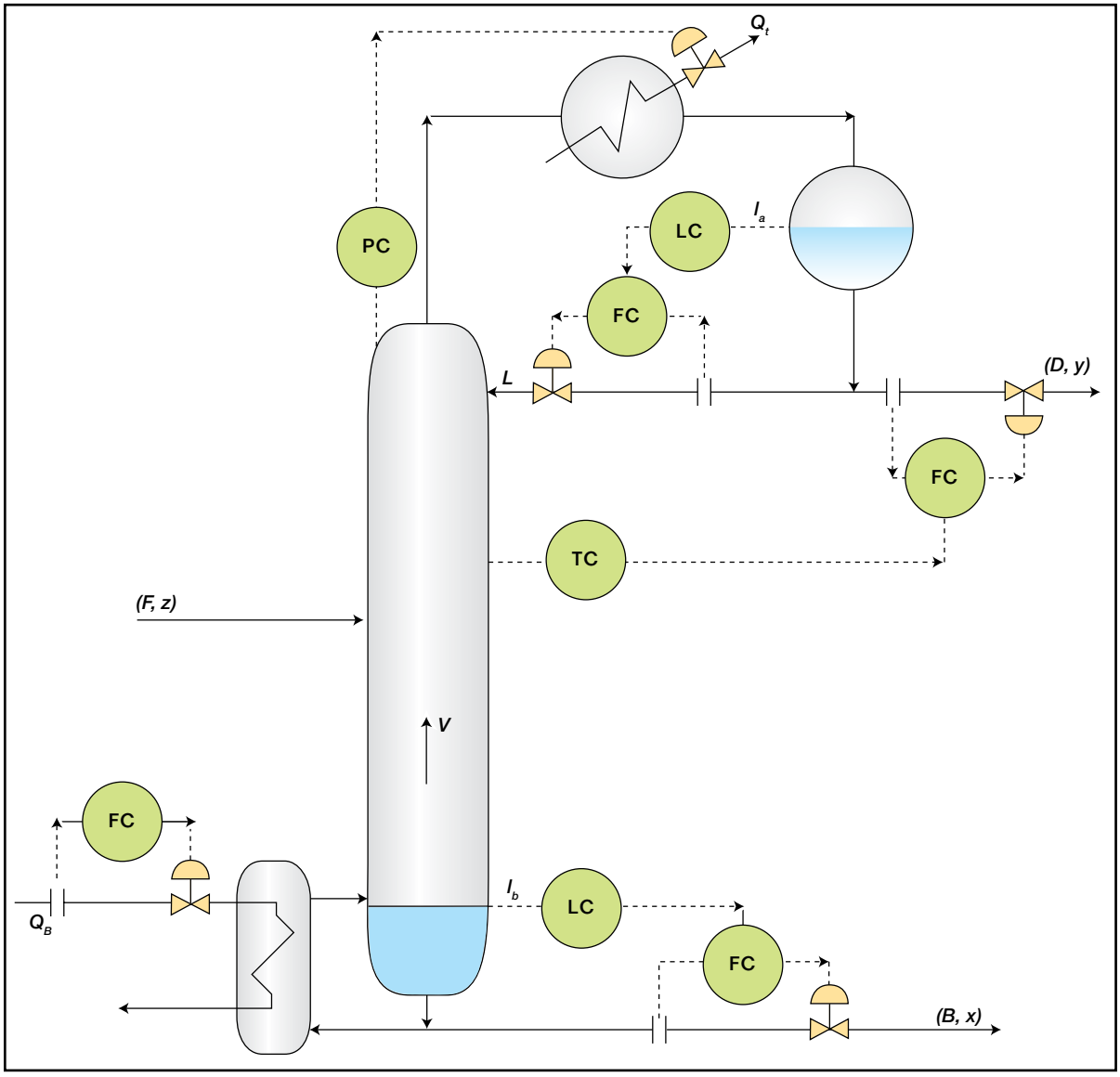
Naturally, the ideal RG is 1.0. $RG = 1$ indicates that the loop gain is unaffected (the tuning of the loop does not need to be changed) when the other loop is switched from manual to automatic or back. Consequently, our goal of selecting the combination of loops with the least interaction is to find pairings with RG values near unity (not much above or below $RG = 1$). For the various equations to be used in making RG calculations, see Chapter 2.25, Vol. 2, *The Instrument Engineer's Handbook*, 4th edition.

If RG is zero, that indicates that the manipulated variable has no direct influence, and the loop can only control through interaction with the other loop. Consequently, the process is controllable only when the

other loop is closed (in automatic). Similarly, a very high RG value (say, 100), indicates that the process is controllable only if the other loop is open (in manual). Negative RG values will reverse the controller action and should never be considered. If the RG is between 0 and 1, the feedback from the interaction is negative and when $RG > 1$, the feedback from the interaction is positive. Loop configurations with RG values that are < 0.5 or > 10 should be rejected, while RG values in the range of 0.5 to 10 are controllable. Naturally, the further the RG is from 1.0, the worse the interaction. About the same amount of interaction is indicated by an RG value of 0.5 and an RG value of 10.

Figure 3 illustrates a possible result of calculating the RG values. Here it was concluded that fixing the production rate (heat input to the column Q_B) and controlling only the distillate composition (y), while allowing the bottoms composition (x) to float, will give the most stable and efficient operation.

FIGURE 3.



In this example, the five manipulated variables are so assigned to the five controlled variables that the heat input at the reboiler (Q_B) and the distillate composition (y) are fixed and, therefore, the bottoms flow (B) and composition (x) are allowed to change with the variations in feed flow (F) or composition (z).

COMPOSITION CONTROL

Conceptually, product quality is determined by the heat balance of the column. The heat removal determines the internal reflux flow rate, while the heat addition determines the internal vapor rate. These internal vapor and liquid flow rates determine the circulation rate, which in turn determines the degree of separation between two key components.

The first task in configuring the control system for a distillation column is to configure the primary composition control loops. This configuration must consider the interaction between the proposed control loops, the column's operating objectives and the most likely disturbance variables. The measurements of the composition control loops can either be direct or inferred. Figure 4 provides some guidance on how to select the manipulated variables for controlling the compositions (and levels) of distillation columns.

FIGURE 4.

Dynamic Response and Sensitivity Limitations on the Paring of Distillation Control Variables

(If both compositions are important, they should not both be controlled by material balance [B, D])

Manipulated Variables / Controlled Variables	Distillate Flow (D)	Bottoms Product Flow (B)	Vaporization Rate (V) or Heat Input at Reboiler (Q)	Reflux Flow Rate (L)
Composition of Overhead Product (y)	OK if L/D ≥ 6 Note 3		Notes 1 and 2	Note 2
Composition of Bottoms Product (x)		Note 3	Notes 1 and 2	OK if trays ≤20
Accumulator Level	OK if L/D ≤ 6		Not good with furnace. OK if V/B ≥ 3	OK if L/D ≥ 0.5
Bottoms Level		OK if V/B ≤ 3	Not good if furnace is used. OK if diameter at bottom ≤ 20 ft.	

Notes:

1. Controls the concentration (x or y) which has the shorter residence time by throttling vapor flow (v).

2. More pure product should control separation (energy).

3. Less pure product should control material balance.

4. When controlling both x and y, the only choices for possible pairings are

a. Control y by D and x by V,

b. Control y by D and x by L,

c. Control y by L and x by V,

d. Control y by B and x by L.

Of these choices, d is not recommended because a y/B combination is not responsive dynamically.

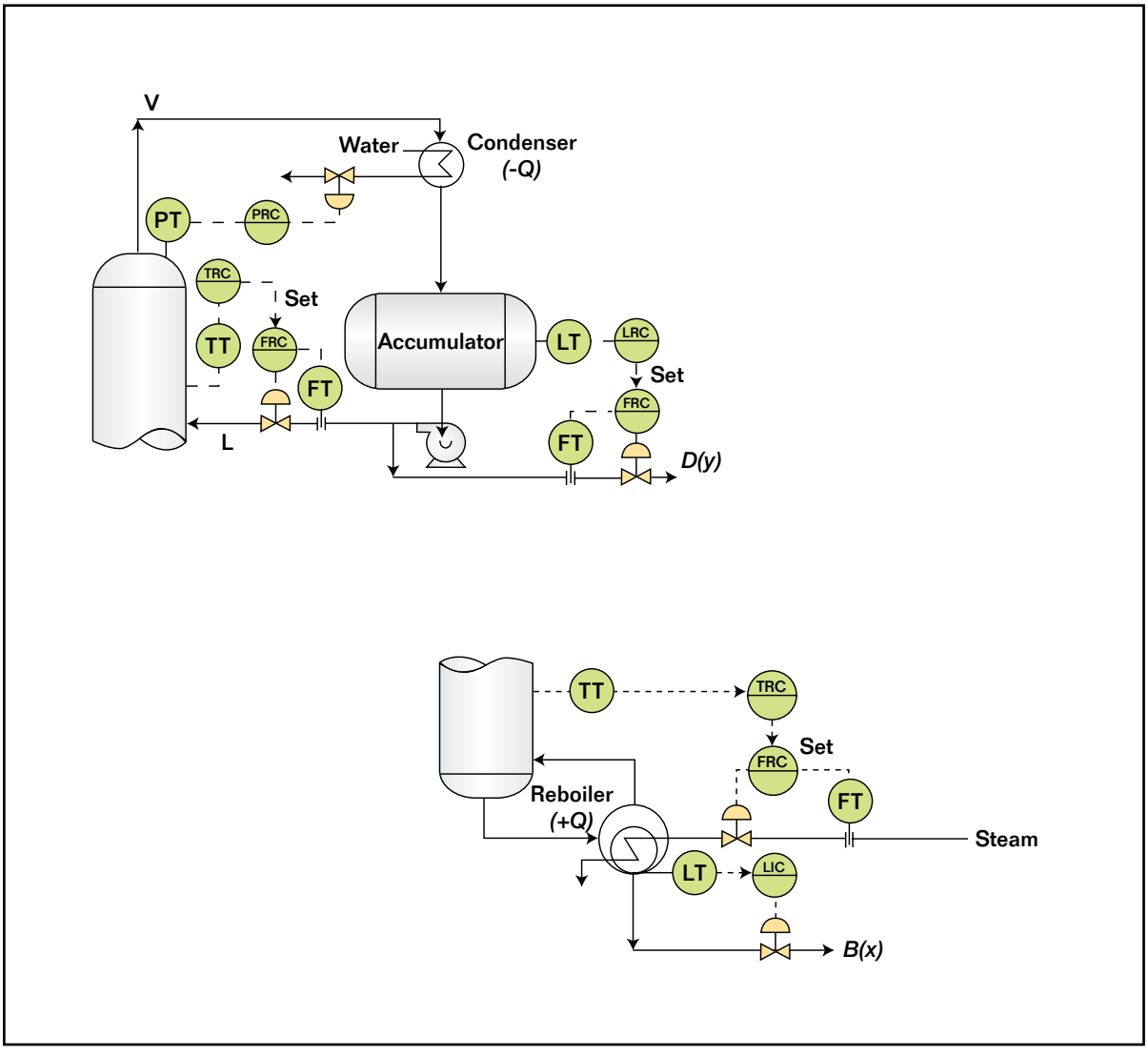
Indirect (Temperature-Based) Composition Control

If the feed composition and the column pressure are constant, temperature can be used as an indirect measure of composition. If the column pressure is not constant, the temperature measurement must be pressure-compensated. When the bottom product composition is controlled, the temperature sensor is located in the lower half of the column, and when overhead composition is controlled, in the upper half.

The temperature sensor should be located on a tray that strongly reflects changes in composition. This means a 1% change in the manipulated variable (reflux or steam flows in Figure 5, p. 10) should result in a temperature change of at least 0.1 °C to 0.5 °C, and that this change should be symmetrical, meaning that a 1% drop or rise in these flows should result in approximately the same size of drop or rise in the temperature on the control tray.

When two compounds of relatively close vapor pressures are to be separated (for example normal butane from isobutene), temperature measurement is not sensitive enough to measure composition. In such cases, two temperatures or a temperature difference (say, between Trays 5 and 15) can be used instead of a single sensor. This configuration can also be used to eliminate the effects of column pressure variations.

FIGURE 5.



Top: Distillate composition can be controlled by a cascade temperature master on the upper part of the column, which manipulates the reflux flow L. Bottom: Similarly, the bottoms composition can be controlled by a cascade temperature master located on the lower half of the column, throttling the reboiler heat input.

Composition Measurement

The direct measurement of composition is more expensive, but also more accurate and versatile than is indirect temperature measurement. Intermittent analyzers, such as chromatographs (with cycle times of a few minutes), are often provided with dead time compensation for closed-loop control. The analyzer update time must always be less than the response time of the process. For improved accuracy, one usually measures the impurity concentration in the controlled stream. This way the upsets caused by feed composition changes, tower pressure or efficiency variations can be more accurately corrected.

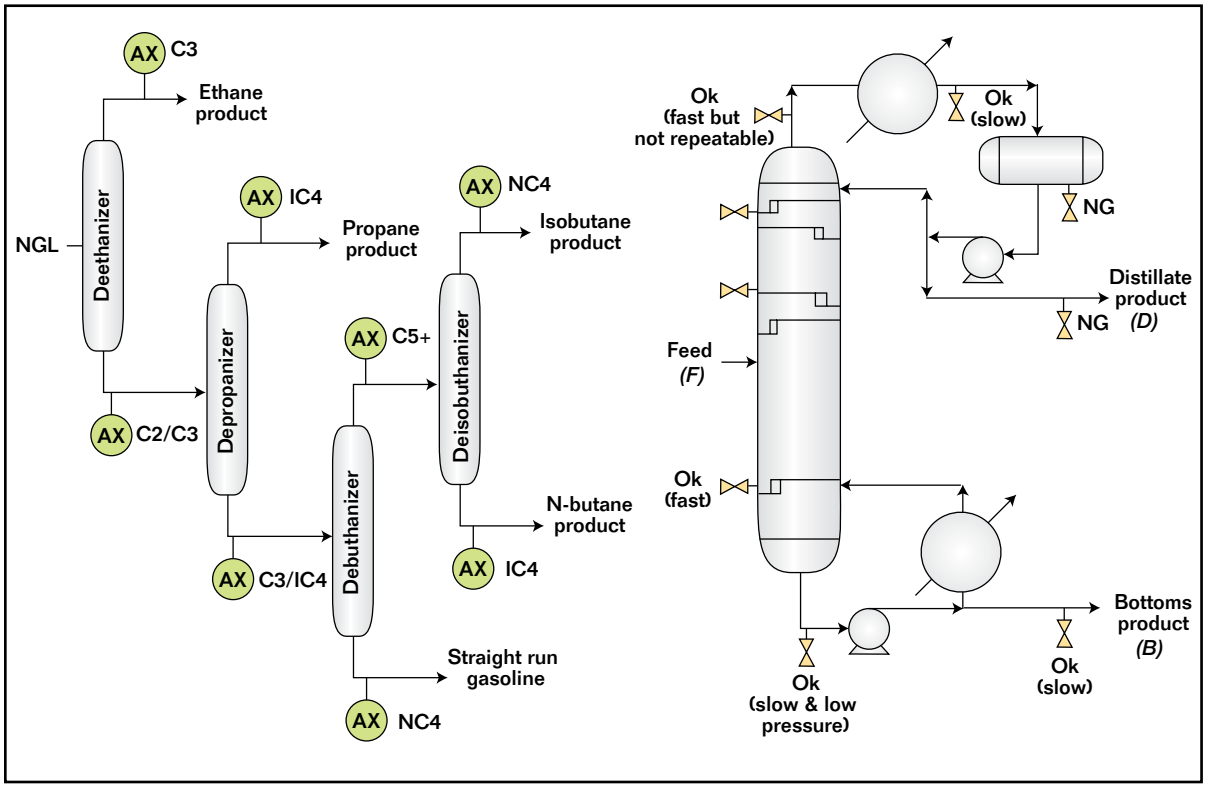
In the case of fractionator trains, the bottom or distillate analyzer on one column can also be the feed analyzer on the next one. In such processes, one can control either the concentration of a single component or the ratio of two components (Figure 6, left, p. 11). As will be shown later, such decisions can be based on profitability and market considerations.

In order for the composition to be close to the terminal composition, the point at which the measurement is made must also be near the column terminals. This will also minimize the transportation lag (dead time). In some applications though, the sampling point is moved closer to the column feed because that provides earlier recognition of feed composition changes or because the key control component is present at a higher concentration. Figure 6, right, describes some of the factors that should be considered in sample point selection. The samples should always be filtered, dried and cooled, and their dead times (transportation lags) must not exceed 30 seconds. Transportation lags can be reduced by the installation of high-flow bypasses.

Analyzer Selection

Most of the 66 types of analyzers that are discussed in Chapter 8, Vol. 1, *The Instrument Engineer's Handbook*, 4th edition, can also be used to control the distillation process. On the LNG train (Figure 6, left), in the depropanizer (where isobutane is to be measured in the presence of ethane, propane and normal butane) and in the deisobutanizer (where isobutane is to be measured in the presence of normal butane and isopentane), an infrared analyzer is recommended. In the debutanizer, where the combined isopentane plus normal pentane concen-

FIGURE 6.



Left: Recommended analyzer locations on fractionator trains. Right: Sampling point considerations on individual columns.

trations are measured in the presence of isobutane and normal butane to control the butane-pentane separation, gas chromatography is recommended.

Some physical properties analyzers, such as boiling-point analyzers, are reliable enough to be used for online control (see Chapter 8.48, Vol. 1, *The Instrument Engineer's Handbook*, 4th edition). Cut points between overhead products and side-cuts can be controlled on the basis of temperature, but doing so results in the downgrading of the more valuable product to the stream of lesser value. This downgrading can be minimized through the use of online boiling point analyzers. Justification of a boiling point analyzer depends upon the value of the products and the cost of analyzer maintenance.

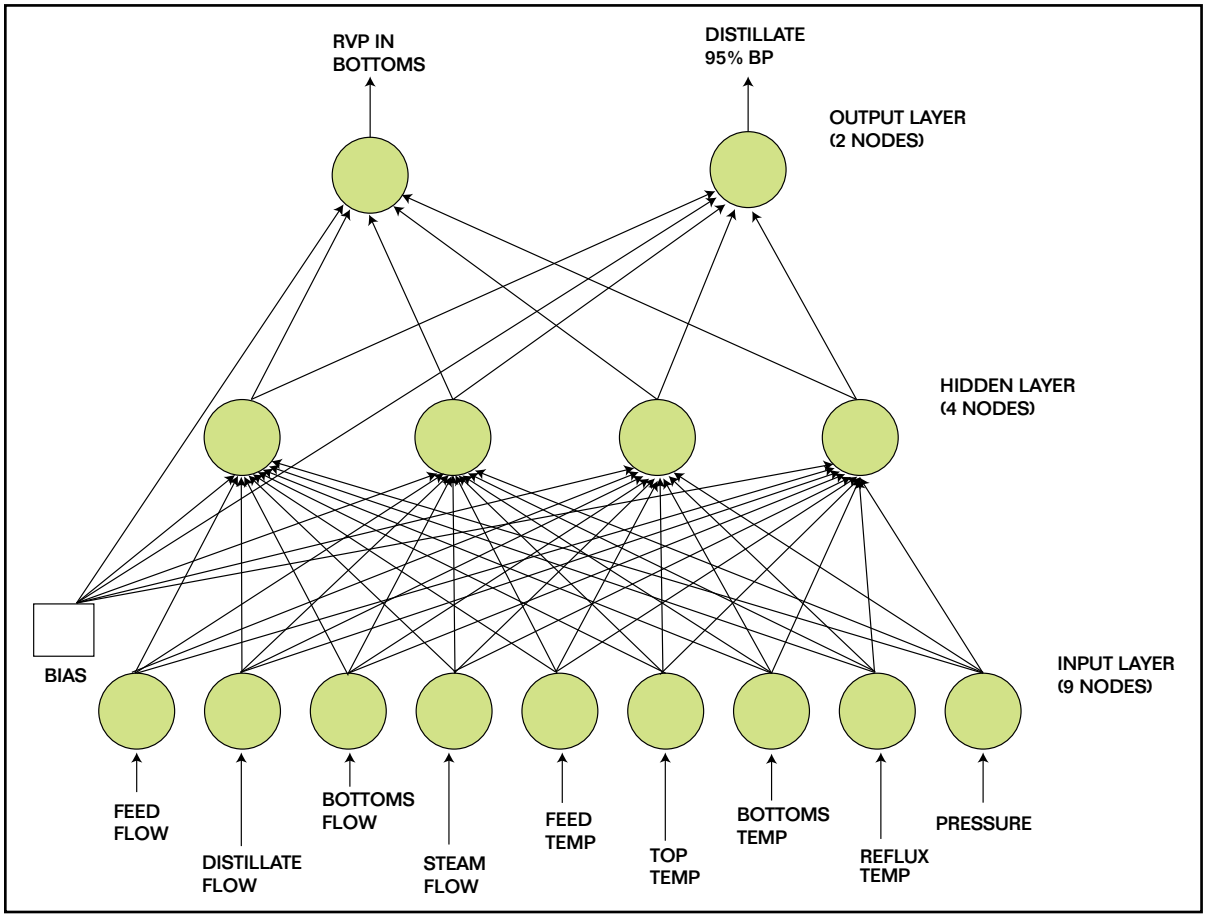
Viscosity can also be measured continuously to give faster control in vacuum distillation. The use of a viscosity analyzer minimizes downgrading during major upsets and large feed composition changes. With such an arrangement, low-viscosity vacuum bottoms can be detected quickly and diverted to recoverable feed for profitable reprocessing.

Many other analytical instruments are being moved out of the laboratory and into the processing area. Mobile units containing several different kinds of analyzers can be used to learn the best place to locate the on-stream analyzers. In cases in which permanent analyzers cannot be justified, the mobile unit is connected to the process long enough to find the best operating conditions. Then the mobile unit can be moved elsewhere.

Recently artificial neural networks (ANN) have also been used as indirect analyzers. In Figure 7, the product specifications are based on the Reid vapor pressure in the bottoms product and on the 95% boiling point of the distillate. Using ANN software eliminates the problems of dead time, cost and availability limitations of direct analyzers. ANN uses a nonlinear neural network model, which infers the analytical readings on the basis of other measurements. If historical data exists, the model can be developed based on collected laboratory or analyzer results, as commonly recorded on log sheets.

Figure 7 illustrates the case where the Reid vapor pressure of the bottoms product and the boiling point of the distillate are predicted by an ANN model. This particular model has nine measurements (input nodes), four hidden nodes and two output nodes with bias.

FIGURE 7.



ANN model-based software can provide indirect analytical reading predictions, overcoming dead time, availability and cost limitations of online analyzers.

Analyzer Sampling

The sample system must condition the sample to remove traces of foreign materials through filtering, maintain pressure and temperature, and maintain or change phase for introduction into the analyzer. The description of the wide variety of samplers, probes, filters and other sampling system components is beyond the scope of this work and is covered in detail in Chapter 8.2, Vol. 1, *The Instrument Engineer's Handbook*, 4th edition.

The system must transport the sample from the sample point to the analyzer with a minimum of transport lag (preferably less than 30 seconds and definitely not greater than one minute). Transportation times are minimized by using high-flow-rate bypass streams taken from the process sample point. With a circulation sample pump, care must be taken to prevent cavitation by locating the pump close to the sample take-off.

Single-line transport is the most direct approach and is used when the sample line volume is small in relation to the analyzer sample consumption, so that the transport time lag is reasonably short. (See the tabulation at the bottom of Figure 8, p. 14).

After selecting the appropriate sample transport method, a calculation of the sample time lag should be made, based on the following:

- Available differential pressures,
- Total length of the fast loop from the sample take-off point to the analyzer location and back to the sample return point,
- Line sizes,
- Viscosity of the sample.

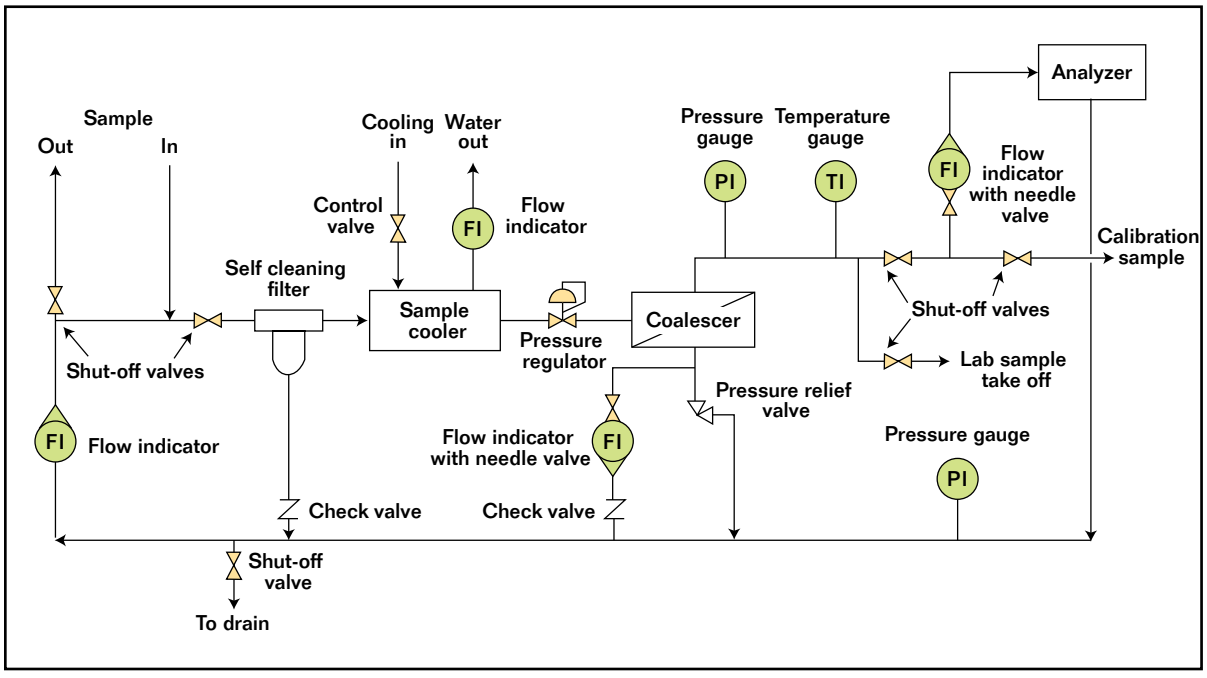
Sample disposal is a critical consideration, both from an economic and an ecological perspective. One goal is to prevent the emission of most hydrocarbons into the air. When there is an economic justification for saving the sample, as when dealing with liquids at the boiling point and viscometer analyzers, a sample collection and return system must be furnished to collect the sample at atmospheric pressures and pump it back at high pressure into the process. For gases with no sample return point, the sample can be pressurized back in to the process, or as is most frequently done, vented in to the flare system.

For chromatographs, liquid sample points are generally preferred (Figure 8) because of condensation at the sample probe and in the sample lines when hydrocarbons with high boiling points are present in the sample. When the sample lines are long, some separation between components can also occur. A satisfactory point for measuring bottoms product composition is at the point of highest pressure immediately after the product pump. However, if liquid holdup in the reboiler and kettle is large, a long lag is introduced, so an alternative sample point, such as a bottoms tray or seal pan may be used.

A satisfactory sample-point location for measuring the distillate is the outlet liquid of the overhead vapor condenser. Sampling the overhead accumulator liquid after the reflux or distillate pump should be avoided because of the tremendous process lag it introduces. Sampling the overhead vapor reduces the process lag of sampling after the condenser if a repeatable, representative sample can be obtained.

Figure 9, Part B, (p. 15) shows the same configuration as Figure 9, Part A, except that the analyzer controller is equipped with a first-order Smith predictor (discussed on page 15) which provides dead time compensation.

FIGURE 8.



Top: Refinery chromatograph liquid sampling system and its components. Bottom: Tubing or pipe volumes and dimensions.

Dimensions and Volumes of Tubing and Pipe Used in Sample Systems

Type	Nominal Volume pER DIAmeter, in.	Inner Diameter, in.	Internal Area	
			in²	cc
316 stainless steel tubing	1/8	.0787	.0048	.9571
	¼	0.1850	0.0268	5.3035
	3/8	0.0253	0.0684	13.4417
	½	0.4055	0.1290	25.2984
Schedule 40 pipe	¼	0.3642	0.1040	20.4521
	3/8	0.4921	0.1891	37.4904
	½	0.6220	0.3038	59.7408
	¾	0.8268	0.5863	106.6800

Composition Control Using Analyzers

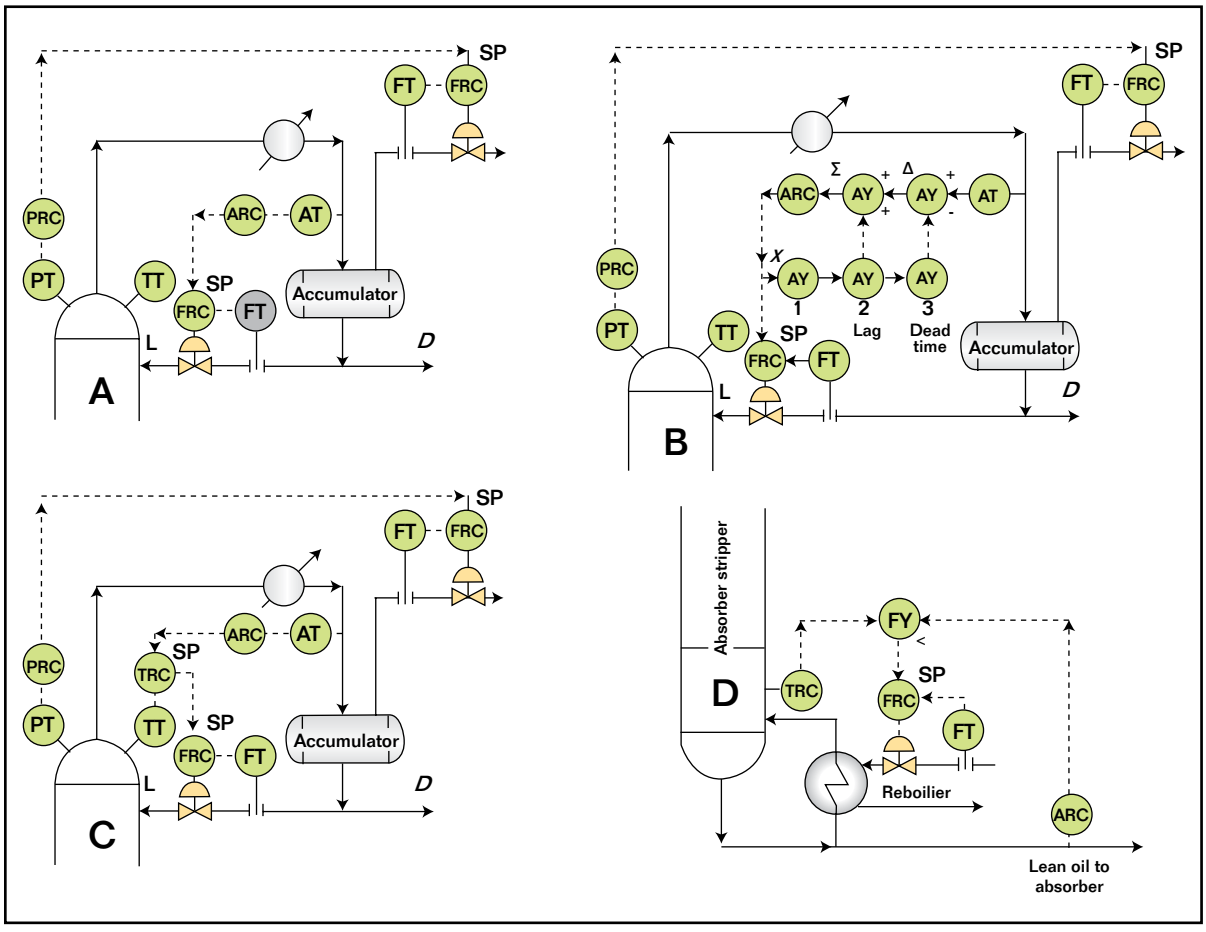
Analyzer controllers in a feedback configuration can be considered only when the dead time caused by analysis update is less than the response time of the process. Naturally, direct control of the composition of the product by an analyzer gives more accurate results than indirect control by temperature. The composition controller provides a feedback correction in response to feed composition changes, pressure variations or changes in tower efficiencies.

In Part A of Figure 9, the analyzer controller (ARC) uses the chromatographic measurement to manipulate the reflux flow by adjusting the set point to the reflux flow controller (FRC). A liquid sample from the condenser run-down line is obtained by a sample probe, and a sample system is used to condition and vaporize the liquid sample to provide a representative vapor sample to the chromatograph.

Smith Predictor

Often the analyzer is slow and introduces a significant delay time that degrades the controllability of the process. In that case, some type of dead time compensation is used. (Section 2.30, Vol. 2, *The Instrument Engineer's Handbook*, 4th edition.) A Smith-predictor compensator can serve to model the process to

FIGURE 9.



Part A: Overhead composition controls by cascading reflux flow as the slave controller (FRC). Part B: The Smith predictor, which is the same as Part A, but with dead time compensation added. Part C: Overhead composition control by triple cascade of ARC to TRC to FRC. Part D: Absorber bottoms composition control (ARC) cascaded to reboiler heat input (FRC) with temperature override (TRC).

predict what the analyzer measurement should be between analysis updates. When the actual measurement is completed, the model's prediction is compared to the actual measurement, and the input to the controller is biased by the difference.

Controllability of the process is degraded by the dead time between measurement updates. (Dead time compensation in detail is discussed in Chapter Section 2.30, Vol. 2, *The Instrument Engineers Handbook*, 4th edition). As shown in Part B of Figure 9, the Smith predictor compensator provides a process model in terms of its time constant and dead time, and thereby predicts what the analyzer measurement should be between analysis updates. When an actual analysis is completed, the model's prediction is compared to the actual measurement, and the input to the controller is biased by the difference.

In Figure 9, Part B, the multiplier (AY*1*), first-order lag (AY2) and dead time (AY3) provide the required inputs to the calculation of the predicted analysis. This predicted response is subtracted from the actual measurement (AY±) to give a differential of the actual process from its own model. This delta is added to the model (AYΣ) without dead time to provide a modified pseudo-measurement to the analyzer controller. Thus, the analyzer measurement, which has a significant dead time due to sampling and cycle times, is provided with a trim to obtain the predicted measurement of the model.

Triple Cascade

The purpose of all cascade systems is to provide slaves that will correct for disturbances before they can upset the primary or master controller. Part C in Figure 9 illustrates a triple-cascade loop, where a temperature controller is the slave of an analyzer controller, while the reflux flow is cascaded to temperature. This configuration is used when maintaining a stable temperature on a particular tray, while the tower operates at a constant and controllable pressure is desired. Since temperature is an indicator of composition at constant pressure, the analyzer controller serves only to correct for variations in feed composition.

Cascade loops will work only if the slave is faster than the master, which adjusts its set point. Therefore, in the case illustrated in Part C, the time constants of the flow controller (FRC) must be much smaller than those of the TRC and similarly, the TRC must be faster than the ARC. Another important consideration in all cascade systems is that the integral mode in the master will cause the output to saturate when that output is blocked from reaching and modulating the set point of the slave (when the slave is switched to local set point). This is called "reset windup," and it is prevented by providing the integral mode of the master with an input (an "external reset") that is never blocked. The external reset of the master is always the measurement of the slave (temperature for the ARC, flow for the TRC, in Part C). These signals are not shown in Part C

Selective Control

Part D of Figure 9 illustrates a limit control configuration where the analyzer controller is overruled when the temperature reaches its high limit. The temperature controller is a constraint controller preventing the temperature from exceeding a limit at the bottoms of an absorber stripper. The reason for this limit is energy conservation, because no additional stripping of the light component can be accomplished once the boiling point of the impurity is exceeded. Therefore, even though an analyzer controller may call for more heat, this heat would only increase the bottoms temperature without removing the impurity, thereby wasting heat.

Selective control configurations also require external feedback to protect them from reset windup. In Part D, we have a combination of selective and cascade systems as the master of the FRC is selected to be either the TRC or the ARC. In such a configuration, the external reset (ER) signal (not shown in the figure) is taken from the measurement of the slave controller (FRC).

Part 2

PRESSURE CONTROL

In controlling the pressure of a column, the key pieces of equipment are the condenser and the accumulator. First the overhead vapors enter the condenser (partial or total), and next the liquid condensate is collected in an accumulator vessel. Some of the accumulated condensate is returned to the column as reflux, while the remainder is withdrawn as overhead product (distillate). If the condensation is incomplete, the condenser is called a "partial" condenser, and the overhead product is withdrawn in both vapor and liquid phases.

Total condensers are usually designed for accumulator pressures up to 215 psia (1.48 MPa) at an operating temperature of 120 °F (49 °C). A partial condenser is usually used between 215 psia and 365 psia (1.48 to 2.52 MPa), and refrigerant coolant, such as propane, is used if the operating pressure is greater than 365 psia (2.52 MPa). The condenser pressure drop is usually about 5 psia (34.4 KPa).

Most distillation columns are operated under constant pressure, because at constant pressure, temperature measurement is an indirect indication of composition, but floating the operating pressure can have advantages in many applications. When the column pressure is allowed to float, the composition must be measured by analyzers or by pressure-compensated thermometers. The primary advantage of floating-pressure control is that one can operate at minimum pressure, and this reduces the required heat input needed at the reboiler. Other advantages of operating at lower temperatures include increased reboiler capacity and reduced reboiler fouling.

In the following paragraphs, floating and constant pressure control strategies will be described for operations when (1) liquid distillate is withdrawn in the presence of non-condensables; (2) vapor distillate is withdrawn in the presence of non-condensables; and (3) liquid distillate is withdrawn when the amount of non-condensables is negligible.

Cooling Water Control, Negligible Inerts

In distillation processes where the distillate is in the liquid phase and the quantity of inerts is negligible, the column pressure is usually controlled by modulating the rate of condensation in the condenser. In the control system shown for Column A in Figure 10 (p. 18), the column pressure is controlled by throttling the cooling water flow through the condenser. This control scheme is recommended only when the cooling water is treated, because condenser tube fouling due to high temperature rise across the tubes can occur.

The advantage of this configuration is simplicity and low maintenance cost, because the control valve is on the water side and gives acceptable control performance, provided the condenser is the bundle type, and the cooling water is flowing through the tubes at a rate over 4.5 feet per second (1.35 m/s), or the water has residence time of less than 45 seconds. With such short residence time, the pressure controller requires only a narrow proportional mode. As the residence time increases, the controller will require wider and wider throttling ranges and will also need the addition of an integral mode to compensate for the load changes.

Once the proportional band is wide, the control quality will no longer be satisfactory for precision distillation applications, because the recovery time from upsets will be long, and because proper tuning of the integral mode is prevented, due to the dead time varying with load. Therefore, one should not use this control system when the process time lag is large, such as in a case where a condenser box with submerged tube sections is used, because of the large volume of water in the box.

If the amount of inerts is nearly zero, and one would like to have a more responsive control system, one can relocate the control valve from the water to the condensate liquid line on Column A in Figure 10. In such a configuration, when the column pressure is dropping, the pressure recorder controller (PRC) reduces the opening of this valve, which causes the condensate level to build up, and the tube surface exposed to the condensing vapors is reduced due to the flooding of the heat transfer tubes. This reduces the rate of condensation and increases the column pressure. If it is expected that over time the inerts will accumulate and blanket the condenser, a vent valve should be added to the condenser.

Distillate with Inerts

The problem of pressure control can be complicated by the presence of large quantities of inert gases, which must be removed, because otherwise they will blanket the condensing surface and cause the loss of pressure control. In refinery applications, the non-condensables are often sent to the fuel gas system or to flare. In other cases, a fixed flow of inerts and vapors is purged to a lower pressure absorption tower or to a vent condenser, where the condensable vapors are recovered.

When the amount of inerts is variable, the purge stream has to be modulated. In the control system shown for Column B in Figure 10, as the inerts build up in the condenser, the pressure controller will first open up the control valve (PCV-1), which lowers the flooding level in the condenser and, by increasing the heat transfer surface area, also increases the rate of condensation. When PCV-1 is nearly fully open (say 95%), the valve position controller (VPC-2) is activated. Its set point is 95%, so when PCV-1 has opened to 95%, it starts purging the inerts by opening control valve (VPCV-2).

The control system shown for Column C in Figure 10 is used when the distillate is in the vapor phase and inerts are present. As the overhead product is removed under pressure control, the system pressure will quickly respond to changes in the distillate vapor flow.

Here, the level of the overhead receiver regulates the condensate generated by throttling the cooling water supply. This way it will condense only enough material to provide the column with reflux.

This control system will give acceptable performance only if the condenser residence time on the coolant side is short. If this is not the case, the cooling water flow should be held constant, and the control system shown for Column

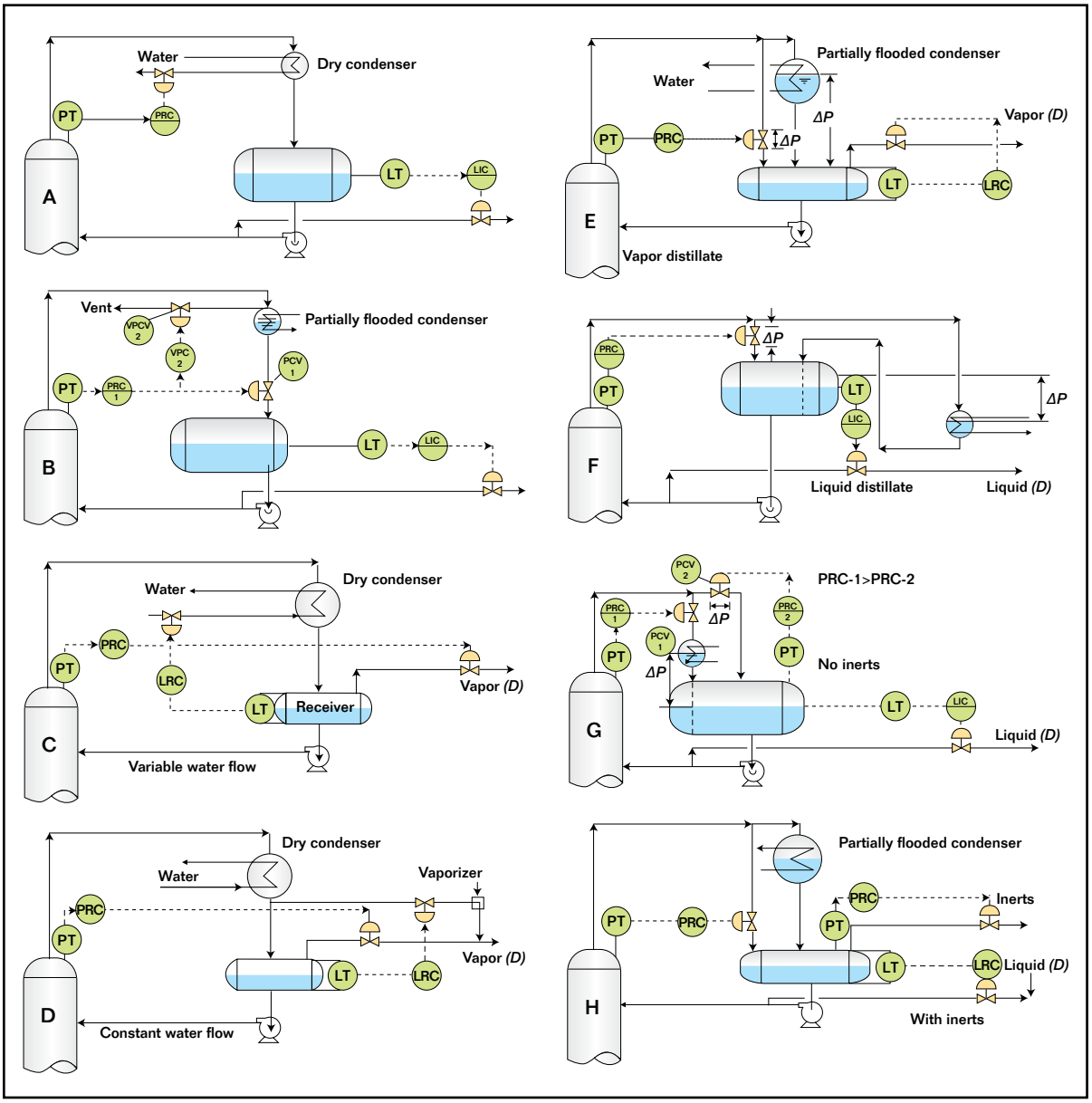
D in Figure 10 should be used. In this case, some liquid condensate is sent to a small vaporizer, which flow is regulated by the accumulator level, and the resulting vapors are mixed with the vapors from the pressure-control valve.

In the main fractionators of fluidized catalytic cracking units (FCCU) and in the crude towers, steam turbine driven compressors are used to "draw" the vapors from the columns that operate at essentially atmospheric pressure. In these cases, the column pressure is usually controlled by modulating the steam supply to the turbine.

Bypass Control Configurations

If the cooling water can cause fouling, or if the use of a vaporizer represents a substantial waste of energy, the control system shown for Column E in Figure 10 is recommended for vapor distillate applications. Here, instead

FIGURE 10.



Pressure control system configurations for no inerts: A: By CW (cooling water) throttling, with inert purging; B: By adjusting level of condenser flooding; C: By CW throttling; D: With vaporizer bypass throttling; E: By bypass throttling with vapor distillate; F: By bypass throttling with liquid distillate; G: By bypass throttling without inerts; H: By bypass throttling with inerts.

of re-vaporizing the condensate, part of the overhead vapor stream is never condensed, but is sent to a vapor bypass valve around the condenser. This allows the condensate flow to stay constant, while varying the rate of condensation by changing the degree of flooding of the tubes.

Column F in Figure 10 illustrates the controls when, in the case of liquid distillates with negligible inerts, the condenser needs to be mounted below the accumulator for ease of maintenance or to save on the steel supports. In this configuration, when the bypass valve is open, the condenser is flooded. When the column pressure drops, the PRC increases the opening of the bypass valve, which increases the pressure in the accumulator and pushes some of the condensate back into the condenser to reduce its rate of condensation. If the column pressure rises, the opposite occurs, and the rate of condensation increases.

It is usual practice to elevate the bottom of the accumulator 10 ft to 15 ft (3 m to 4.5 m) above the suction of the pump in order to provide the required positive suction head. Under normal operating conditions, the subcooling that the condensate received in the condenser is sufficient to reduce the vapor pressure in the receiver. The difference in pressure permits the condensate to flow up the 10 ft to 15 ft of pipe between the condenser and the accumulator.

When the condenser is mounted above the accumulator and no inerts are present, two bypass valves can be used, as shown for Column G in Figure 10. The column pressure is maintained by PRC-1, which is throttling the flow of vapor through the condenser, while PRC-2 is maintaining the accumulator pressure by throttling the bypass. This configuration provides faster pressure regulation for the column.

In most applications, some inerts are also present, and in such cases the controls shown for Column H in Figure 10 are applicable. Here, the column pressure controller is throttling the hot vapor bypass, and the accumulator pressure controller is throttling the vent valve of the inerts. The accumulator PRC is set at about 5 psig below the tower pressure.

Optimization: Minimum Pressure

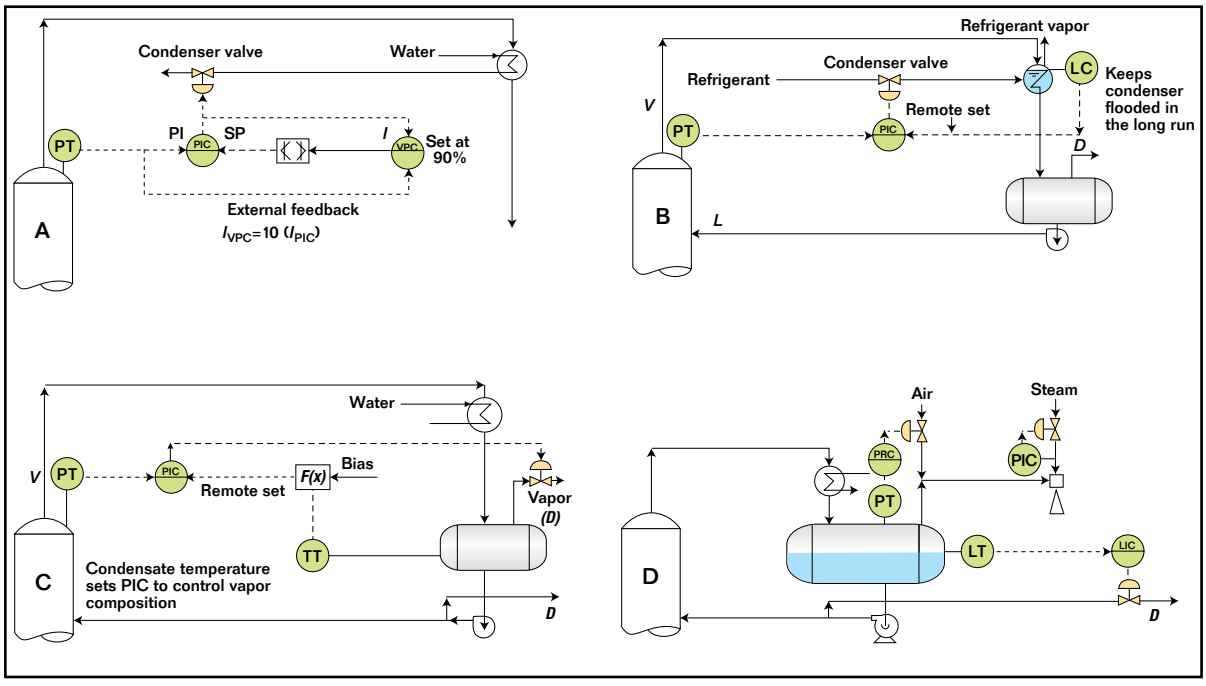
Operating the column at the minimum possible pressure minimizes the energy cost of separation within the constraints of the system. Lowering this pressure increases the relative volatility of distillation components, and thereby increases the capacity of the reboiler by reducing the operating temperature, which also results in reduced fouling. Reducing pressure also affects other parameters, such as tray efficiencies and latent heats of vaporization.

Yet composition control must take precedence over pressure optimization, and therefore the pressure optimization loop response must be much slower than that of the composition control loop. The effects of pressure changes also must be considered on the upstream and downstream units. For example, if the pressure of an upstream tower provides the driving force to move product to a downstream tower, pressure minimization may not be practical. Fractionators using vapor recompression, such as a propylene splitter (Figure 13, p. 21) with a heat pump, may actually benefit from increasing the pressure rather than reducing it.

Minimum pressure operation can be achieved by adjusting the set point of the pressure controller so as to keep the condenser fully loaded at all times. In order to prevent upsets caused by rapid set point changes, the valve position control (VPC) scheme shown for Column A in Figure 11 (p. 20) can be used to minimize the set point of the pressure controller, and thereby keep the condenser control valve in nearly the fully open position. The VPC should be an integral-only controller, with an integral time setting approximately 10 times that of the overhead composition controller. In addition, bumpless transfer from VPC to direct PIC control should be guaranteed by providing external reset, and the set point adjustment capability of the VPC should be limited if required by the process.

The controls for Column B in Figure 11 serve to minimize the operating pressure of a partial condensing system with a vapor distillate product. Here the level controller (on the refrigerant side) serves to fully load the condenser in the long term, while the pressure controller corrects the short-term upsets.

FIGURE 11.



Optimization and vacuum control strategies. A: Minimizing (floating) pressure by maximizing coolant valve opening; B: Floating pressure control of partial condenser by maximizing condenser level; C: Floating pressure control when the distillate is both vapor and liquid; D: Vacuum controlled by air bleed-in.

An additional control loop is required to control the composition of the vapor product (Column C, Figure 11) if the column pressure is floated and both liquid and vapor distillates are produced. In this control system, the column pressure is controlled by throttling the flow of the vapor distillate, while the set point of this PIC is adjusted if changes occur in the accumulator (condenser outlet) temperature, which is characterized to represent the desired composition. If changes are to be made in the product composition, the bias is adjusted.

Vacuum Systems

To separate some liquid mixtures, the temperature required to vaporize the feed would need to be so high that decomposition would result. These columns are operated under vacuum and steam jet ejectors can be used, singly or in stages, to generate the required vacuum. Ejectors have no moving parts and require little maintenance, but are designed to operate at a fixed capacity and steam condition. Increasing the steam pressure above the design level will not increase and sometimes will decrease the capacity of the ejector because of choking. Reducing the steam pressure causes unstable operation.

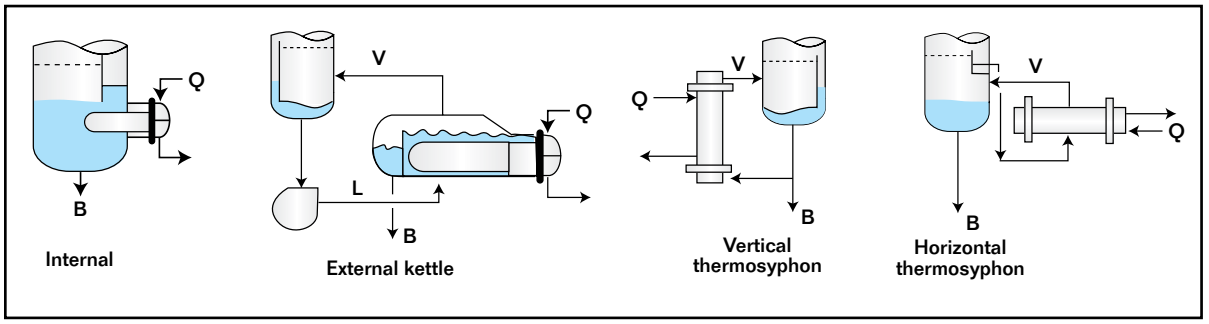
For the above reasons, the steam pressure is controlled at a constant value (PIC) in the configuration for Column D, Figure 11. In order to reduce the time lag, instead of the column pressure, it is the accumulator pressure that is controlled by the PRC, which modulates the amount of air or inert gas that is bled into the system. At low loads this results in a substantial waste of steam. Therefore, operating costs can be lowered by installing two ejectors and by automatically switching between the larger one and a smaller one to match the load.

Reboilers

As shown in Figure 12 (p. 21), reboilers can be internal or external, and operated by either natural or forced circulation. The kettle reboiler is the most common design for external forced-circulation applications. Thermosyphon reboilers (vertical or horizontal) operate by natural circulation, which is induced by the hydrostatic pressure imbalance between the liquid inside the tower and the two-phase mixture in the reboiler tubes.

A newer development is the use of self-cleaning, shell-and-tube heat exchangers for applications where heat exchange surfaces are prone to fouling. Common heat sources include hot oil, steam, fuel gas (fired reboilers) and the condensation of compressed vapors in vapor recompression systems.

FIGURE 12.



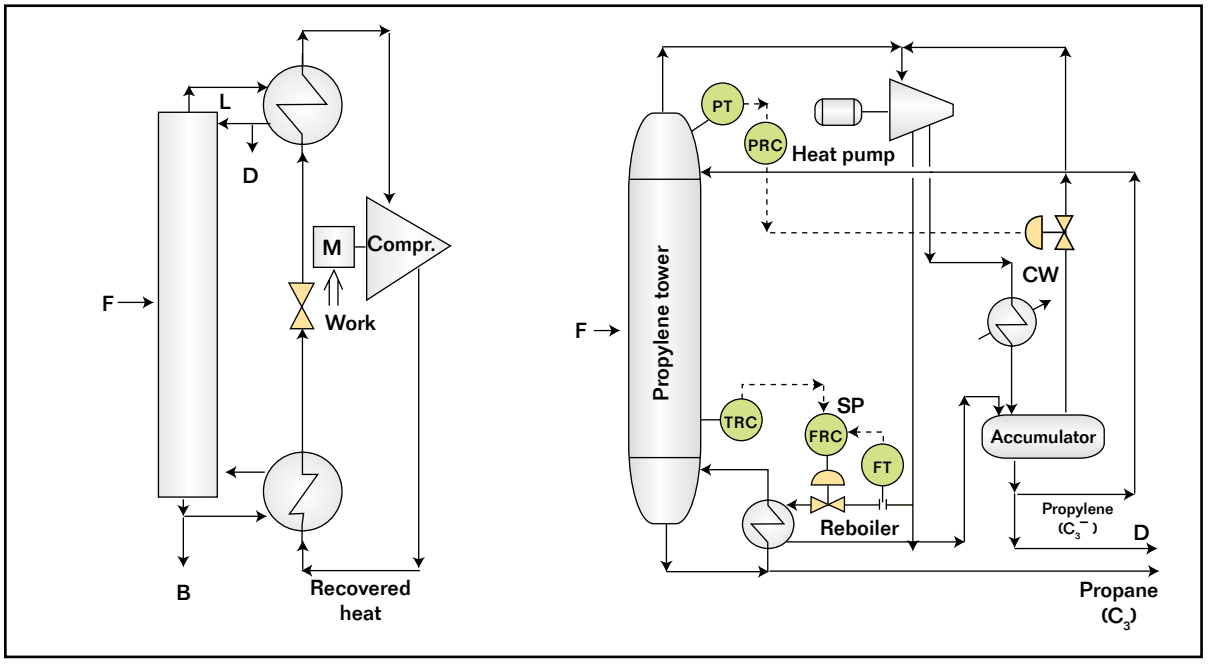
Reboiler design variations are distinguished by the type of circulation—natural and forced (pumped)—and by the point from which they take their liquid. Most reboilers take their inlet from the column bottoms, while horizontal thermosyphon reboilers take it from the bottom trays.

Vapor Recompression

Vapor recompression is another means of improving energy efficiency. As shown in Figure 13, the overhead vapor from the distillation column is compressed to a pressure at which the condensation temperature is greater than the boiling point of the process liquid at the tower bottoms. This way the heat of condensation of the column overhead is reused as the heat for reboiling the bottoms. This scheme is known as vapor recompression.

It is often used when the boiling points of the top and bottom products are similar. Examples of such processes are the cryogenic demethanization processes, where the column pressure is controlled by throttling the speed of the recompression compressors or the propylene fractionation process. As shown on the right of Figure 13, the propylene column pressure is controlled by throttling the bypass valve around the vapor recompression heat pump.

FIGURE 13.



Left: The vapor recompression system uses recovered heat. Right: The pressure of such a distillation process can be controlled by modulating the speed of the compressor or by throttling the bypass around it.

FCCU main fractionators and crude towers also make use of compressors to "draw" vapors from their essentially atmospheric towers and maintain the column pressure by manipulating the steam flow to the turbine, which sets the speed of the compressor.

FEED CONTROLS

One of the best means of stabilizing the operation of a distillation column is to hold both the feed flow and temperature constant. Feed composition is seldom subject to adjustment. Having constant feed conditions simplifies the operation of the control system. For example, if the feed flow to a column is controlled by a liquid level controller, that controller can be tuned with a low gain, (wide proportional band) so that the level will be allowed to swing over a wide range without causing much change in flow. Nevertheless, if the feed comes from another column, the feed to the second column will eventually change if the load on the first column changes.

As shown in the control loop for Column A in Figure 14 (p.23), the temporary flow fluctuations from the previous column and/or the pressure fluctuations caused by the distillate/reflux pump can be minimized by cascading the feed-flow controller to a nonlinear level master (LRC) on a surge tank. Such level controllers can be so configured that as long as the level is between 25% and 75%, the set point to the FRC remains constant. Naturally, if the level drops below 25% or rises above 75%, the FRC set point is changed to protect the surge tank from being drained or flooded.

Feedback controls are capable of compensating for upsets only after the upsets have occurred and been detected. With feedback controls, the operators have to manually adjust the set points of the loops when plant conditions change. Feedback control is usually sufficient to keep the distillation column in operation, but it is not sufficient to provide optimal performance. Usually, feed-forward-based product composition control of distillation can provide 5% to 15% energy savings. The goal of a feed-forward control application can be to maintain the material balance of the column as feed rate varies.

Feed-forward controls represented the first steps towards model-based process control. They were first applied in well-understood processes, such as heat transfer and distillation, where the firm knowledge of material and heat balance equations made it possible to predict and anticipate the consequences of changes in the inputs before they had time to evolve, and to take corrective action before the tower is significantly affected. This correction is accomplished by considering process dynamics (dead times and time lags), the nonlinearities between separation efficiency and column loading, loop interactions and process measurements.

If feed-flow variations are unavoidable, the impact of these disturbances can be reduced by feed-forward correction of the material balance (Column B in Figure 14). In this configuration, if the distillate flow is changed in the right proportion (*m*) and at the right time, the upsets resulting from feed-flow variations can be minimized. This dynamic compensation (*FY*) is not only a ratio adjustment, but must also be tuned to reflect the time constant and dead time of the process. However, if the feed composition changes, the value of m will have to be readjusted.

Maximizing the Feed Flow

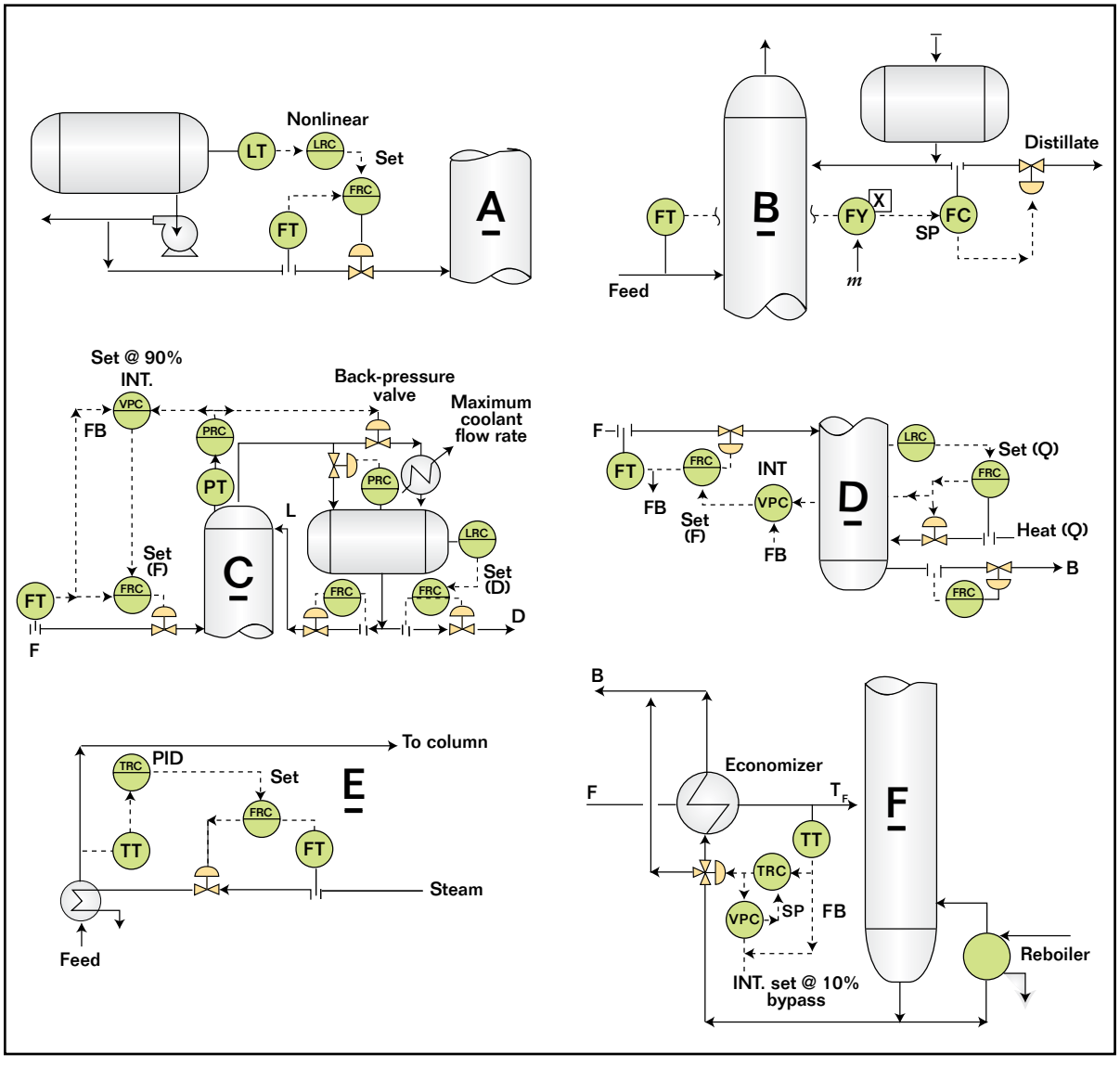
In cases where both the demand for the product and the availability of feedstock is unlimited, increasing the throughput of the column maximizes profitability. In such installations, a valve-position controller can be cascaded to the feed-flow controller in order to increase the feed rate until an equipment constraint is reached. In many cases, it is not known which constraint will be encountered first, as feed flow is maximized because the limiting constraint may vary over time. In such cases, a multiple constraint network is implemented. In such a configuration, the outputs of the valve-position controllers (VPCs) detecting the loadings of the critical equipment are sent to a low-signal selector, and the feed flow is set by that.

If the constraint on maximum production is the cooling capacity of the condenser, the opening of the column-pressure control valve can be measured (VPC on Column C in Figure 14), and the feed rate can be manipulated by the valve-position controller to keep the back-pressure control valve always nearly open (VPC set at 90%). If a hot-vapor bypass around the condenser exists, as is the case on Column C, the opening of the bypass valve can also be used to indicate the condenser load.

If the constraint on maximum production is the heating capacity of the reboiler, a similar cascade control configuration can be used to keep the reboiler fully loaded (Column D, Figure 14). This strategy is particularly effective when the cost of reboiler heat is negligible, such as when a waste steam is used that otherwise would be vented.

The VPCs used on Columns C and D are usually selected as integral-only controllers and are set at around 90% of valve stem lift, which on an equal-percentage valve corresponds to about 70% of maximum flow. The integral time setting of the VPCs is slow, about 10 times the integral setting of the FRC, which the VPC adjusts in a cascade arrangement. In order to eliminate reset windup, the VPC is provided with external feedback (FB) from the slave transmitter (FT, the feed-flow transmitter).

FIGURE 14.



Feed control systems. A: Surge tank with non-linear level controller as cascade master; B: Feed-forward adjustment of distillate product flow; C: Maximizing throughput by fully loading the condenser; D: Maximizing throughput against a reboiler constraint; E: Feed pre-heater controls; F: Maximum recovery of the heat content of bottom product by economizer.

Feed Temperature and Enthalpy Control

The thermal condition of the feed determines the required heat input at the reboiler, but constant temperature alone does not necessarily mean constant feed quality. For best separation efficiency, the feed should be preheated to its bubble point (the temperature at which bubbles first appear when a liquid mixture is heated—when the lightest component in the feed starts to boil). If the composition of the feed varies, its bubble point will also vary. As the feed becomes lighter, some of it will vaporize, but this variation can be handled by subsequent controls.

The controls shown for Column E in Figure 14 show the cascade controls for a steam-heated preheater. In this configuration, the temperature controller usually is a three-mode (PID) controller. The role of the derivative mode (*D*) during start-up is to provide a large correction initially, which helps to quickly get the unit up to stable operation.

An economizer can be used to preheat the feed by sending it through an economizer (Column F, Figure 14), which is heated by the bottoms product. The economizer operation is optimum when the amount of heat recovered from the bottoms product is maximized. To achieve this goal, a valve position controller (VPC) is used as the cascade master of the feed temperature controller, which controls the bypass around the economizer. The goal of optimization is to keep the bypassed flow at a minimum. Therefore the VPC is usually set at about 10% of valve opening.

Control of feed enthalpy instead of temperature can be achieved if the right measurements are available. For example, if the cold feed (at temperature T_F) first passes through the bottoms-to-feed economizer and then through the steam preheater, one can calculate the steam flow required to maintain the feed enthalpy constant by the calculation:

$$S = [F(\Delta H_F) - C_{PF}T_F) - B(C_{PB}\Delta T)]/\Delta H_{stm}$$

Where:

ΔH_F = feed enthalpy as it enters the tower, BTU/lb (kcal/kg)
F = feed flow to preheater, lb/hr (kg/hr)
S = steam flow, lb/hr (kg/hr)
C_{PF} = feed heat capacity, BTU/lb °F (kcal/kg °C)
T_F = feed temperature before preheaters, °F (°C)
B = bottoms flow to economizer, lb/hr (kg/hr)
C_{PB} = bottoms heat capacity, BTU/lb (kcal/kg)
ΔT = bottoms temperature difference through the economizer, °F (°C)
ΔH_{stm} = steam heat of vaporization minus condensate heat, BTU/lb °F (kcal/kg °C)

The primary effect of increasing feed enthalpy is to decrease vapor-liquid circulation below the feed tray relative to that above the feed tray. When preheating the feed is less expensive than reboiler heat, or when the reboiler is the limiting constraint on production rate, the process is optimized by maximizing feed preheat. When condenser capacity is limiting or flooding is encountered above the feed tray, preheat should not be used.

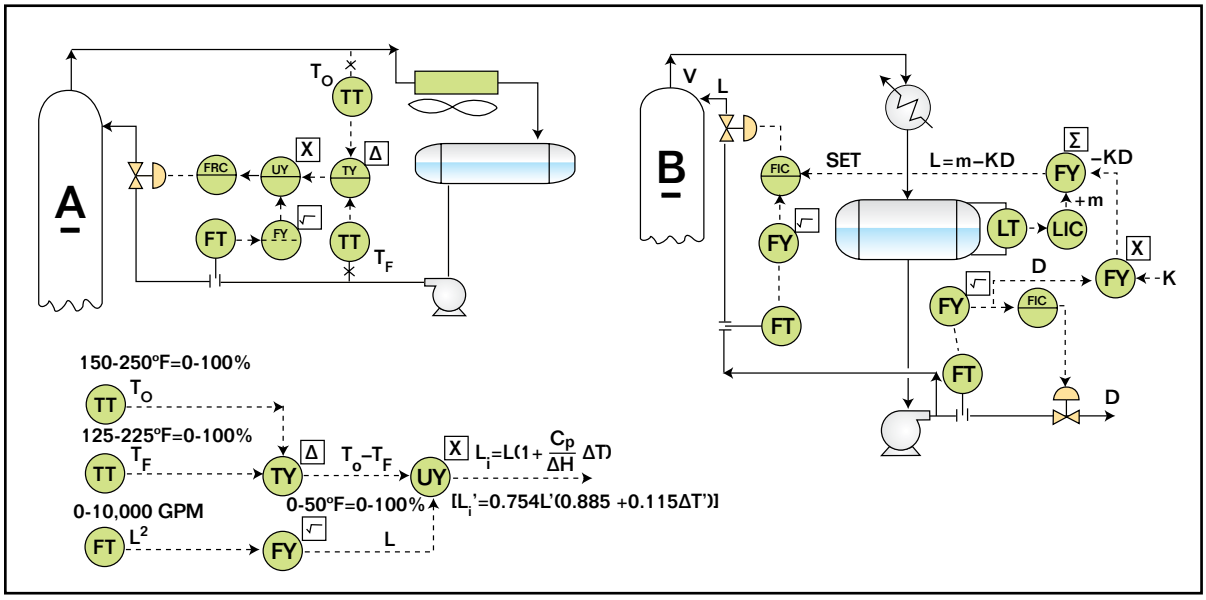
REFLUX CONTROLS

Stable column operation is guaranteed by keeping the internal reflux of the distillation tower constant. Consequently, internal reflux controls are designed to compensate for changes in the temperature of the external reflux caused by ambient conditions. Column A, Figure 15 (p. 25) is controlled by a typical internal reflux control system. The equations for calculating the required external reflux rate are shown at the bottom. This control system corrects for either an increase in overhead vapor temperature or a decrease in external reflux liquid temperature and, obviously, the required control actions are in the opposite direction.

The need for internal reflux control can be eliminated in some cases when the flow of distillate product is controlled under the cascade control of the accumulator level. If the speed of response of this control system is not sufficient, it can be speeded up by reducing of the accumulator volume. To overcome the accumulator lag, the reflux rate, *L*, is manipulated in direct proportion to a change in distillate rate, *D*, rather than by waiting for the response of a level controller. In the control system for Column B, Figure 15, when $K = 0$, the reflux is adjusted by the level controller only. In other cases, the reflux flow is immediately altered by some percentage for a change in distillate, and the level controller forces the balance of the change. The response is a first-order lag.

If $K = 0.5$, the reflux flow is changed to the exact new steady-state value, because *K* equals the ratio of D_{max}/L_{max}, and therefore the computation is exact. The lead equals the lag and the net effect is the elimination of dynamic contribution. If $K = 1.0$, the initial response is a first-order lead-lag function. In this case, the reflux is greater than required for the new steady state, and the level controller eventually corrects the flow. The value of *K* affects the transient response only, but does not change the steady-state flow; therefore, it can be used to adjust the dynamics of the loop. The greater the value of *K* the faster is the response. In order to maintain stability, *K* should not be set greater than 0.75.

FIGURE 15.



Left: The controls required to keep the internal reflux flow of the column constant; Right: The controls needed to eliminate the effect of accumulator lag in controlling the external reflux to a column.

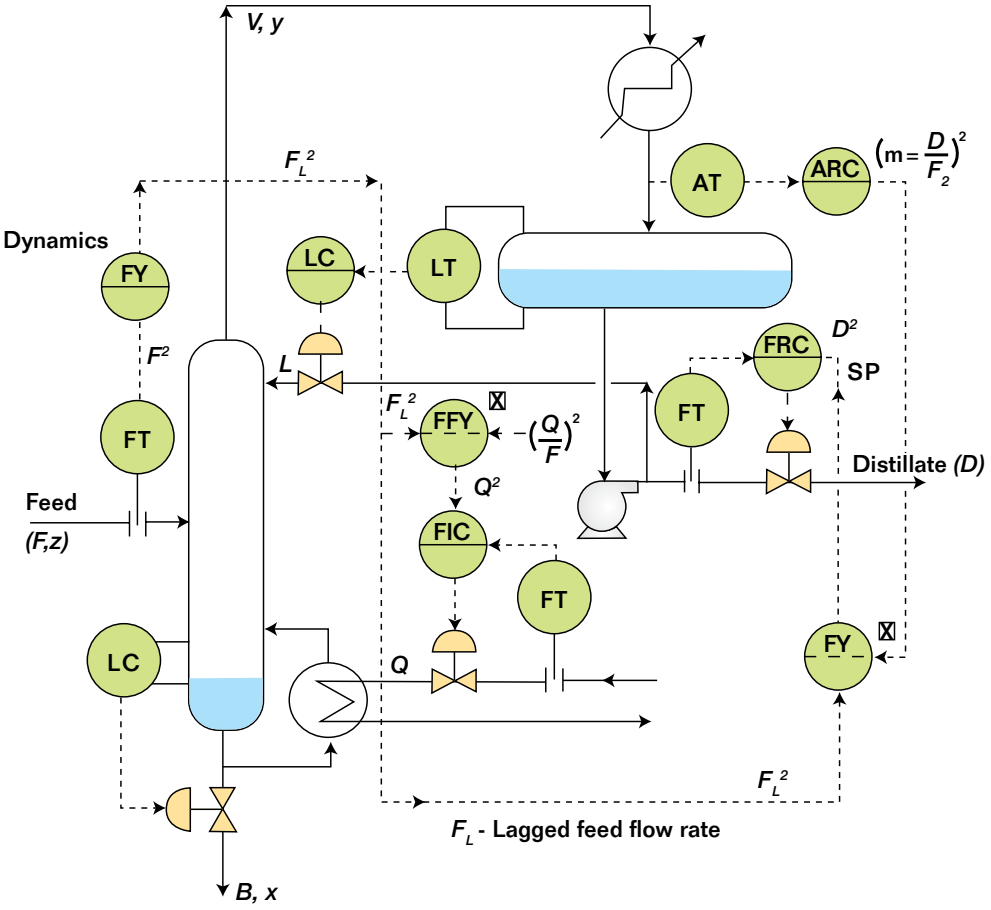
Part 3

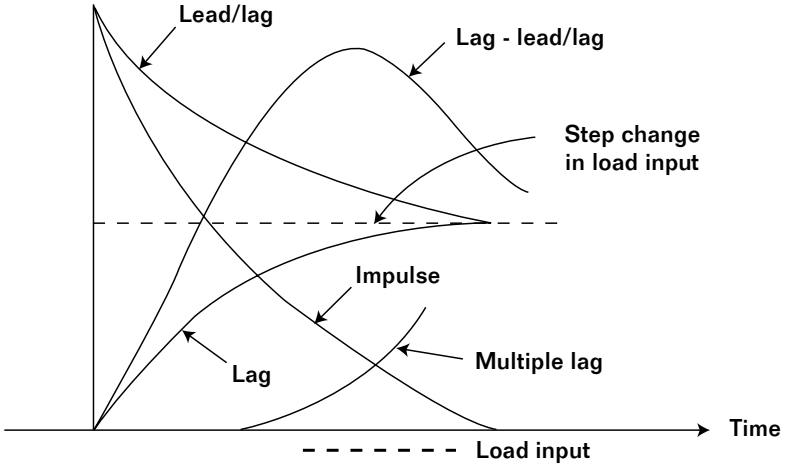
CONTROLLING THE WHOLE COLUMN

Previously, I have discussed the individual loops used to control the different segments of the distillation process. Now I will talk about various control strategies that can be used to control and optimize the complete column. Here, I will treat distillation as a single unit operation. In unit operations' control, the individual column variables are treated only as constraints and so long as the values of these constraints are within acceptable limits, the column is controlled (optimized) to maximize production rate, profitability, etc.

In order to optimize the unit operation of distillation, one not only wants to keep the top and bottom product compositions within specification, but might also want to maximize the throughput, minimize energy use, enhance column stability or operate against equipment constraints. In addition, one might shift less profitable components into more profitable products. To optimize the profitability of the operation, one must know the acceptable variations in product specifications, the relative economic values of the product streams and cost of energy used in the separation.

FIGURE 16.





Top: Feed-forward control system that provides constant separation by manipulating the distillate flow. Bottom: A variety of dynamic compensators that can be used to match the "dynamic personality" of the process.

Economics of individual fractionators may continually change throughout the life of the plant, because energy savings can be important at one particular time, while product recovery can be more important at other times. Other goals include the desire to minimize the disturbances propagated to downstream units, minimize rework or recycle of off-spec products and maximize the consistency of product quality. Thus, a given column's operating economics, and, therefore, its optimization objectives may change with time.

CONSTANT SEPARATION

A distillation column operating under constant separation conditions has one fewer degree of freedom because its energy-to-feed ratio is constant. This means that for each concentration of the key component in the distillate, a corresponding concentration exists in the bottoms. Therefore, if the feed composition is constant, and if the concentration of a component in one product stream is held constant, that fixes the concentration of that component in the other. Figure 16 (p. 26) shows a feed-forward control system for constant separation in which distillate is the manipulated variable.

The control system is so adjusted that the output (m) of the trim analyzer controller ARC is 50% when the design or normal distillate-to-feed ratio is required. If the gain of the multiplier (FY) is set at 2, the output tracks the load when this normal distillate-to-feed ratio occurs.

The block labeled "dynamics" in Figure 16 is a special module that serves dynamic correction by modifying the transient response. This matches the time response of the distillate to a feed-rate change. The dynamic block most often is a combination of a dead-time and a lead-lag module in series, which in the steady state makes no correction at all; its output equals its input. Because the "dynamic personalities" of the various distillation processes are different, a variety of dynamic compensators are available to match them, as shown at the bottom of Figure 16.

Maximum Recovery

If one product, such as the distillate (D), is worth much more than the other, the control system can be designed to maximize the more valuable stream. If we assume that energy is free, the material balance for distillate in such a column can be expressed as $D = m(KF + K_2F^2)$ where:

D = distillate rate
F = feed rate
K = adjustable coefficient
$K_2 = 1 - K$
m = feedback trim

Figure 17 (p. 28) shows the controls for this maximum recovery system. In this configuration, the boil-up (heat input rate) is constant, and consequently the distillate product flow is not linear with the feed rate. To increase the response of the system (minimize accumulator lag), the reflux flow set point is adjusted by the distillate flow measurement.

A summing block (FY-1) is provided to compute ($KF + K_2F^2$). The values of m can be calculated on the basis of feed composition. A typical range for m is 0.35 to 0.65. This is the output signal range of the feedback controller (ARC-2). While the coefficients can be calculated with reasonable accuracy, online adjustment is also easy. (These coefficients are accessible in most DCS and PLC systems.)

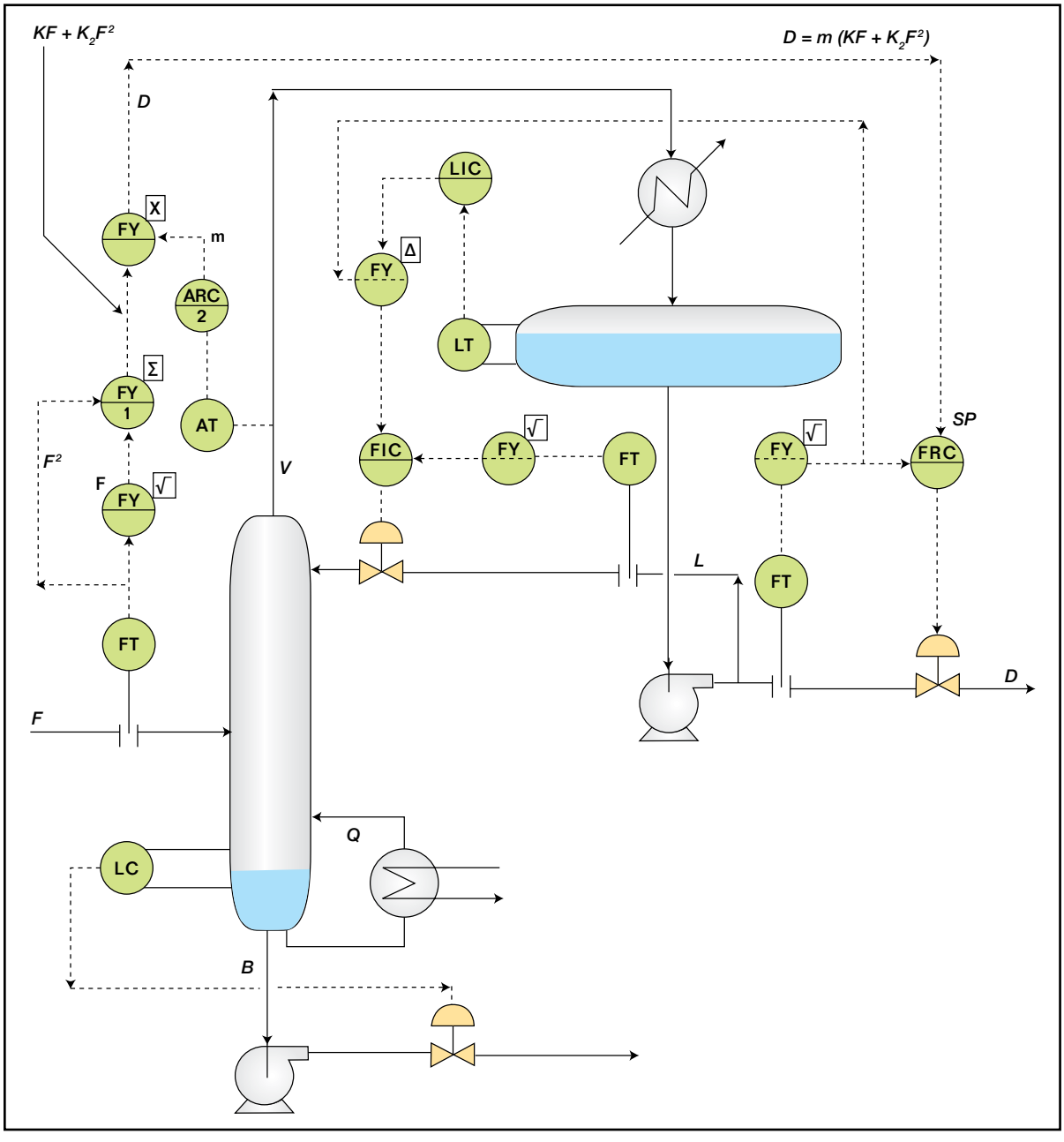
If energy is not assumed to be free, and the composition of only one product needs to be controlled, then the distillate flow can be determined by the following linear relationship: $D = m_1(K_3F)$.

Composition Control of Two Products

If the composition specifications for both products are tight, the controls described in Figure 17 (for the constant separation strategy) are not good enough. This is usually the case whenever the feed composition is unpredictable. In such cases, one must directly control the compositions of both products. The main benefit of dual composition control is minimized energy consumption. The main limitation is caused by the severe interactions between the two composition loops.

Naturally, the feed composition and the tower design determine the compositions that can be achieved. The left side of Figure 18 (p. 29) gives an example of a feed-forward dual composition control system. In this con-

FIGURE 17.

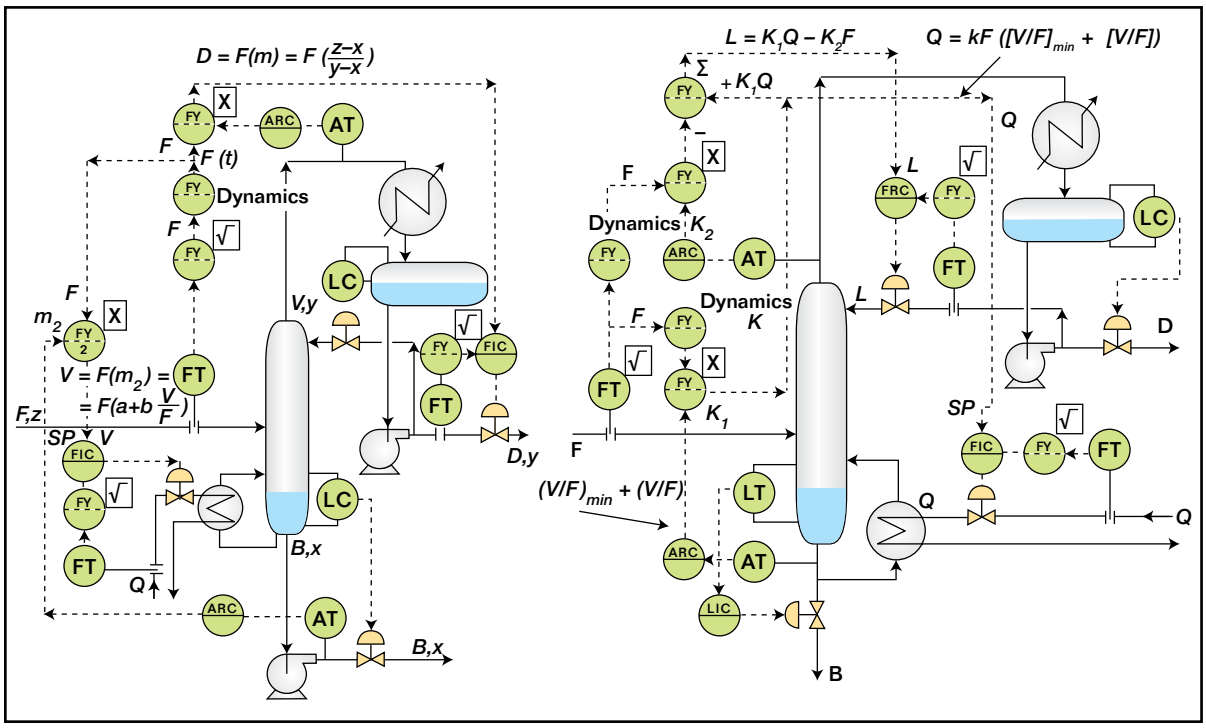


Control system that guarantees the maximum recovery of the more valuable product, in this case, the distillate (D).

figuration, the distillate flow is manipulated to control distillate composition by maintaining the relationship $D = F\,[(z\text{-}x) / (y\text{-}x)]$. However, in order to also enforce composition control of the bottom product, an additional manipulated variable is needed. This cannot be another product stream, because that would conflict with the material balance and change the accumulation in the column. Consequently, the bottoms composition (x) has to be controlled by manipulating the energy balance of the column.

The relationship between separation (S) and the ratio of boil-up to feed (V/F) over a reasonable operating range is $V/F = a + bS$, where a and b are functions of the relative volatility, the number of trays, the feed composition and the minimum V/F. The control system therefore computes V based on the equation $V = F(a + b[V/F])$, where $[V/F]$ equals the desired ratio of boil-up to feed.

FIGURE 18.



Left: Configuration for controlling the composition of both products of a distillation column without much interaction; Right: With interaction.

The column on the left side of Figure 18 illustrates how the two product composition controllers (ARC) are configured to throttle the material balance (*D*) and energy balance (*V*) to maintain the compositions of both products.

The loops include multipliers (*FY-1* and *FY-2*) and the FY block labeled "dynamics." This last block serves the function of dynamic compensation, the role of which has already been explained in connection with Figure 16.

Two Products with Interaction

Interaction is unavoidable between the material and energy balances in a distillation column. The severity of this interaction is a function of feed composition, product specification and the pairing of the selected manipulated and controlled variables.

Severe interaction is likely to occur when the composition controllers of both products are configured to manipulate the energy balance of the column. An example of such a case is when reflux flow and steam flow are manipulated by the two product composition controllers (Figure 18). In such cases, the heat input is $Q = kF([V/F]_{min} + [V/F])$ while the reflux rate is $L = K_1Q - K_2F$. Therefore, both product flow rates are dependent on energy balance terms.

This configuration, without decoupling, results in severe interaction, because if the composition controller on the bottom product is increasing heat input to the reboiler, this action will force the overhead composition controller to increase reflux flow in order to increase heat withdrawal. Therefore the two composition controllers "fight" each other.

In the decoupling equations just defined, the values of K_1, K_2 and k are determined by using the actual process values of $[L/F]$, $[V/F]_{min}$ and $[V/F]$. The decoupling scheme shown on the right side of Figure 18 serves to minimize the tendency of a change at one end of the column upsetting the controller at the other end by implementing the decoupling equation $L = K_1Q - K_2F$. The system is now half-decoupled: A change in heat input at the bottom will not upset the top composition, because the decoupling loop adjusts the reflux independently. However, the heat input is still coupled to reflux, because a change in reflux will still cause the bottom temperature controller to adjust steam flow. This type of half-decoupling is enough to reduce the interaction approximately twenty-fold.

The limitations of decoupling include the fact that overrides can drive the loops to saturation when con-

straints are encountered. Also, in some distillation columns, small measurement errors can transform a system that provides complete decoupling into one that provides no control at all. Since the proper decoupler gains (K_1 , K_2 *and* k) depend on the process gains, and because process gains inevitably change with variations in feed rate, product specifications and column characteristics, these systems require constant attention and adjustment. Such adjustments are beyond the capability of the average plant operating personnel and require sophisticated column models.

The difficulties associated with the application of decoupling systems have prompted a re-examination of interaction itself. The problem may be postulated in two ways:

1) For a given column, is the interaction equally strong in each of the possible control configurations?

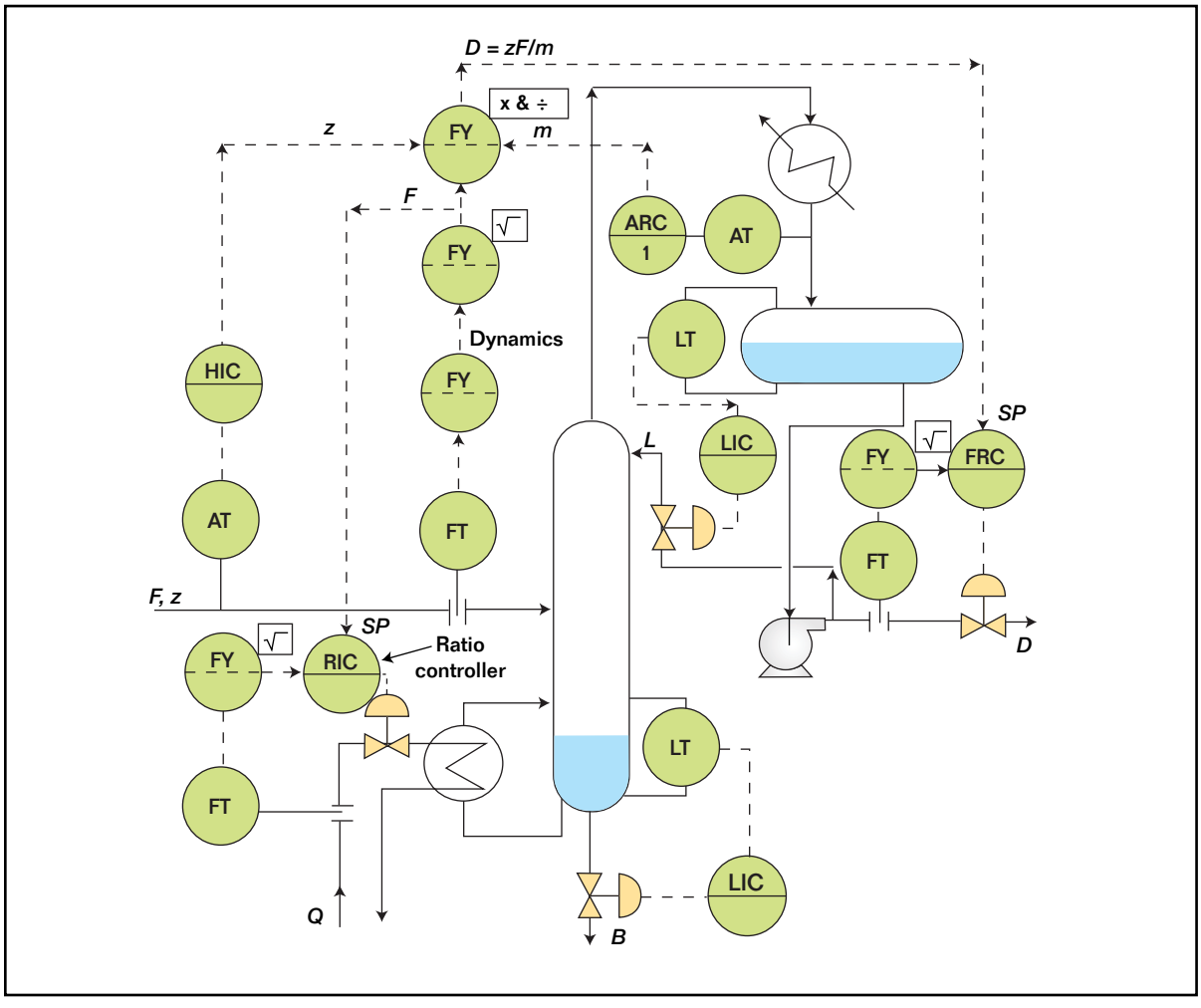
2) For a given control configuration, will the interaction be the same for every column to which it is applied?

Some of the answers to these questions appear in the earlier discussion of relative gain calculations. One general rule worth remembering is that the composition controller for the component with the shorter residence time should adjust vapor flow, and the composition controller for the component with the longer residence time should adjust the liquid/vapor ratio.

Feed Composition Compensation

If the changes in feed composition occur too fast to be handled by feedback control, feed-forward compensation is required. If z is the concentration of the key component in the feed, the material balance, solved for distillate

FIGURE 19.



Feed composition based feed-forward control of distillation.

flow is $D = F\,[(z\text{-}x) / (y\text{-}x)]$, and therefore, when z is measured, the equation for distillate can be simplified to $D = zF/m$, where m is the output of the overhead analyzer feedback trim controller (ARC-1 in Figure 19, p. 30).

The auto manual station (*HIC*) is used in the event of analyzer failure. Dynamic compensation is included (*FY*) in the measurement of the feed flow (*F*). The control of the bottoms flow (*LIC*) is indirectly obtained by the feed-forward control of the reboiler heat input, which is also based on the dynamically compensated feed flow rate.

MULTIPLE PRODUCT DISTILLATION

Most separations involve multiple components and produce two or more liquid or vapor products. Sometimes only one product is withdrawn at a time. When there is a side-stream product in addition to the overhead and bottom products, an additional degree of freedom is available for the control system, because the overall material balance becomes $F = D + C + B$, where C is the side-stream flow rate. Therefore, two product streams can be manipulated to control variables, while the material balance can still be closed by the third product flow. The presence of this added degree of freedom makes the careful analysis of the process even more essential to avoid mismatching of the manipulated and controlled variables.

If the feed rate and column pressure are constant, five degrees of freedom exist: three composition specifications and two levels. These five controlled variables can manipulate the material and heat balance by throttling three product flows and the loading of two heat exchangers (V and L). As was detailed in Part 1, interactions among the loops can be minimized by determining the relative gains of the possible combinations of the manipulated and control variables.

After such evaluation, one might arrive at the controls shown on the left side of Figure 20 (p. 32). Here the ratio of heat input to feed and, therefore, boil-up to feed, is held constant, and separate dynamic elements are used for the distillate loop, the heat input and the side-stream loops.

Multi-product fractionators—crude towers, vacuum towers and FCCU main fractionators—are common in the refining industry. Product quality is often based on true boiling point (TBP) cut points, which approximate the composition of a hydrocarbon mixture and are numerically similar to the American Society for Testing and Materials' 95 percentage points. As was shown in Figure 7 (p. 12), an artificial neural network can calculate—on the basis of local temperature, pressure, steam flow and reflux data—the 95% boiling point or TBP cut point of the products.

If there is a boiling point analyzer, the ANN-based approximation can be used as the measurement for a fast inner slave loop of a cascade configuration in which the analyzer provides the measurement for the slower master, which trims the set point of the slave. Because of the volume of liquid/vapor loads within most multi-product fractionators, the greatest control sensitivity and the quickest response are usually obtained by throttling the product flows. Heat balance is often controlled by throttling the pump around reflux flows, as shown on the right side of Figure 20. The goal is to maximize the amount of heat that is transferred to the feed.

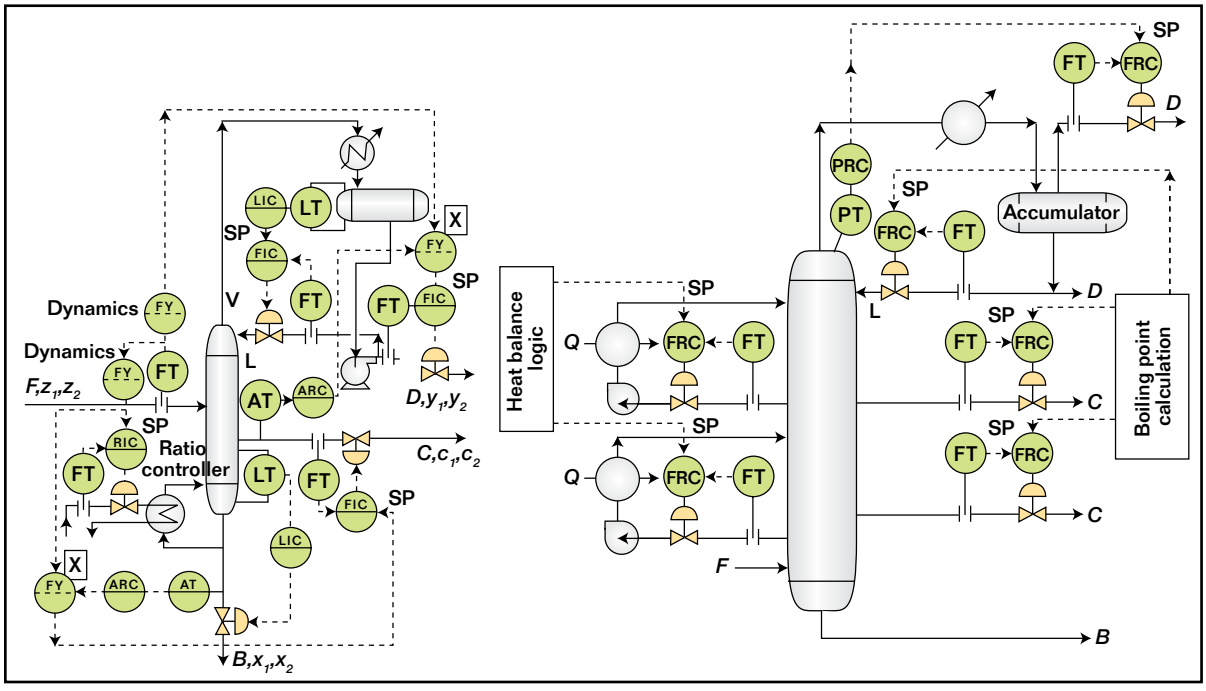
Super-fractionators are used to separate streams with close relative volatilities of the light and heavy key components. These include deisobutanizers, which separate isobutane from normal butane, propylene splitters, which separate propane from propylene, ethylbenzene towers, which separate ethylbenzene from xylene, xylene splitters, which separate para and ortho-xylene from meta-xylene.

Super-fractionators require many trays; therefore, the column heights often necessitate dividing them into two or even three sections. Super-fractionators require high internal vapor and liquid rates and high flow ratios of both reflux-to-distillate and vapor-to-bottoms. These result in large tower pressure drops, long time constants and long dead times. Therefore, the response of compositions to the manipulation of feed and reflux flows is slow. As a consequence, dynamically compensated material balance control is recommended to control the distillate compositions.

Multivariable Control

Traditional PID controls usually implement SISO (single input, single output) algorithms, while advanced controls work with multiple inputs and outputs (MIMO). The modeling of a process can be of the "white box" or "black box" type. White-box modeling is used for well-understood processes, such as distillation, where the knowledge of mass, energy and momentum balances allow the development of ac-

FIGURE 20.



Left: Multi-product fractionator controls where, after dynamic correction, the boil-up (V), side-draw (C) and distillate (D) flows are ratioed to the feed flow (F); Right: The true boiling points are controlled by throttling the product flows, while heat balance is controlled by manipulating the reflux flows.

curate dynamic models. These internal model control (IMC) systems are useful in optimizing the process and in anticipating future events.

"Black box" or model-free controls (MFC) include the ANN, fuzzy logic and statistical process control strategies. These algorithms are trained on the data obtained from the past operation of the controlled process. Their limitations include their required relatively long learning periods and the fact that their knowledge is based on the past. Therefore, they are not well-suited to anticipating future events, and if conditions change, they require retraining.

Model-Based Control

Once a process model has been established, it is possible to build the inverse of that model, which can be used as a controller. In that sense, the PID controller is a linear inverse model of a single process loop. A simple IMC is shown at the bottom of Figure 21 (p. 33). It has the same structure as a Smith predictor (Figure 9, Part B, p. 15), which is a first-order system with dead time combined with a PI controller.

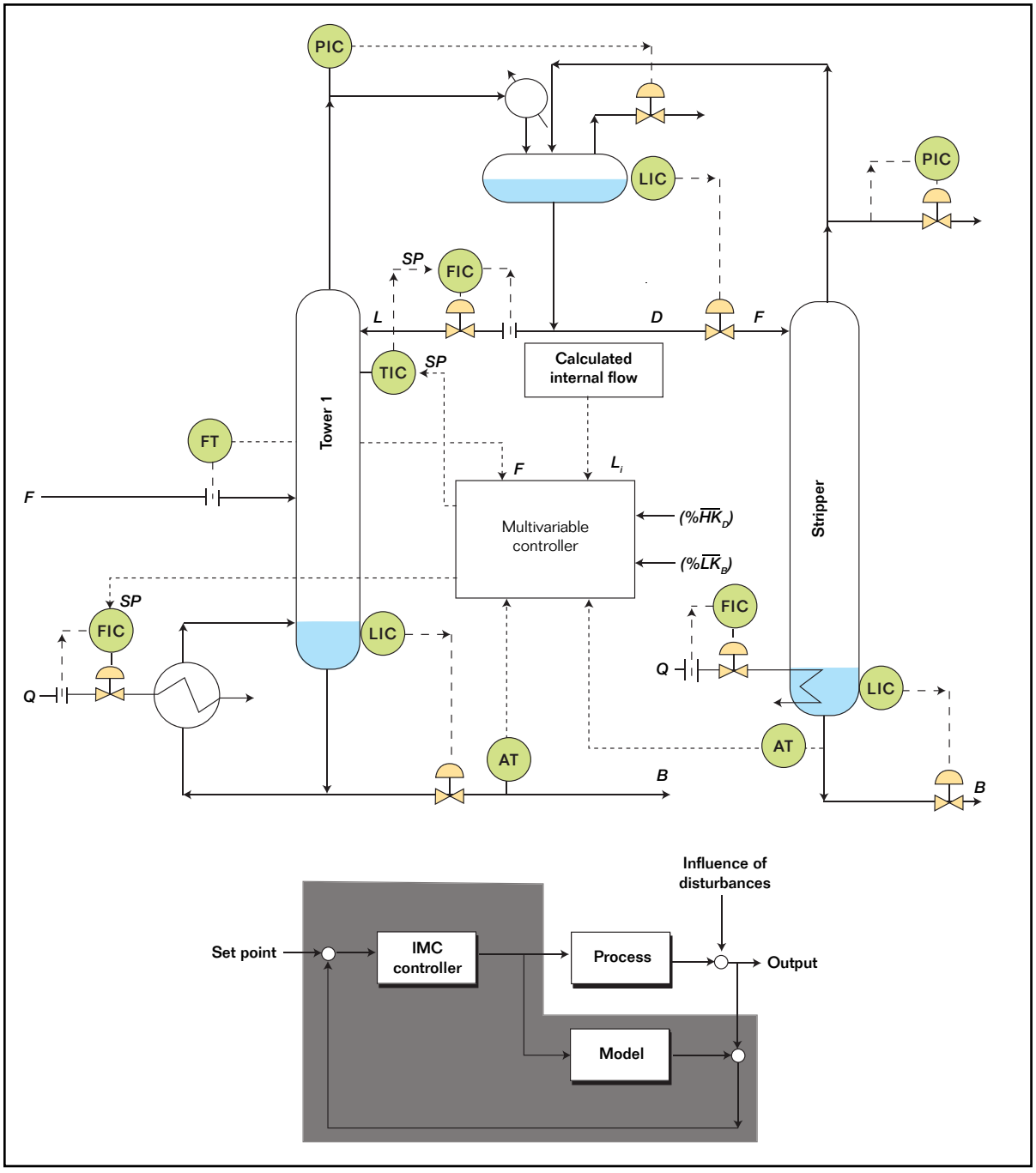
Multivariable control (MVC) is particularly well-suited for controlling highly interactive fractionators, where several control loops need to be simultaneously decoupled. Figure 21 illustrates an MVC configuration, which simultaneously considers all the process lags and applies safety constraints and economic optimization factors in determining the required manipulations to the process.

In Figure 21, two products and an impurity stream are produced by two towers. The objective is to control the composition of both products; therefore, when an adjustment is made to control one composition, that manipulation causes a disturbance in the operation of the other composition loop.

In this example, the MVC controls the two product compositions, while keeping the operation within the constraints of the process equipment capacities and considering the dead time introduced by the stripper. In this process, the feed-flow rate is the main disturbance variable. The steam to the first column and the temperature at the top of that column are the manipulated variables. A constraint variable is the internal reflux flow, which is calculated from tower temperatures and flows.

The technique of multivariable control requires the development of dynamic models based upon fractionator testing and data collection. Multivariable control applies the dynamic models and historical information to predict future fractionator characteristics. Predicted fractionator responses result in planned controller actions on the manipulated variables to minimize error for the dependent controlled variable. For towers that are subject to many constraints, towers that have severe interactions, and towers with complex configurations, multivariable control can be a valuable tool.

FIGURE 21.



Multivariable internal model controls (IMC) for controlling two product compositions (B), while keeping the operation within the constraints of the process equipment and while taking into account interactions and the dead time introduced by the stripper.

Dynamic Matrix Control (DMC) is also an MVC technique, but it uses a set of linear differential equations to describe the process. The DMC method obtains its data from process-step responses and calculates the required manipulations using an inverse model. Coefficients for the linear equations describing the process dynamics are determined by process testing. During these tests, manipulated and load variables are perturbed, and the dynamic responses of all controlled variables are observed. This identification procedure is time-consuming and requires substantial local expertise.

Artificial Neural Networks

Figure 22 (p. 35) shows a three-layer, back-propagation ANN, which predicts the manipulated steam and reflux flows of a column. The process model is stored in the ANN by the way its processing elements (nodes) are connected and by the importance that is assigned to each node (weight). The ANN is "trained" by example, and therefore, it contains the adaptive mechanism for learning from examples (somewhat similarly to the learning process of a child). During the "training" of these networks, the weights are adjusted until the output of the ANN matches that of the real process. Naturally, when process conditions change, the network requires retraining. The hidden layers help the network to generalize and even to memorize.

In the SISO configuration, the ANN network builds an internal nonlinear model relating the controlled and manipulated variables. It builds this model by learning or "training" from a data set of known measurements and process responses. This makes the neural controller more useful and more robust than the standard PID. Because the neural network paradigm can accommodate multiple inputs and outputs, an entire fractionator model can be built into a single controller. The neural controller can be thought of in the same terms as model-based control algorithms, whereby the neural network is used to obtain the inverse of the process model. As shown on the top right of Figure 22, the back-propagation network can be trained to behave as an inverse model of the process, with load and controlled variables being input vectors and manipulated variables output vectors.

To build such a model, all inputs and outputs must first be normalized based upon expected minimum and maximum values, and be presented to the network for the training

By using the same historical data, the network can be trained, and a nonlinear internal model can be created. The network's ability to do the prediction of the dynamics of the fractionator improves as more data become available for training. Thus, the neural controller can be considered as a specific type of nonlinear, multivariable, model-based control algorithm. Instead of creating the nonlinear process model with explicit equations, the neural controller builds its own process model from actual tower operation.

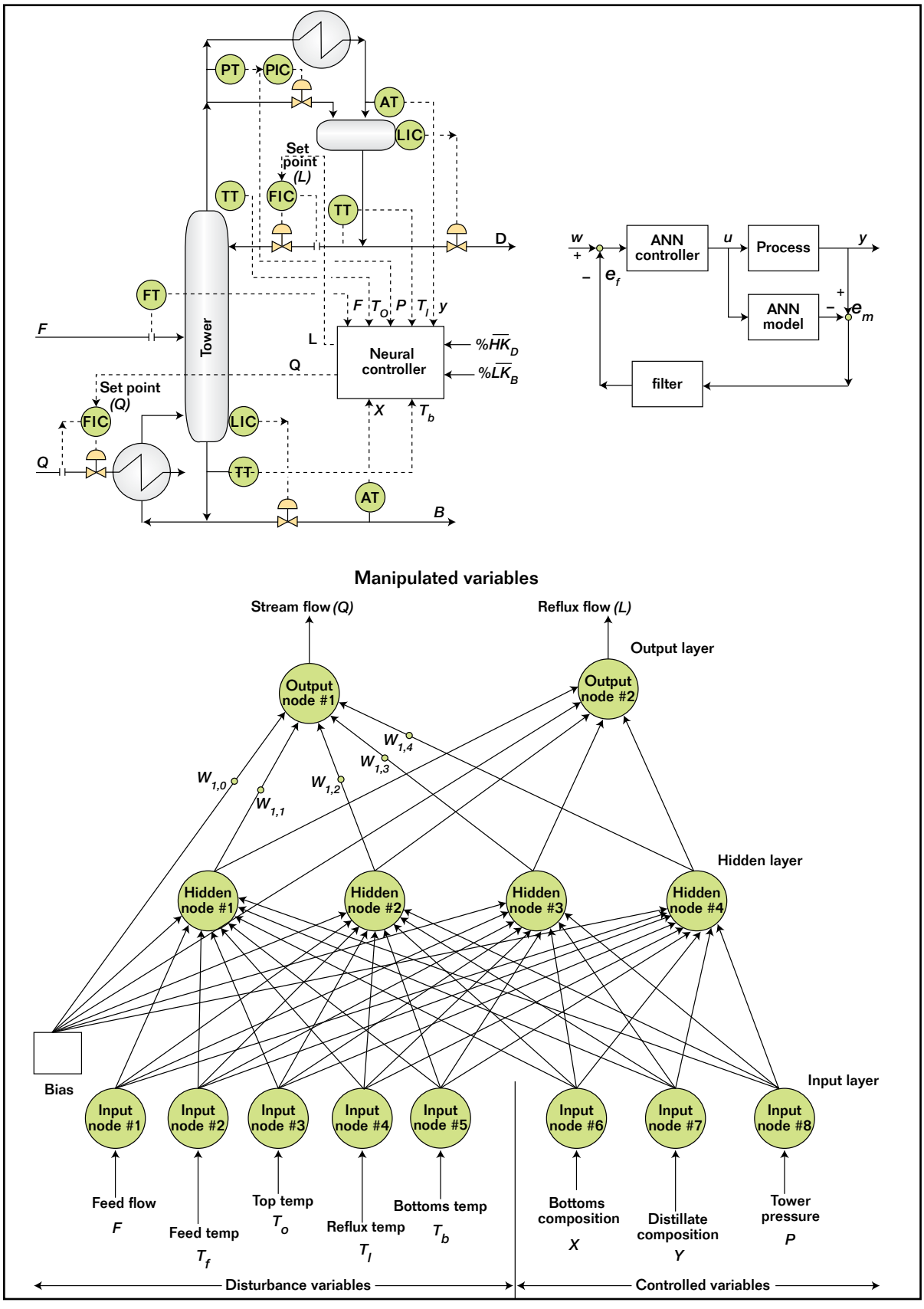
Since the neural controller is an empirical, not a theoretical model, it is susceptible to errors if operated outside the conditions of the training set. Data for the training set need to be continually gathered and the network retrained whenever novel conditions occur in order to increase the robustness of the neural controller throughout its operating life.

The Total Model

It is possible to design a control system which will compensate for changes in any of the load variables: feed rate, composition, enthalpy, reflux and bottoms enthalpy. The goal of these systems is to eliminate interactions and to protect the column from the consequences of changes in ambient conditions. To provide a model that describes both the material and energy balance of the column, one has to develop both the steady-state and the dynamic equations for the following:

- Feed enthalpy balance,
- Bottoms enthalpy balance,
- Internal reflux computation,
- Reboiler heat balance,
- Overall material balance.

FIGURE 22.



The configuration of a back-propagation neural network (ANN) and its use as an internal model controller.

Sub-Optimization of Unit Operations

Every distillation column is unique and therefore, the control strategies described in Figure 23 (p. 37) are for illustrative purposes only and should not be considered to be the sub-optimal or optimal control solutions for every column. The goal of optimization of a single column is to safely operate at maximum profit, but this can only be done if the market value of each product is known. This is not the case when the products of a column are not final products, but the feed flows to other unit processes.

When the product prices are unknown, it is still possible to perform optimization, but the optimization goal changes. The criteria in that case becomes the generation of the required products at minimum operating cost. This can be called an optimum with respect to the column involved, but only a "sub-optimum" with respect to the system of which the column is a part.

When the market values of the products are known, the column can be fully optimized, but additional variables, including the type of the market that exists for the products, must still be considered. If the market is limited, the goal is to generate the products at optimum separation and minimum operating cost. This cost varies as the feed flows and their compositions vary. When the market is unlimited and sufficient feedstock is available, the optimization task is more difficult, because one must determine both the optimum separation and the value of the feed streams. In this case, the goal of optimization is either maximum loading or maximum energy efficiency.

In this case, one of three constraints can be the limiting one:

1. Throughput can be limited by the maximum cooling capacity of the overhead vapor condenser,
2. The maximum heat input capacity of the reboiler,
3. The maximum separation rate of the column itself.

In some cases, this constraint will change from time to time, depending upon product prices and other independent variables. Therefore, the design of an optimal control system for a single column should follow three logical steps:

1. Designing the basic controls to regulate the operating variables, such as pressures, temperatures, levels and flows;
2. Configuring the controls to regulate the reboiler heat input, the internal reflux flow rate, feed enthalpy and the sources of heat to the reboiler and preheater(s);
3. Determining the controls required to maintain the specified separation.

If the above control loops are provided for a single column, that column can be said to have been "sub-optimized." This sub-optimum will generate products at close to the specified separation, whether or not that separation is ideal with regard to the total system. If product purities are higher than specified, the operation is not considered sub-optimum.

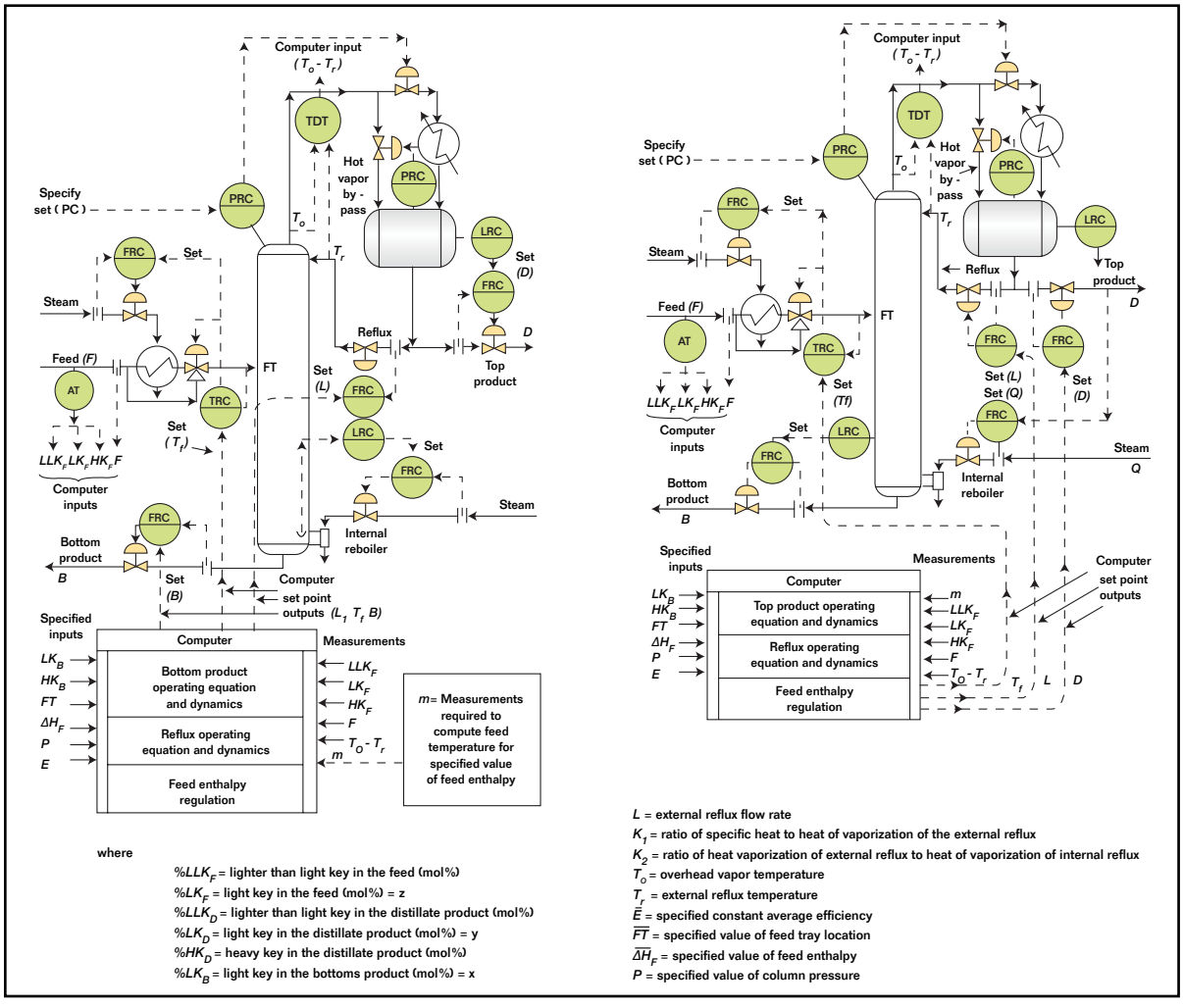
Operating Equations

When a single column is automated through the sub-optimum operation stage, it will still exhibit up to five degrees of freedom. The left side of Figure 23 describes the controls of a column that has been "sub-optimized" by controlling bottom product flow to obtain the specified separation. Figure 24 (p. 38) shows some typical operating control equations used in operating the predictive controls for maintaining the energy balance (reflux flow rate) and the material balance (bottom product flow rate) to give the specified separation (right of Figure 23). Both control systems shown in Figure 23 will achieve their goal of operating at sub-optimum; the difference is mainly in their basic controls.

The theoretical operating equations are normally developed by tray-to-tray runs of calculations, which are usually performed by an off-line digital computer. Next, a statistically designed set of runs is made, and the information thus obtained is curve-fitted to an assumed equation form.

Once the steady-state theoretical equation is developed and placed in service, the experimental part is determined by online tests. These tests involve operating the column at different loads to determine the correction required to (L_i/F) in order to obtain the specified separation. Average overall efficiency $\overline{E}$ is set to make (L_i/F) equal to the actual L_i/F which exists. The loading tests are carried out under this condition.

FIGURE 23.



Suboptimization distillation controls aimed at producing the specified separation by controlling bottom product flow (left) or reflux flow (right).

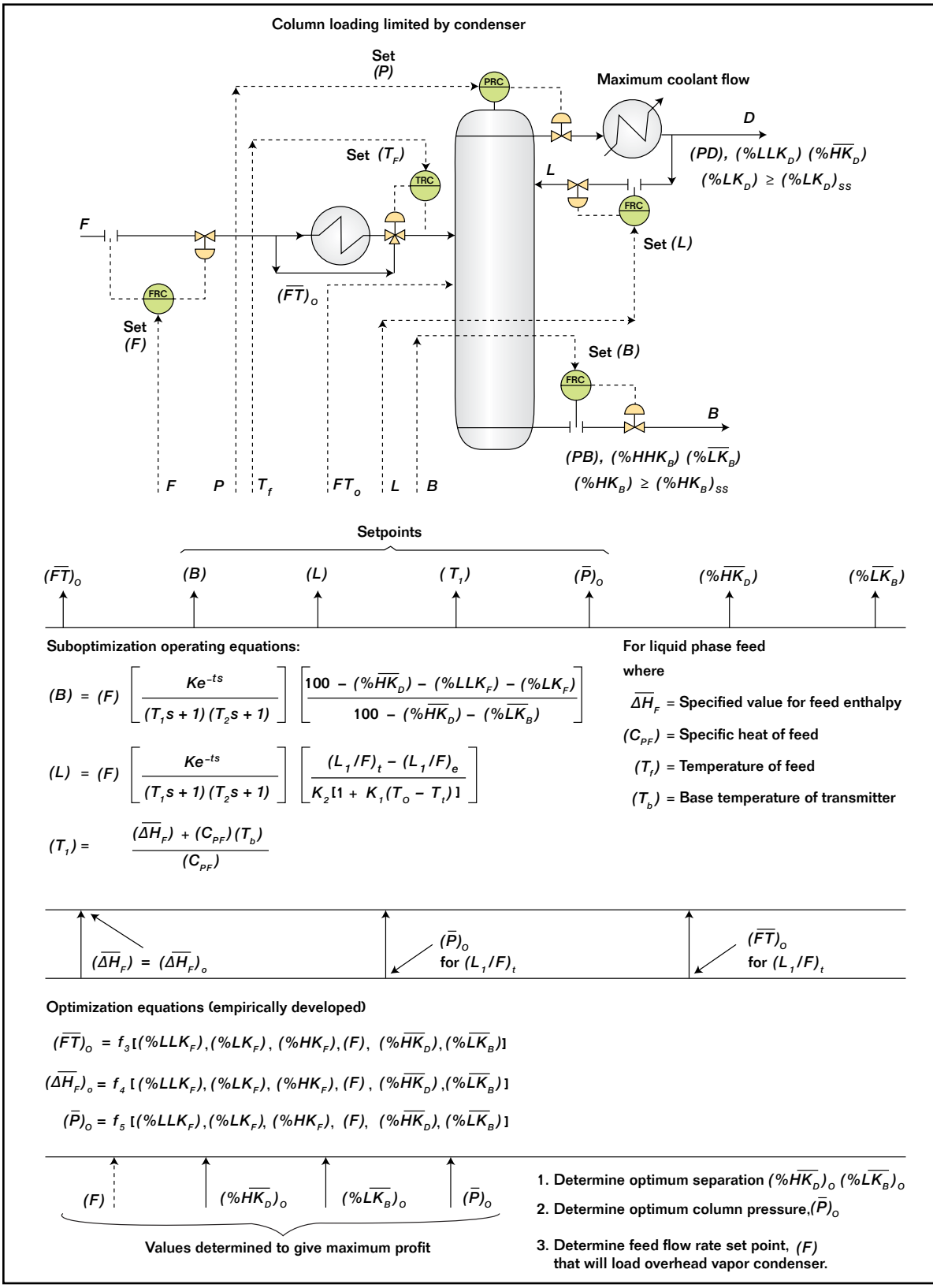
In other cases, plant tests can be performed to determine L_1/F without consideration of the theoretical term. As to the internal reflux flow rate, one can approximate it by making a heat balance around the top tray.

If the operating equations that have been obtained for the steady state are used without alteration, the column will be upset when changes, such as sudden feed flow rate increase or decrease occur. Feed composition changes cause less severe upsets and, therefore, providing dynamic compensation is less of a concern.

The purpose of dynamic compensation is to compensate for changes (such as feed-flow rate changes) in such a way that the column's terminal flows respond to them at the proper time, in the correct direction and without overshoot. The simplest dynamic element, which can be used is the sum of a dead time plus a second-order exponential lag response. As shown in Figures 18 (p. 29) and 23, the feed flow rate signal is passed through this dynamic element before being used to obtain the set points for the bottom product flow rate (B) and the reflux flow rate (L) set points.

The bottoms and overhead dynamic constants, t, T_1 and T_2 in the operating equations listed in Figure 24, are not necessarily the same values. With the use of the operating equations in Figure 24 (p. 38), the system still has five degrees of freedom. Therefore, feed tray location ($\overline{FT}$), feed enthalpy ($\overline{\Delta H}_F$), column pressure ($\overline{P}$), concentration of heavy key component in the top product ($\%\overline{HK}_D$), and concentration of light key component in the bottom product ($\%LK_B$) can all be controlled according to specifications. Although tray efficiency is also included in the operating equations, it is a fixed value.

FIGURE 24.



Top: The control configuration; Bottom: Operating equations to minimize operating costs when the market is unlimited and product prices (*PD & PB*) are known.

TOTAL OPTIMIZATION

Optimization implies maximum profit rate. An objective function is selected, and manipulated variables are chosen that will maximize or minimize that function. This is similar to the PID equation that is designed to minimize the error between the set point and measurement, but it views the process at a higher level. Optimization can be applied in several layers. Local optimization is the optimization of a single column. Normally, the goal of optimization of a single tower is to obtain minimum energy consumption or maximum throughput. For a detailed treatment of distillation optimization and of the development of the optimization equations, refer to Chapter 8.21, Vol. 2, *The Instrument Engineer's Handbook*, 4th edition.

Unit optimization addresses several columns in series or parallel. It is concerned with the effective allocation of feedstocks and energy among the members of that system. Plant-wide optimization involves coordinating the control of distillation units, furnaces, compressors, etc. to maximize profit from the entire operation. All lower-level control functions respond to set points received from higher-level optimizers.

Product-Price Considerations

If the prices of the distillation products are not known, it is impossible to maximize the profit rate for that column without taking into account all other aspects of the overall plant. Optimization of a single column whose product prices are unknown involves determining the location of the feed tray ($\overline{FT}$), the feed enthalpy ($\overline{\Delta H}_F$), and the column pressure ($\overline{P}$), which will result in minimum operating costs. Assuming that the appropriate operating equations are available, it is a relatively simple matter to establish optimum values for these three variables. Because these variables must stay within specific limits, a statistical design study can be made offline that allows correlation of the variables with each of the three optimizing variables.

In a majority of cases, the optimum column pressure ($\overline{P}$) can be found by lowering the operating pressure until the condenser capacity is reached, or until liquid entrainment occurs in the vapor on the trays. The operating equations in Figure 24 are applicable for the case when product prices are unknown.

When terminal product prices for a single column are known, the column is optimized to obtain the specified separation for the least operating cost. In this case, the main task is to find the separation that will maximize profit rate. Any of the following conditions can be true for a particular column.

1. The optimum separation can be determined independently of feed cost.
2. Optimum separation can be obtained by producing the product with the highest unit price at minimum specified purity.
3. The optimum separation is a function of the price difference between products.
4. The optimum separation is a function of the price difference between the heavy key components in the top and bottom ($PHK_D - PHK_B$) and of the price difference between the light key component in the top and in the bottom products ($PLK_D - PLK_B$).

The above four policies are derived by the evaluation of the partial differential equations that describe the profit rate for a single column with respect to the specified separation, noted as ($\overline{LK}_B$), ($\overline{HK}_D$).

Operating Constraints

When product prices are known, complete economic optimization almost always requires that the column be operated against a constraint or at a point where the specified separation, (LK_B), is such that an incremental gain in production is equal to a corresponding incremental gain in operating cost.

Column loading is increased when either the separation is improved or when the feed rate is increased at a constant separation. The operating constraints are a function of the capacities of the condenser, the reboiler and the column. As feed rate or separation is increased, one of these three constraints will be approached. Ambient conditions, steam pressure, feed composition, etc. can also influence the choice of the constraint first reached.

Condenser Constraint: The maximum cooling capacity of a given condenser at maximum coolant flow rate is a function of the difference between the temperatures of the overhead vapor and of the coolant. When operating against the condenser constraint, tests should be conducted to determine the maximum vapor flow rate as a function of this temperature difference. Such a correlation can be obtained by column testing. Column pressure and condenser fouling are also major variables that will affect overhead vapor temperature.

Reboiler Constraint: The maximum heating capacity of the reboiler at maximum heating media flow rate is a function of the temperature difference between the heating media and the liquid being reboiled. As with the condenser, tests can be conducted to correlate maximum vapor flow rate with temperature difference across the reboiler tubes. Column pressure and reboiler fouling are major variables that will affect this correlation. Generally, when using waste streams such as low-pressure steam that would alternatively be vented, it is optimal to operate against a reboiler constraint.

Column Constraint: Column capacity is a function of liquid and vapor flow rates and of column pressure. Often, capacity is limited by liquid entrainment of the vapor. Operation at low internal liquid flow rates allows higher vapor flow rates. Also, column capacity increases as the operating pressure rises. If capacity is limited by entrainment, then loading can be increased by raising the operating pressure. However, if column capacity is limited by the tray downcomers, internal liquid flow rate can be increased by lowering pressures. Therefore, to optimize the operation, the capacity-limiting parameter must be known.

An equation can be developed from test data to cover a limited range of liquid and vapor flow rates and pressure. This relationship can be linear and have a form of $V_{max} = a_1 + a_2 (L) + a_3 (P)$. This equation is useful in predicting the values of feed flow rate and of the separation that will cause maximum vapor flow rate to exist.

For a detailed description of optimization strategies for a variety of product prices and market conditions, refer to Chapter 8.21, Vol. 2, *The Instrument Engineer's Handbook*, 4th edition.

Conclusions

The advanced process control strategies that are most applicable to the optimization of the distillation process are usually based on *white-box* modeling, where the theoretical dynamic models are derived on the basis of the mass, energy and momentum balances of this well-understood process. *Fuzzy-logic* and *black-box* models are less often used, as they are more applicable when it is acceptable to use a complete mechanistic empirical model constructed solely from a priori knowledge.

The amount of energy used for distillation is approximately 8% of the total energy used in the industrial sector of the United States. Refineries spend 50% to 60% of their operating costs (i.e., excluding capital costs and depreciation) on energy, almost twice as much as does the better controlled and optimized chemical industry (30% to 40%). This difference shows the potential for savings through better control and optimization.

While the optimization techniques described in this book can improve refinery productivity and profitability by 25% compared to present operation, this goal will only be achieved if we stop using individual PID loops and treat distillation as a single and integrated unit operation. In multivariable unit operation control, the variables, such as flows, levels, pressures, etc. become only constraints, while the controlled and optimized variable is productivity and profitability.

The advances in process control have replaced or supplemented the single loop PID, but it is still true that we can only control a process if we fully understand it. As was explained, each distillation column has its own "personality" and one cannot control it without fully understanding it. Therefore, the importance of the role of process control engineers, who fully understand the controlled process did not diminish, but in fact increased in importance as our tools of control became more sophisticated.

ACRONYM REFERENCES

ANN – artificial neural networks
ARC – analyzer recording controller
B – bottoms product
CW – cooling water
D – distillate
DMC – dynamic matrix control
$\bar{E}$ – efficiency
FCCU – fluidized catalytic cracking unit
FRC – flow recording controller
FT – flow transmitter or feed tray
HK – heavy key
HHK heavier than heavy key
IMC – internal model control
L – reflux
L_i – internal reflux
LK – light key
LLK – lighter than light key
LNG – liquefied natural gas
LRC – level recording controller
MFC – model-free controls
MIMO – multiple input, multiple output
MVC – multivariable control
PB – price of bottom product
PCV – pressure control valve
PD – price of distillate
PI – pressure indicator
PIC – pressure indicator controller
PID – proportional-integral-derivative
PRC – pressure recording controller
RFC – reflux flow controller
RIC – ratio indicating controller
RG – relative gain
SISO – single input, single output
TBP – true boiling point
TRC – temperature recording controller
VPC – valve position controller
VPCV – valve position control valve
ΔH – enthalpy change
x – bottom product composition
y – distillate composition
z – feed composition

According to Lipták, developing effective control strategies and optimization of distillation columns can improve "productivity and profitability by 25%." Being able to test these control strategies and optimization techniques offline can provide additional significant savings in time and money. By using the MiMiC Distillation Modeling Package, the process or instrument engineer can:

- Test control strategies and optimizations on an accurate, dynamic plant model
- Optimize the operation of the distillation process
- Train operators offline in an environment identical to plant Human Machine Interface

The MiMiC Distillation Modeling Package is a powerful tool for control strategy testing, initial and ongoing tuning of control loops, and dynamic optimization of distillation systems. It provides dynamic and accurate models for distillation, providing time and cost savings to maximize new capital projects and operational excellence initiatives. It provides:

- Accurate dynamics for distillation simulation
- Easy-to-use configuration wizard tool
- Powerful data visualization tools
- Quick and easy integration with process automation systems through MiMiC simulation tags

For more information about the MiMiC Distillation Modeling Package, please contact:

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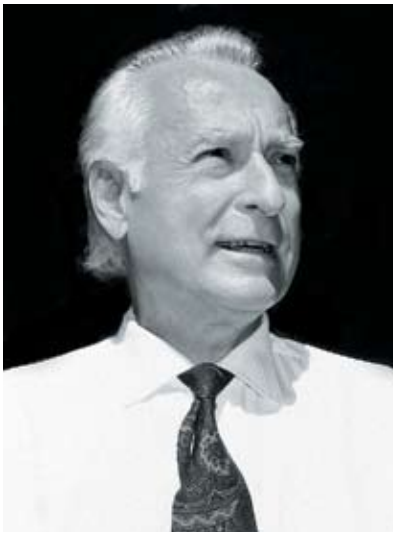
ABOUT MYNAH TECHNOLOGIES

MYNAH Technologies (www.mynah.com) is the leading provider of Process Simulation Software for Automation system software acceptance testing and operator training. Process companies worldwide save time and money on their process automation projects by using MYNAH's award-winning MiMiC Process Simulation Software. MYNAH also produces high-performance products for IO and device connectivity. MYNAH Technologies is a wholly-owned subsidiary of Experitec, Inc., located in Chesterfield, Missouri, USA.



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