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PRINCIPLES

A distillation column obtains separation through energy. Consequently, it seems intuitive that a product composition must be controlled by manipulating a term relating to energy. When the composition of both product streams from a two-product tower must be controlled, this suggests the following approach:

- Control the distillate composition by adjusting the reflux.
- Control the bottoms composition by adjusting the boilup.

For most columns, this control configuration exhibits a substantial degree of interaction, which translates to operational problems in the field.

An alternate approach is as follows:

- Control the composition of one of the products (distillate or bottoms) by adjusting an energy term (reflux or boilup).
- Control the composition of the other product by adjusting the respective product draw.

For most applications, the degree of interaction is much lower.

With this approach, one of the compositions is being controlled by directly adjusting a term in the column material balance. Consequently, this presentation begins with various material balances (entire tower, condenser only,

reboiler only). The discussion proceeds to component material balances for binary distillation, followed by an examination of the relationship between energy and separation. The primary objective is to provide insight into the nature of distillation and make the case that controlling one of the product compositions by adjusting a product draw is not only possible but is likely to be the appropriate approach for most towers.

This chapter reviews the general principles of distillation that are relevant to process control, including

- material balances, energy, and separation;
- composition control, through either energy terms or product flows;
- the stage-by-stage separation models for multicomponent distillation and their utility in control analyses;
- tray towers and packed towers;
- column dynamics.

1.1. SEPARATION PROCESSES

A simple separation process splits a feed stream into two product streams. In a pure separation process, no molecules are created, rearranged, or destroyed. That is, every molecule in the feed stream appears unchanged in one of the product streams.

Examples of industrial separation processes include the following:

- adsorbers
- centrifuges
- crystallizers
- cyclones
- decanters
- distillation columns
- dryers
- evaporators
- filters
- mist extractors

Every separation process relies on some principle to separate the molecules. Some separate by phases—a filter separates solids from liquids, a mist extractor separates liquids from gases, a decanter separates two immiscible liquids. Some separate by forcing a phase change—a dryer vaporizes a component such as water, leaving the nonvolatile solids behind. Distillation separates components based on their differences in volatility.

Separation processes, and distillation in particular, can become quite complex. Multiple feeds are possible. Multiple product streams are very common in distillation applications. Considerations such as energy conservation often add complexity to improve overall energy efficiency. Even reactive distillation systems are now occasionally incorporated into plant designs.

1.1.1. Binary Distillation

A binary separation process is one for which the feed contains only two components. Most presentations begin with such processes, as they are the simplest cases. Binary separations are occasionally encountered in practice, but most industrial columns are multicomponent.

A binary distillation example commonly used in textbooks is a column whose feed is a mixture of benzene and toluene. At atmospheric pressure, benzene boils at 80.1°C; toluene boils at 110.8°C. Consequently, benzene is more volatile than toluene. If a mixture of benzene and toluene is heated to its bubble point, the benzene vaporizes preferentially to the toluene. If the mixture is 50% benzene and 50% toluene, the vapor will contain more than 50% benzene and less than 50% toluene.

In distillation, the terms “light” and “heavy” are used to distinguish the components. But as used in distillation, these terms do not reflect weight, density, and so on. The light component is the more volatile; the heavy component is the less volatile. This notation is also reflected in the subscripts that designate the components:

- x_L = mole fraction of light component in a liquid stream or phase;
- x_H = mole fraction of heavy component in a liquid stream or phase;
- y_L = mole fraction of light component in a vapor stream or phase;
- y_H = mole fraction of heavy component in a vapor stream or phase.

1.1.2. Stages

A stage provides an arrangement where a vapor phase is in equilibrium with a liquid phase. The more volatile components concentrate in the vapor phase. The less volatile components concentrate in the liquid phase. The relationship between the vapor composition and the liquid composition is governed by the vapor–liquid equilibrium relationships for the various components.

A flash drum is a separation process that consists of a single stage. The feed is a superheated liquid that partially vaporizes (or flashes) within the flash drum. The two phases are separated to provide a vapor stream and a liquid stream. These are assumed to be in equilibrium as per the vapor–liquid equilibrium relationships.

Such single-stage separations are only viable when a crude separation is required between materials of significant difference in volatility. In distillation

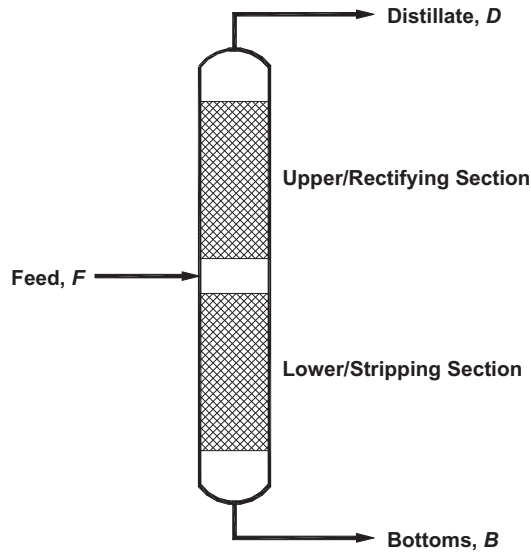


Figure 1.1. Distillation column.

columns, a separation section provides a sequence of stages whereby liquid flowing down the section is successively contacted with the vapor flowing up the section. One approach is to use trays to provide the vapor–liquid contact, with each tray ideally providing one stage (actual trays are not quite that good). The alternate approach is to use packing to provide the vapor–liquid contact. The selection of trays versus packing is a design issue with surprisingly little impact on the column controls.

As illustrated in Figure 1.1, a two-product tower contains two separation sections, one (the upper or rectifying section) between the feed and the distillate, and the other (the lower or stripping section) between the feed and the bottoms. The number of stages required in each section is determined by the design of the column. The controls have no way to influence the number of stages in each section.

Designs are usually based on “ideal stages,” where the vapor and liquid on the stage are in equilibrium. Actual stages rarely achieve this. A parameter known as the stage efficiency quantifies the departure of a stage from ideality. This parameter is used to adjust the actual number of stages installed in a column.

1.1.3. Engineering Units

For operator displays, reports, and other indications in production operations, the engineering units are typically as follows:

Flows. Either mass flow (kg/h, lb/h, etc.) or volumetric flow (L/h, gal/h, etc.).

Compositions. Either weight percent (wt%) or volume percent (vol%) for liquids; usually vol% (= mol%) for gases and vapors.

However, vapor–liquid equilibrium relationships are fundamentally based on molar quantities. Consequently, the equations used for the design, analysis, and so on, of distillation columns are normally developed in molar units:

Flows. Molar flow (mol/h, mol/min, etc.).

Compositions. Mole fractions.

Herein molar units will generally be used for both flows and compositions.

1.1.4. Feed and Product Streams

Figure 1.1 illustrates a two-product distillation column with a single feed stream. The designation of the streams is usually as follows:

Feed. The flow rate of this stream will be designated by F , in mol/h.

Distillate. The flow rate of this stream will be designated by D , in mol/h.

This stream is sometimes referred to as the overheads.

Bottoms. The flow rate of this stream will be designated by B , in mol/h.

Feed composition. The possibilities for the feed stream F are as follows:

- entirely liquid,
- entirely vapor,
- vapor–liquid mixture.

The mole fraction of such streams is normally designated by z . The composition of the light component is z_L ; the composition of the heavy component is z_H .

1.1.5. Distillate Composition

The possibilities for the distillate stream are as follows:

Entirely liquid. The condenser must be a total condenser as illustrated in Figure 1.2a. The overhead vapor V_C that flows into the condenser is totally condensed to provide liquid for the distillate stream and the reflux stream. The composition of the distillate is the same as the composition of the overhead vapor.

Entirely vapor. The condenser must be a partial condenser as illustrated in Figure 1.2b. Only part of the overhead vapor V_C flowing into the condenser is condensed. The resulting liquid is the reflux stream. The

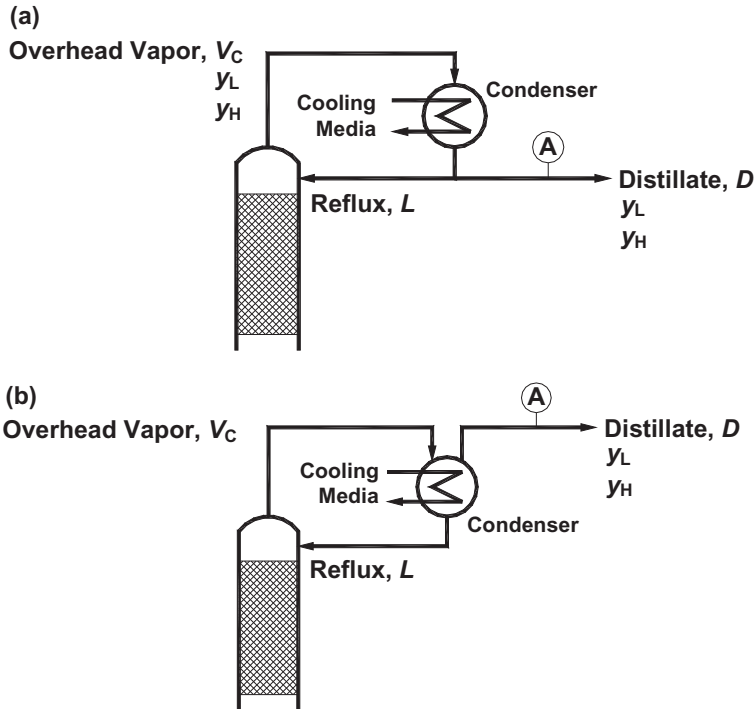


Figure 1.2. Overhead composition. (a) Total condenser. (b) Partial condenser.

distillate stream is the vapor that is not condensed. A partial condenser provides separation and is ideally one stage. The composition of the distillate is not the same as the composition of the overhead vapor.

The distillate composition is either the composition of a vapor stream (partial condenser) or the composition of a vapor stream that is condensed (total condenser) to provide the liquid overhead product. Vapor compositions are normally designated by y , giving the following notation for the distillate composition:

- y_L = mole fraction of the light component;
- y_H = mole fraction of the heavy component.

1.1.6. Bottoms Composition

As illustrated in Figure 1.3, the bottoms stream is always a liquid stream. Only part of the liquid flowing into the reboiler is vaporized, making the reboiler the counterpart of the partial condenser. The vapor stream becomes the boilup to the column; the liquid stream is the bottoms product.

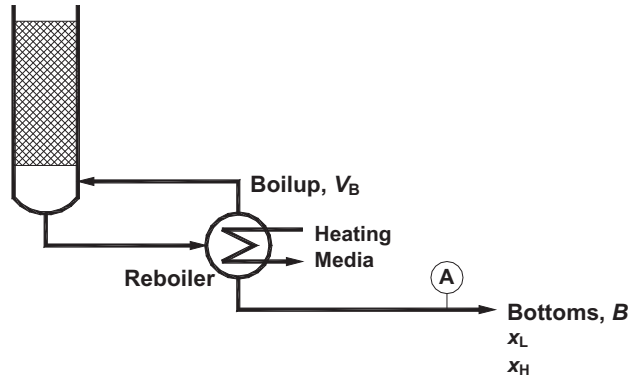


Figure 1.3. Bottoms composition.

Liquid compositions are normally designated by x , giving the following notation for the bottoms composition:

- x_L = mole fraction of the light component;
- x_H = mole fraction of the heavy component.

1.1.7. Composition Measurement

The performance of a column ultimately depends on the composition of the product streams. There are two possibilities:

- Single-end composition control.** The composition of one of the product streams is controlled, and the other is allowed to “float.”
- Double-end composition control.** The composition of both product streams is controlled. This is far more challenging.

The specification for the composition of a product stream can be in many forms, some of which will be examined in the next chapter. Throughout this book, the composition of a product stream will be stated in terms of one or more impurities. For a binary separation, the only impurity in the distillate composition is y_H ; the only impurity in the bottoms is x_L . The smaller the value of y_H , the higher the purity of the distillate product. The smaller the value of x_L , the higher the purity of the bottoms product.

Ideally, a product composition would be sensed by an onstream analyzer installed on the product stream, as is illustrated in Figures 1.2a,b and 1.3. This will be the general practice in the piping and instrumentation (P&I) diagrams presented in this book. But unfortunately, practical considerations often dictate otherwise, the options generally being the following:

Install an onstream analyzer on a nearby stream. As will be discussed in the next chapter, installing the analyzer directly on the product stream is often impractical, but the desire is to select a stream as near as possible to the product stream.

Use temperature in lieu of onstream analyzer. The incentive is obvious—cost. The stage on which the temperature is selected is called a *control stage*. The hope is that maintaining the appropriate temperature on the control stage will give a product of the desired composition. This must always be coupled with an off-line analysis that provides the basis for the process operators to adjust the target for the control stage temperature. The various issues will be explored in the next chapter.

Manual control based on off-line analyses. The operator makes adjustments based on the results of the off-line analyses. The downside of this approach is that the product compositions are conservatively maintained within specification, which results in reduced throughput, lower yields (loss of valuable product through a product stream), increased energy costs, and so on.

The P&I diagrams in this book will generally illustrate composition control based on a composition analyzer installed directly on a product stream. This is the ideal, and the closer it can be achieved in practice, the better.

1.1.8. Manipulated Variables

In distillation applications, the most common final control elements are control valves, although pumps with variable speed drives are certainly viable alternatives. Consequently, the output of most controllers will be a control valve opening. This valve opening in turn determines the flow through the control valve.

Technically, the manipulated variable would be the control valve opening. However, the various relationships (material balances, energy balances, etc.) that will be written for a column invariably involve flows, not valve openings. The variables in distillation simulation programs are always flows, never valve openings. Consequently, in this book, the flow through the control valve will be routinely referred to as the manipulated variable.

In older towers, flow measurements were rather sparingly installed. But in newer towers, flow measurements are more widely applied, and in some, a flow measurement is installed on every stream where metering is possible. The availability of a flow measurement permits a flow controller to be configured in the controls, and cascade control configured for loops such as composition and level. In cascade control, the output of the outer loop (composition, level, etc.) is the set point of the inner loop (flow). Technically, the manipulated variable for the outer loop is a flow set point, but as flow controllers are far faster than composition, level, and so on, the actual flow is

essentially equal to its set point, at least from the perspective of the slower loop. In the cascade configurations, the manipulated variable for the outer loop is essentially a flow.

As composition loops are very slow, providing a flow controller as an inner loop is generally recommended. In this book, cascade will be indicated for composition loops and for temperature loops for the upper and lower control stages. For level loops, providing a flow controller for the inner loop is not essential, especially when close control of level is not required. Within this book, cascade control will not generally be configured for level loops. However, if a flow measurement is available for other reasons, cascade control should be configured in practice.

1.2. TOTAL MATERIAL BALANCE

Material balances are the most fundamental equations that can be written for any process. For the two-product distillation column illustrated in Figure 1.4, the steady-state total material balance is written as follows:

$$F = D + B.$$

On a long-term basis, this equation must close. If the feed flow is constant, then

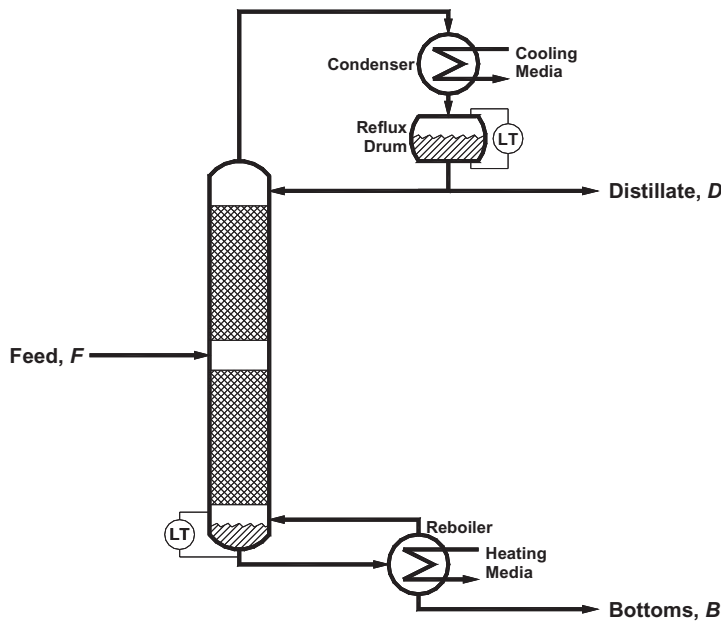


Figure 1.4. Holdups in a column.

1. any long-term change in the distillate flow must be offset by an equal and opposite change in the bottoms flow;
2. any long-term change in the bottoms flow must be offset by an equal and opposite change in the distillate flow.

1.2.1. Degrees of Freedom

The control configuration must be consistent with the degrees of freedom for the process. The equation for the degrees of freedom is as follows:

$$\text{Degrees of freedom} = \text{number of variables} - \text{number of equations.}$$

Most distillation columns are said to operate in a “fixed service,” which means that

1. the feed flow F is explicitly specified or is determined by upstream unit operations;
2. the feed composition is determined by upstream unit operations.

In such columns, the feed flow F is considered to be a known quantity in the material balance equation. This leaves two variables in the material balance equation, specifically, the distillate flow D and the bottoms flow B . Therefore, there are

- two variables (D and B);
- one equation (the total material balance equation);
- one degree of freedom.

1.2.2. Consequences for Control

The significance of this to the controls is as follows. A target for either the distillate flow or the bottoms flow can be independently specified, but not both. If either

1. the process operator specifies the target for the distillate flow or
2. a product composition controller specifies the target for the distillate flow,

then the bottoms flow must be the difference between the feed flow and the distillate flow. If either

1. the process operator specifies the target for the bottoms flow or
2. a product composition controller specifies the target for the bottoms flow,

then the distillate flow must be the difference between the feed flow and the bottoms flow.

1.2.3. Unsteady-State Behavior

At unsteady state, the possibilities are as follows:

1. Feed rate exceeds the sum of the product rates. Material accumulates somewhere within the tower.
2. Feed rate is less than the sum of the product rates. Material depletes somewhere within the tower.

Material accumulates or depletes primarily either in the reflux drum, in the bottom of the column, or both.

The amount of material (holdup) on the tower internals (trays or packing) is not constant. However, this holdup is largely determined by the design of the internals. The internal flows (reflux and boilup) have some influence on this holdup. However, the product flows (distillate and bottoms) have no direct influence on this holdup. Any long-term imbalance in the steady-state material balance will affect the holdup in the reflux drum and/or in the bottoms of the tower.

1.2.4. Level Measurement

As illustrated in Figure 1.4, level measurements are normally provided on both holdups. The capacity of these holdups is limited by the size of the equipment, so high and low level switches are usually installed in the reflux drum and in the bottoms. So that these switches are not actuated, one responsibility of the control configuration is to force the closure of the overall material balance by maintaining the levels within a “reasonable proximity” of their targets.

A level measurement for the bottoms holdup is essentially universal, but for the condenser, there are exceptions:

Flooded condenser. The condenser is partially filled with liquid, which reduces the effective area for condensing the overhead vapor. The level within the condenser is allowed to seek its own equilibrium, which means that sufficient heat transfer area is exposed to condense the overhead vapor. The level is never controlled and usually not measured.

No reflux drum. In small-diameter towers that require an external structure for support, the condenser is often physically mounted on the top of the tower. The reflux is returned directly to the tower, so no reflux drum is required.

These will be discussed in more detail in the subsequent chapter devoted to condenser arrangements.

1.2.5. Integrating Process

Consider the behavior of the process under the following conditions:

1. Process is within its design limits (no vessel capacities exceeded; no vessel empty).
2. No controls are on automatic.

Let H be the total holdup of material within the column. Changes in holdup affect the head for fluid flow. This is significant only for gravity flow applications, which are rare in distillation. Otherwise, changes in the holdup H have no direct effect on either the feed flow F , the distillate flow D , or the bottoms flow B .

The unsteady-state material balance can be written in either its differential or its integrated form:

$$\text{Differential: } \frac{dH(t)}{dt} = F(t) - D(t) - B(t)$$

$$\text{Integrated: } H(t) = \int [F(t) - D(t) - B(t)] dt$$

When H has no effect on F , D , or B , a process described by such equations is referred to as an *integrating process*. An alternate term is *ramp process* (the response to any upset is a ramp in the holdup or level) or *non-self-regulated process* (the process will not seek an equilibrium unless control actions are taken).

1.2.6. Level Control

An integrating process does not seek its own equilibrium. If there is an imbalance in the total material balance, the result is one of the following:

$F > B + D$. The holdup increases until some limiting condition is attained, the limiting condition being either

1. the level in the reflux drum actuates the high level switch or
2. the level in the bottoms actuates the high level switch.

$F < B + D$. The holdup decreases until some limiting condition is attained, the limiting condition being either

1. the level in the reflux drum actuates the low level switch or
2. the level in the bottoms actuates the low level switch.

The responsibility of every level controller is to close some material balance. To assure that the column material balance closes, every column control configuration must contain one of the following:

1. The reflux drum level is controlled by manipulating the distillate flow.
2. The bottoms level is controlled by manipulating the bottoms flow.

Providing both is also an option.

1.3. REFLUX AND BOILUP RATIOS

The reflux L and boilup V are associated with energy. The heat supplied to the reboiler generates the boilup V . In a partial condenser (distillate product is a vapor stream), the heat removed by the condenser generates the reflux L . In this context, several ratios arise, most of which involve the ratio of a liquid flow and a vapor flow.

1.3.1. External Reflux Ratio

The external reflux ratio is the ratio of the reflux flow L to the distillate flow D :

$$\text{External reflux ratio} = \frac{L}{D}.$$

In many towers, flow measurements can be installed for these two flows, and if so, the external reflux ratio can be computed.

However, there are tower designs where measurement of the reflux flow is not possible. To minimize pressure drops in vacuum towers, the condenser is often physically mounted on the top of the column. For a partial condenser, all of the condensate is returned directly to the column to provide the reflux. For a total condenser, part of the condensate is withdrawn with the remainder returned directly to the column to provide the reflux. In neither arrangement is it possible to measure the reflux flow.

1.3.2. Boilup Ratio

The counterpart to the external reflux ratio (that pertains to the top of the tower) is the boilup ratio (which pertains to the bottom of the tower). The boilup ratio is the ratio of the boilup V to the bottoms flow B :

$$\text{Boilup ratio} = \frac{V}{B}.$$

Direct measurement of the boilup flow V is never possible. Therefore, the boilup ratio cannot be computed from direct flow measurements.

When sufficient measurements are available to compute the energy transferred from the heating media to the reboiler, the boilup can be estimated by dividing this heat transfer rate by the latent heat of vaporization of the

material in the reboiler. The simplest case is a steam-heated reboiler with a measurement for the steam flow S . The boilup V can be computed as follows:

$$V \equiv \frac{S \cdot \lambda_S}{\lambda_B},$$

where

λ_B = latent heat of vaporization of liquid in the reboiler;

λ_S = latent heat of vaporization of the steam.

Unfortunately, there is always some error in the resulting value.

If the objective is to maintain a constant boilup flow, one possibility is to measure the pressure drop across a few of the lower stages and adjust the heat to the reboiler to maintain a constant pressure drop. One must use enough stages so that the pressure drop being sensed is above the noise invariably associated with such measurements. Furthermore, the pressure drop is related to the square of the vapor flow, so this approach works better at high vapor flows than at low vapor flows.

1.3.3. Internal Reflux Ratio

The internal reflux ratio R_I is the ratio of the reflux flow L to the vapor flow V at a point within the tower:

$$R_{I,k} = \frac{L_k}{V_k},$$

where

L_k = reflux flow at location k within the tower;

V_k = vapor flow at location k within the tower;

$R_{I,k}$ = internal reflux ratio at location k within the tower.

The vapor and liquid flows within most columns vary from stage to stage, so the internal reflux ratio is not constant. Furthermore, the internal reflux ratio above the feed stage will be different from the internal reflux ratio below the feed stage.

1.3.4. Above Feed Stage

For a location above the feed stage, Figure 1.5 presents the streams for a total material balance from that location through the top of the column. The total material balance is as follows:

$$V_k - L_k = D.$$

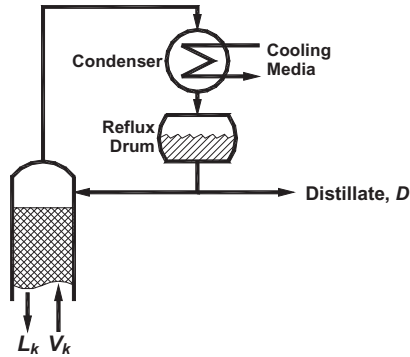


Figure 1.5. Internal reflux ratio above the feed stage.

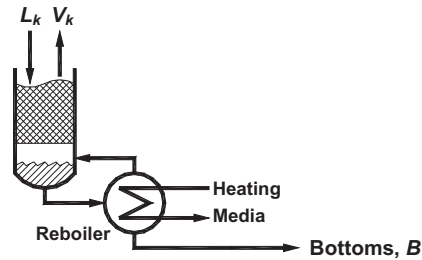


Figure 1.6. Internal reflux ratio below the feed stage.

Since the distillate flow D cannot be negative, the following conclusions can be made for the flows above the feed stage:

$$V_k \geq L_k,$$

$$R_{I,k} \leq 1.$$

1.3.5. Below Feed Stage

For a location below the feed stage, Figure 1.6 presents the streams for a total material balance from that location through the bottom of the column. The total material balance is as follows:

$$L_k - V_k = B.$$

Since the bottoms flow B cannot be negative, the following conclusions can be made for the flows below the feed stage:

$$L_k \geq V_k,$$

$$R_{I,k} \geq 1.$$

1.3.6. At Feed Stage

If one proceeds from the stages below the feed stage to stages above the feed stage, there is an abrupt change in the liquid flow relative to the vapor flow at the feed stage. Below the feed stage, the liquid flow exceeds the vapor flow. Above the feed stage, the vapor flow exceeds the liquid flow.

What happens at the feed stage depends on the enthalpy of the feed relative to conditions on the feed stage. There are five possibilities:

Feed is subcooled. All of the feed is added to the liquid flowing below the feed stage. In addition, some vapor is condensed at the feed stage to heat the feed to column temperatures. The condensed vapor is added to the liquid flowing below the feed stage, but is removed from the vapor flowing above the feed stage.

Feed is at its bubble point. All of the feed is added to the liquid flowing below the feed stage. No vapor is condensed at the feed stage.

Feed is between its bubble point and its dew point. Some feed flashes and is added to the vapor flowing above the feed stage. The remaining feed is added to the liquid flowing below the feed stage.

Feed is at its dew point. All of the feed is added to the vapor flowing above the feed stage. No liquid is vaporized on the feed stage.

Feed is superheated. All of the feed is added to the vapor flowing above the feed stage. Some liquid is vaporized to cool the feed to column temperatures. The vaporized liquid is added to the vapor flowing above the feed stage, but is removed from the liquid flowing below the feed stage.

Most process designs avoid highly subcooled feeds and highly superheated vapors.

1.3.7. Total Reflux

Most towers can be operated with the feed shut off and both product draws shut off. Sometimes this is during startup; sometimes this is during a temporary interruption in production operations.

If no distillate product is being withdrawn, all of the overhead vapor is condensed and returned to the column as reflux. The external reflux ratio is infinite, but the internal reflux ratio above the feed stage is exactly 1.0.

If no bottoms product is being withdrawn, all of the bottoms liquid is vaporized and returned to the column as boilup. The boilup ratio is infinite, but the internal reflux ratio below the feed stage is exactly 1.0.

At least theoretically, columns can operate indefinitely at total reflux. But in practice, total reflux is a temporary situation, although temporary could be hours or perhaps days. Energy is being consumed, but no product is

being made—not a good mode of operation with regards to the profit and loss statement. Production personnel must weigh the costs of continuing operation at total reflux versus the cost of shutting the tower down and re-starting it.

1.3.8. Equimolal Overflow

On every stage within a separation section, some vapor is condensed and some liquid is vaporized. Equimolal overflow means that for each mole of vapor that is condensed, exactly one mole of liquid is vaporized. This is definitely not assured. Separations involving light hydrocarbons (ethane, propane, etc.) deviate less than separations involving more complex components.

When equimolal overflow is assumed, the liquid and vapor flows within a separation section do not change from stage to stage. The liquid flow on all stages within the upper separation section is the reflux L . The vapor flow on all stages within the lower separation section is the boilup V .

At the feed stage, there will be a change in the liquid and/or vapor flows. One way to characterize the enthalpy of the feed is by its quality q , which is the fraction of the feed that vaporizes at the feed stage. The value of q for various types of feed is as follows:

$q < 0$. Subcooled feed; some vapor is condensed at the feed stage to heat the feed to column temperatures.

$q = 0$. Liquid feed at its bubble point; none of the feed is vaporized.

$0 < q < 1$. Partially vaporized feed.

$q = 1$. Vapor feed at its dew point; none of the feed is condensed.

$q > 1$. Feed is a superheated vapor; some liquid is vaporized at the feed stage to cool the feed to column temperatures.

When equimolal overflow is assumed, the liquid flow L_B in the lower separation section is computed as follows:

$$L_B = L + (1 - q)F.$$

The vapor flow throughout the upper separation section is the same as the overhead vapor flow V_C into the condenser and is computed as follows:

$$V_C = V + qF.$$

The assumption of equimolal overflow permits the liquid and vapor flows throughout the column to be easily computed. However, the results are approximate. For some separations, the liquid and vapor flows within a separation section change by a factor of 2 or more.

1.4. TOTAL MATERIAL BALANCE AROUND CONDENSER

A subsequent chapter is devoted to the wide variety of possible condenser configurations. A mechanism to influence the heat removed in the condenser is required, but the exact nature of this mechanism has no effect on the discussion that follows. The illustrations will only show a generic “cooling media” for a total condenser, but the discussion herein also applies to a partial condenser.

For small-diameter towers that require a structure for support, the condenser and reflux drum are usually physically located at the top of the column. But for a tower whose diameter is large enough that a structure is not required for support, cost issues favor the following configuration:

- The overhead vapor line extends to grade level.
- The condenser and reflux drum are physically at grade level.
- A reflux pump is required to return the reflux to the top stage.

No control issues are associated with any of this, so this detail will not be included in any of the illustrations in this book.

1.4.1. Condenser Material Balance

In the context of the material balance, the term “condenser” also includes the reflux drum, if one is present. The material balance contains a term for each of the three streams illustrated in Figure 1.7:

Distillate D (an output term). This is one of the product streams from the column. The controls influence the distillate flow via a control valve on the distillate stream.

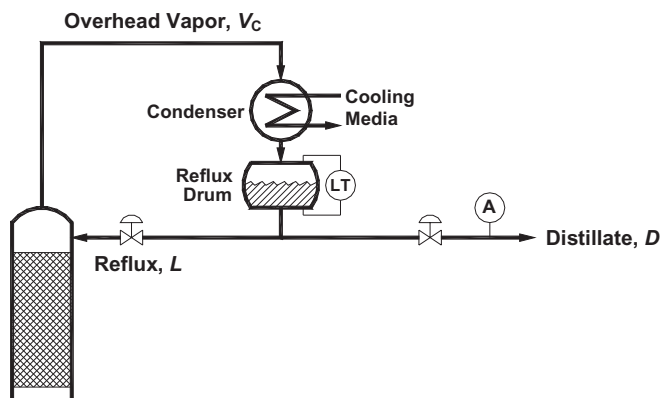


Figure 1.7. Material balance streams for condenser/reflux drum.

Reflux L (an output term). Part of the overhead vapor must be returned to the column as a liquid stream known as reflux. In most columns, the controls influence the reflux flow via a control valve on the reflux stream.

Overhead vapor V_C (an input term). This is determined by the heat removed in the condenser, which for most total condensers is adjusted by the tower pressure controller to maintain constant tower pressure. The material balance controls at the top of the column have no way to influence the overhead vapor flow.

The unsteady-state material balance around the condenser is written as follows:

$$V_C - (D + L) = \frac{dH_C}{dt},$$

where H_C is the reflux drum holdup (mole).

1.4.2. Control Configurations

The two manipulated variables, the distillate flow D and the reflux flow L , associated with the condenser are used to control the following two variables:

Distillate composition. When the distillate product is a salable product, good distillate composition control is crucial.

Reflux drum level. Rarely does the drum level affect any term in the profit-and-loss statement.

In selecting the control configuration, controlling the distillate composition must take priority, as reflected in the following approach:

1. Determine if the distillate composition is to be controlled by manipulating the reflux flow L or by manipulating the distillate flow D . This takes precedence over the usual preference to control level by manipulating the larger of the two flows (D or L).
2. Control reflux drum level with the other flow. However, level cannot be controlled by manipulating a very small flow. If $L/D \ll 1$, drum level cannot be controlled by manipulating L . If $L/D \gg 1$, drum level cannot be controlled by manipulating D .

Figure 1.8 presents the two possible control configurations, which are designated *direct material balance control* and *indirect material balance control*. The distillate flow D appears explicitly in the total material balance for the column:

$$F = D + B.$$

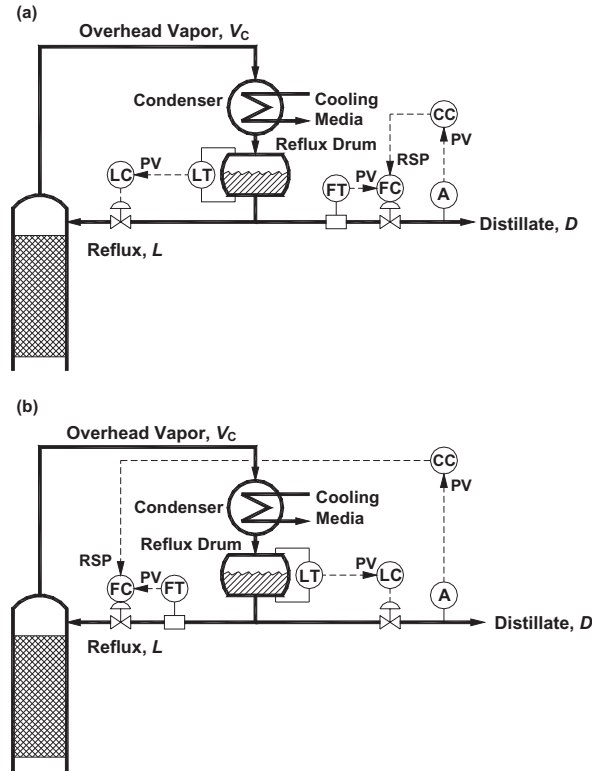


Figure 1.8. Control configurations for distillate composition. (a) Direct material balance control. (b) Indirect material balance control.

TABLE 1.1. Control Configurations for Distillate Composition

	Direct Material Balance Control	Indirect Material Balance Control
Control configuration	Figure 1.8a	Figure 1.8b
Manipulated variable for composition	Distillate D	Reflux L
Manipulated variable for drum level	Reflux L	Distillate D
Solution of condenser material balance	$L = V_c - D$	$D = V_c - L$
Preferred for level control if	$L > D$	$D > L$
Impractical if	$L/D \ll 1$	$L/D \gg 1$

The terms direct material balance control and indirect material balance control pertain to how the value of the distillate flow is determined. Table 1.1 summarizes the attributes of the two configurations.

The configuration in Figure 1.8a is the direct material balance control configuration. Values for D and L are determined as follows:



- D —specified by the distillate composition controller;
- L —determined by the level controller to satisfy the steady-state material balance for the condenser:

$$L = V_C - D.$$

The manipulated variable D for the composition controller appears explicitly in the column material balance.

The configuration in Figure 1.8b is the indirect material balance control configuration. Values for D and L are determined as follows:

- L —specified by the distillate composition controller;
- D —determined by the level controller to satisfy the steady-state material balance for the condenser:

$$D = V_C - L.$$

The manipulated variable L for the composition controller does not appear explicitly in the column material balance. Instead, the composition controller specifies L , from which the level controller determines the value of D .



1.5. TOTAL MATERIAL BALANCE AROUND REBOILER

A subsequent chapter is devoted to the wide variety of possible arrangements for reboilers at the bottom of the column. A mechanism to influence the heat added in the reboiler is required, but the exact nature of this mechanism has no effect on the discussion that follows. The illustrations will be for a steam-heated reboiler with a control valve and possibly a flow controller on the steam supply.

1.5.1. Reboiler Material Balance

In the context of the material balance, the term “reboiler” also includes the bottoms holdup. In Figure 1.9, the holdup for bottoms liquid is within the tower itself, but for kettle reboilers, this is within the reboiler. The material balance contains a term for each of the three streams illustrated in Figure 1.9:

Bottoms B (an output term). This is one of the product streams from the column. The controls influence the bottoms flow via a control valve on the bottoms stream.

Boilup V (an output term). Part of the liquid leaving the lower separation section of the column must be returned to the column as a vapor stream



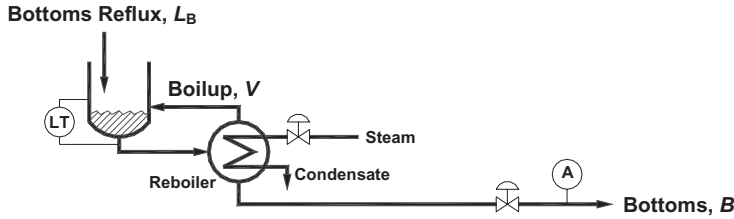


Figure 1.9. Material balance streams for reboiler.

known as boilup. Installing a control valve (or any other final control element) on the vapor stream leaving the reboiler is impractical. Instead, the controls must influence the boilup via the heat input to the reboiler. In Figure 1.9, the heat is supplied by steam, and a control valve is provided on the steam supply.

Bottoms reflux L_B (an input term). This is the liquid flow leaving the lower separation section within the column. The controls at the bottom of the column have no way to influence the bottoms liquid L_B .

The unsteady-state material balance around the reboiler is written as follows:

$$L_B - (B + V) = \frac{dH_B}{dt},$$

where H_B is the bottoms holdup (mole).

1.5.2. Control Configurations

The two manipulated variables, the bottoms flow B and the boilup V , associated with the reboiler are used to control the following two variables:

Bottoms composition. When the bottoms product is a salable product, good bottoms composition control is crucial.

Bottoms level. Rarely does the bottoms level affect any term in the profit-and-loss statement.

In selecting the control configuration, controlling the bottoms composition must take priority, as reflected in the following approach:

1. Determine if the bottoms composition is to be controlled by manipulating the boilup V or by manipulating the bottoms flow B . This takes precedence over the usual preference to control level by manipulating the larger of the two flows (B or V).

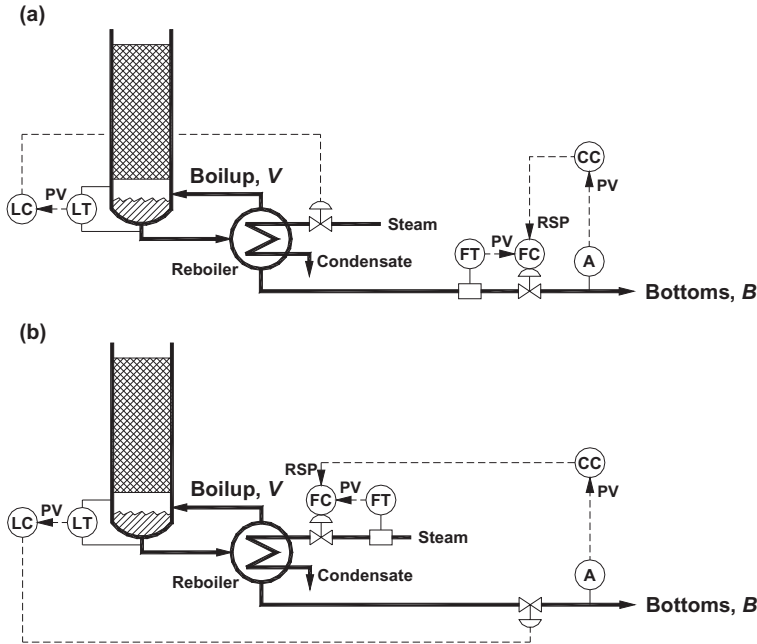


Figure 1.10. Control configurations for bottoms composition. (a) Direct material balance control. (b) Indirect material balance control.

- Control bottoms level with the other flow. However, level cannot be controlled by manipulating a very small flow. If $V/B \ll 1$, drum level cannot be controlled by manipulating V . If $V/B \gg 1$, drum level cannot be controlled by manipulating B .

Figure 1.10 presents the two possible control configurations, which are designated *direct material balance control* and *indirect material balance control*. The bottoms flow B appears explicitly in the total material balance for the column:

$$F = D + B.$$

The terms direct material balance control and indirect material balance control pertain to how the value of the bottoms flow is obtained. Table 1.2 summarizes the attributes of the two configurations.

The configuration in Figure 1.10a is the direct material balance control configuration. Values for B and V are determined as follows:

- B —specified by the bottoms composition controller;
- V —determined by the level controller to satisfy the steady-state material balance for the reboiler:

TABLE 1.2. Control Configurations for Bottoms Composition

	Direct Material Balance Control	Indirect Material Balance Control
Control configuration	Figure 1.10a	Figure 1.10b
Manipulated variable for composition	Bottoms B	Boilup V
Manipulated variable for bottoms level	Boilup V	Bottoms B
Solution of reboiler material balance	$V = L_B - B$	$B = L_B - V$
Preferred for level control if	$V > B$	$B > V$
Impractical if	$V/B \ll 1$	$V/B \gg 1$

$$V = L_B - B.$$

The manipulated variable B for the composition controller appears explicitly in the column material balance.

The configuration in Figure 1.10b is the indirect material balance control configuration. Values for B and V are determined as follows:

V —specified by the bottoms composition controller;

B —determined by the level controller to satisfy the steady-state material balance for the reboiler:

$$B = L_B - V.$$

The manipulated variable V for the composition controller does not appear explicitly in the column material balance. Instead, the composition controller specifies V , from which the level controller determines the value of B .

1.6. COMPONENT MATERIAL BALANCES

Herein component material balances will only be developed for the entire column. Component material balances can be made for the condenser and the reboiler, but these seem to have no significant implications for control.

1.6.1. Steady-State Equations

A component material balance can be written for each component in the feed. For binary distillation, there are two components (light and heavy), hence two equations:

$$\text{Light component: } F z_L = D y_L + B x_L,$$

$$\text{Heavy component: } F z_H = D y_H + B x_H.$$

The respective mole fractions must sum to unity:

$$x_L + x_H = 1,$$

$$y_L + y_H = 1,$$

$$z_L + z_H = 1.$$

Summing the above two component material balance equations gives the total material balance:

$$F(z_L + z_H) = D(y_L + y_H) + B(x_L + x_H),$$

$$F = D + B.$$

To obtain a set of independent equations, the total material balance can be used in lieu of either of the component material balances.

1.6.2. Degrees of Freedom

The analysis will be based on the following two independent equations:

$$\text{Total material balance : } F = D + B,$$

$$\text{Component material balance, light } F z_L = D y_L + B x_L.$$

A fixed service is assumed, which means that the feed flow F and the feed composition z_L are otherwise specified. The degrees of freedom are as follows:

Number of variables: 4 (B , D , y_L , and x_L)

Number of equations: 2

Degrees of freedom: $4 - 2 = 2$

For control, this means that independent targets can be provided for two of the four variables (B , D , y_L , and x_L). However, this does not mean “any two.”

1.6.3. Control Options

For a total of four variables, there are six possible subsets of two. But for the distillation column, it is possible to provide independent targets for only five of the six possible subsets:

Subset 1: D and y_L

Subset 2: D and x_L

Subset 3: B and y_L

Subset 4: B and x_L

Subset 5: y_L and x_L

Because the mole fractions must sum to unity, y_H can be used in lieu of y_L and/or x_H in lieu of x_L .

The sixth possible subset of two is D and B . However, degrees of freedom also apply to subsets of the equations. One of the equations in the set is the total material balance. This equation does not permit targets for D and B to be specified independently.

1.6.4. Composition Control

The degrees of freedom analysis suggests that the following are possible:

1. For one of the product streams, specify a target for the flow and a target for the composition:
 - Specify distillate flow D and distillate composition y_L or y_H .
 - Specify bottoms flow B and bottoms composition x_L or x_H .

In practice, this is not common.
2. Specify a target for the flow of either product stream and a target for the composition of the other product stream:
 - Specify distillate flow D and bottoms composition x_L or x_H .
 - Specify bottoms flow B and distillate composition y_L or y_H .

This is commonly used for single-end composition control.
3. For both product streams, specify a target for the composition.
 - Specify distillate composition y_L or y_H and bottoms composition x_L or x_H .

This is double-end composition control.

The latter combination is of particular interest. Specifically, the degrees of freedom are sufficient to control both compositions.

1.6.5. Double-End Composition Control

Many difficulties were experienced in the early attempts, and applications of double-end composition control remained rare until the 1970s. The degrees of freedom analysis only suggests that something is possible; it does not propose a control configuration that will be successful.

The root of most problems was interaction between the two composition loops. There is inherently some interaction in every double-end composition control configuration. Any change that affects the composition of one product stream will have some effect on the composition of the other product stream. For each component of the feed, if one additional unit of that component is removed in the distillate stream, then one unit less of that component must be removed in the bottoms stream.

The degree of interaction depends on many factors, including the purities of the products, the external reflux ratio, and the relative volatility of the components. A proposed control configuration must be analyzed in light of the degree of interaction exhibited by the column on which it will be installed. Eventually, double-end composition control will be implemented on about 80% of the distillation columns.

1.6.6. Values for Targets

Suppose the degrees of freedom analysis suggests that two targets can be independently specified. This does not mean that all combinations of values for the targets are acceptable.

Probably the best way to express this is that the values for the targets must be “within reason.” Basically, this means that the values specified for the targets do not result in values for other variables that are impossible to attain. For distillation applications, the values specified for the targets must not give results such as the following:

1. A value for a composition that is less than 0% or greater than 100%.
2. A value for a flow that is negative. Reversible flow is not permitted for the distillate product, the bottoms product, reflux, and so on.

Mathematically, negative values could certainly be computed. In the formulation of the problem, inequalities such as $D \geq 0$, $B \geq 0$, and $0 \leq y_L \leq 1$ should be included. But instead of writing these explicitly, phrases such as “within reason” are sometimes applied.

1.6.7. Recovery

The recovery is the fraction of the feed that goes to a respective product stream. For the distillate product, the recovery is D/F ; for the bottoms product, the recovery is B/F . The recovery is often an important measure of column efficiency. If the distillate product is the salable product, improvements in the distillate recovery increase the amount of the desirable product that is available for sale.

The recovery is related to the various compositions (feed, distillate, and bottoms) and vice versa. This is vividly illustrated when the component material balance for the light component is rewritten as follows:

$$F z_L = D y_L + B x_L = D y_L + (F - D)x_L = D(y_L - x_L) + F x_L$$

$$F(z_L - x_L) = D(y_L - x_L)$$

$$\frac{D}{F} = \frac{z_L - x_L}{y_L - x_L}$$

When controlling product compositions, the usual approach is to focus on the energy terms (reflux and boilup). However, ignoring the role of the column material balance is an invitation for problems.

1.7. ENERGY AND THE SEPARATION FACTOR

In a distillation, column separation is attained by successive stages that essentially involve vaporization of a liquid and condensation of a vapor. Both involve energy. Except in towers with side heaters and/or side coolers, the energy for vaporization is provided largely by the reboiler, and the energy released by condensation is removed largely by the condenser.

Since energy is providing separation, the intuitive conclusion is that product compositions must be controlled through energy, which in most towers means the boilup and the reflux. The result is the double-end composition control configuration in Figure 1.11, in which the distillate composition is controlled by adjusting the reflux and the bottoms composition is controlled by adjusting the boilup. This is indirect material balance control for both product compositions—both the distillate flow and the bottoms flow are determined by the difference in two energy terms.

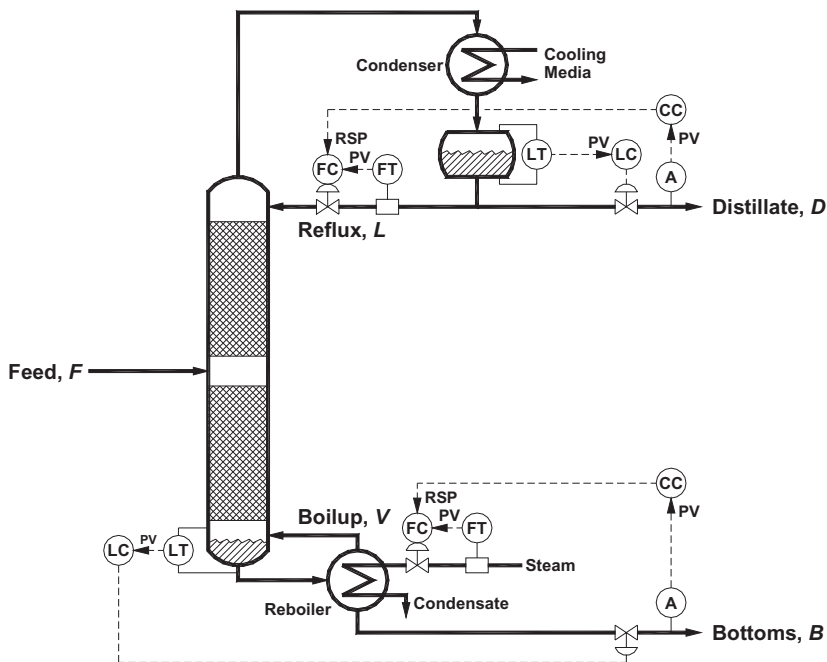


Figure 1.11. Double-end composition control configuration using an energy term for each product composition.

In a subsequent chapter, interaction analysis will be introduced as the tool for analyzing the degree of interaction in a proposed control configuration for distillation. In most cases, the degree of interaction for the configuration in Figure 1.11 is high, which translates into operational problems in the field. The degree of interaction is usually much lower for configurations in which one composition is controlled by manipulating an energy term and the other composition is controlled by manipulating a product draw.

Sometimes, it is difficult to convince people that what seems intuitive is perhaps off-base, at least in some cases. Distillation is a complex process, which complicates making the argument that controlling a product composition with a product draw is not only possible but appropriate. For double-end composition control, one of the compositions must be controlled by an energy term (D and B are not independent variables). But the other composition can be controlled using a product draw, and in most towers, this provides the least degree of interaction.

The objective of this section is to present the argument that controlling a product composition with a product draw just might make sense. To make this argument, a relationship between separation and energy is required. This is a complex relationship, even for binary distillation. The objective herein is to provide an insight into the issues, not to use the relationship for computational purposes. To keep it simple, the presentation will rely on the following:

1. An approximate relationship between separation and energy;
2. A binary separation.

However, the conclusions apply to multicomponent columns as well.

1.7.1. Fenske Equation

Most approximate relationships for separation are derived in some manner from the Fenske equation that relates the product compositions to the relative volatility and the number of stages:

$$\frac{y_L(1-x_L)}{x_L(1-y_L)} = \frac{y_L x_H}{x_L y_H} = \frac{y_L/x_L}{y_H/x_H} = \alpha^n,$$

where

- n = number of theoretical stages;
- α = relative volatility (ratio of vapor pressures) of the light component relative to the heavy component.

Unfortunately, the Fenske equation has a serious restriction—it only applies at total reflux.

1.7.2. Separation Factor

When the tower is not operating on total reflux, the term α^n in the Fenske equation is replaced by the separation factor S :

$$S = \frac{y_L(1-x_L)}{x_L(1-y_L)} = \frac{y_L x_H}{x_L y_H} = \frac{y_L/x_L}{y_H/x_H}.$$

For most columns, the numerical value of the separation factor will be large, especially if the products are low in impurities (y_H in the distillate product; x_L in the bottoms product). Suppose both products are 95% pure, which is not an especially high purity. The value of the separation factor is

$$\begin{aligned} y_L &= 0.95, \\ y_H &= 0.05, \\ x_L &= 0.05, \\ x_H &= 0.95, \\ S &= \frac{y_L/x_L}{y_H/x_H} = \frac{0.95/0.05}{0.05/0.95} = 361. \end{aligned}$$

In practice, values of 1000 or more for the separation factor are typical.

1.7.3. Separation Factor and Control

The above example computed the separation factor from the distillate and bottoms compositions. But in practice, the distillate and bottoms compositions depend on the separation factor and the column material balances.

The value of the separation factor depends on the following:

Number of theoretical stages n . Largely determined by the column design; operating variables have only a minor influence.

Relative volatility α . Depends primarily on the materials being separated. Column pressure has some influence and is occasionally used for optimization but never for regulatory control.

Energy input Q . Variable that the control system can influence through the reflux and boilup rates.

In order to affect the separation factor in an operating tower, the control system must change the energy terms. In a sense, this reinforces one's intuition that product compositions should be controlled through energy.

Although a few relationships have been proposed, relating the separation factor to the number of theoretical stages n , the relative volatility α , and the energy (either as reflux ratio or boilup ratio) is a challenge. Fortunately, this

is not necessary for the discussion that follows—again, the objective is to gain insight, not to perform computations.

1.7.4. Coupling Material Balance with Separation

For a binary tower, the following equations relate the product compositions (y_L and x_L) to the D/F ratio (the recovery for the distillate product) and the separation factor S :

$$\begin{aligned} \text{Material balance: } \frac{D}{F} &= \frac{z_L - x_L}{y_L - x_L} \\ \text{Separation: } S &= \frac{y_L(1 - x_L)}{x_L(1 - y_L)} \end{aligned}$$

With four unknowns (D , S , y_L , and x_L) in two equations, the solution can be viewed in two ways:

1. Though its final control elements, the control system specifies the product draws (which determine D/F) and the energy terms (which determine the separation factor S). The above two equations can be solved for the product compositions y_L and x_L .
2. In a double-end composition control application, the product specifications provide targets for y_L and x_L . The above two equations can be solved for the recovery D/F and the separation factor S . Basically, this is the solution that the controls must obtain in basically a trial-and-error fashion.

Even for binary columns, the solution of the two equations requires iterative procedures. Consequently, these equations are of little (or no) computational value. However, they provide the basis for gaining insight into the control options for a column.

1.7.5. Approximations in Separation Factor Equation

In many columns, the impurities y_H and x_L in both products are small, which permits the following approximations to be made:

$$\begin{aligned} 1 - y_H &\cong 1, \\ 1 - x_L &\cong 1. \end{aligned}$$

With these approximations, the expression for the separation factor simplifies to the following:

$$S = \frac{y_L/x_L}{y_H/x_H} = \frac{(1-y_H)/x_L}{y_H/(1-x_L)} \cong \frac{1}{y_H x_L}.$$

In a previous example, the purity of both product streams was 95%, giving a separation factor of 361. With the above approximation, the separation factor is

$$S = \frac{1}{0.05 \times 0.05} = 400.$$

The higher the purity, the less the difference.

1.7.6. Logarithmic Equation for Separation Factor

When analyzing the expressions for the separation factor, the nonlinear nature of the equation leads to complications. But when the impurities in both products are small, expressing the relationship in terms of logarithms gives a linear result:

$$\ln S = -\ln y_H - \ln x_L = (-\ln y_H) + (-\ln x_L).$$

The compositions y_H and x_L are both less than 1, so the quantities $(-\ln y_H)$ and $(-\ln x_L)$ are positive values.

1.7.7. Graphical Representation

The objective of the graphical representation in Figure 1.12 is to illustrate this point. There are two scales:

Upper scale. The composition y_H of the impurity in the distillate.

Lower scale. The composition x_L of the impurity in the bottoms.

Both scales are logarithmic. The two scales are joined for a composition of 1.0 (which is zero on a log scale).

Starting with values for the separation factor S and the distillate draw D gives product compositions of y_H impurity in the distillate and x_L impurity in the bottoms. These are represented on the graph as the distances $(-\ln y_H)$ for the impurity in the distillate and $(-\ln x_L)$ for the impurity in the bottoms. The sum of these two distances is $(\ln S)$. Starting from the solution designated as the “base case” in Figure 1.12, the effect of increasing the separation factor S and then increasing the distillate flow D will be illustrated.

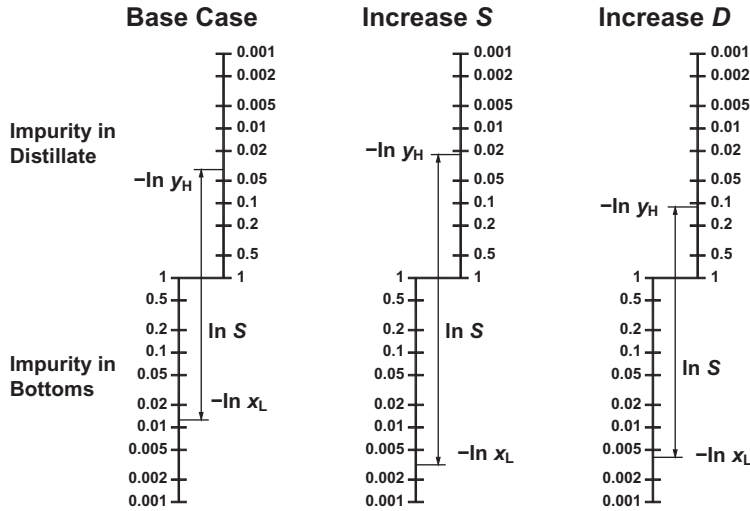


Figure 1.12. Effect of the separation factor S and the distillate draw D on product compositions.

1.7.8. Increasing the Separation Factor

The impact of increasing the separation factor is simple: the value of $(\ln S)$ increases. This means that either $(-\ln y_H)$ increases, $(-\ln x_L)$ increases, or both. In practice, there is some increase in both, as illustrated by the “Increase S ” solution in Figure 1.12.

However, the increase is usually not by the same amount. The extremes for the possibilities are as follows:

- The major impact is on y_H (the impurity in the distillate), with little change in x_L (the impurity in the bottoms).
- The major impact is on x_L (the impurity in the bottoms), with little change in y_H (the impurity in the distillate).

However, it is also possible for the impact to be about evenly distributed between y_H and x_L . No general statements can be made about what result to expect. The only way to obtain answers is to use a distillation column model to examine the effect of increasing the energy input to the column.

1.7.9. Increasing the Distillate Draw

If the separation factor S is held constant, increasing the distillate draw D increases the concentration of the heavy component in every separation stage. Consequently, the results will be as follows:

1. The composition y_H (the impurity in the distillate) increases.
2. The composition x_L (the impurity in the bottoms) decreases.

This is illustrated by the “Increase D ” solution in Figure 1.12. There is no change in $(\ln S)$, so on the logarithmic scales, the magnitude of each change in the compositions is exactly the same. The change is an increase on one scale, but is a decrease on the other by exactly the same amount.

1.7.10. Impact for Controlling One Composition

For the example in Figure 1.12, suppose y_H (impurity in the distillate) is to be controlled using energy. The separation factor S has a greater impact on the bottoms composition x_L than on the distillate composition y_H . The process disturbances cause some variance in the distillate composition y_H . As the variance propagates from the distillate composition to the separation factor and then to the bottoms composition, it is amplified:

1. To compensate for the variance in y_H , the composition controller changes the separation factor S (through changing the energy input to the column). The smaller the effect of the separation factor S on a composition, the larger the changes required in the separation factor S in order to maintain the composition at its target.
2. The changes in separation factor S will affect the bottoms composition. For the example, in Figure 1.12, the effect of the separation factor S on x_L is larger than its effect on y_H . This significantly increases the variance in x_L .

Since only y_H is being controlled, is variance in x_L of any concern? Variance in x_L is likely to impact some downstream operation. Situations where reducing the variance in one variable greatly amplifies the variance in another should be avoided.

1.7.11. Double-End Composition Control

Controlling only one composition can usually be accomplished via the energy streams that affect the separation factor S . Potentially, the control actions taken to control that composition could propagate significant variance to the other composition. But since this composition is not being controlled, there will be no closed-loop response to the variations propagated to the other stream.

When both compositions are to be controlled, the issues pertaining to interaction must be resolved. Normally, the composition most affected by the separation factor must be controlled by making changes in the energy streams. For the case illustrated in Figure 1.12, this means that the bottoms composition

must be controlled using the energy streams. Changes in the separation factor S have more influence on the bottoms composition than on the distillate composition.

If the bottoms composition is controlled via the separation factor, how does one control the overhead composition? The relationships on which the graphs in Figure 1.12 are based suggest that the distillate composition can only be controlled by changing a product draw. In essence, the bottoms composition is controlled through changes in energy; the distillate composition is controlled through the material balance.

1.8. MULTICOMPONENT DISTILLATION

One always likes to start with the simple, which in distillation means binary distillation. A few binary distillation columns are found in production facilities, but most are multicomponent.

The next section will discuss the stage-by-stage separation models that are now routinely used in column design. These provide very accurate solutions, but at the expense of considerable complexity. Design and control are fundamentally different. For a column in a specified service (feed flow and composition), the problems are stated as follows:

Design. Calculate the reflux, boilup, and so on, required to give specified product compositions.

Control. The current operating conditions in the tower are known (reflux flow, boilup, product compositions, etc.). Calculate the change in the manipulated variable, such as the boilup, required to change the controlled variable, such as the bottoms composition, from its current value to its target.

To summarize, design works on actual values, and for this, accuracy is crucial. However, control works on changes (a change in the manipulated variable leads to a change in the controlled variable), and especially when the changes are small, approximations would certainly be acceptable.

The Hengstebeck approximation is one example that will be explained shortly. Prior to the computer era, columns were designed based on such approximations, but detailed models are now used in lieu of such approximations.

1.8.1. Heavy and Light Keys

In binary distillation, the components are referred to as the light component and the heavy component. The objective is to separate these two components. In multicomponent distillation, the corresponding terms are “light key” and “heavy key.” A column effects a separation between the two keys.

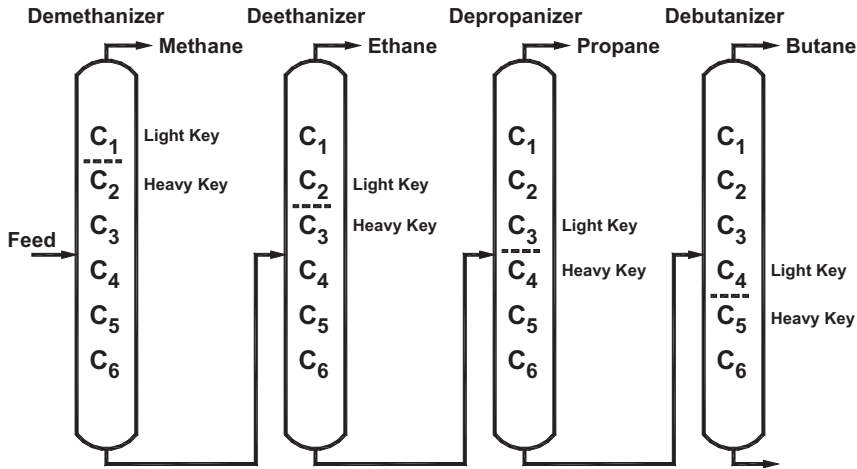


Figure 1.13. Columns in series.

Figure 1.13 illustrates a sequence of columns such as found in a gas plant. The feed to the first column (the demethanizer) is a mixture of methane (C₁), ethane (C₂), propane (C₃), butane (C₄), and so on. The columns and their key components are as follows:

- Demethanizer.** Separates methane (the light key) from ethane (the heavy key).
- Deethanizer.** Separates ethane (the light key) from propane (the heavy key).
- Depropanizer.** Separates propane (the light key) from butane (the heavy key).
- Debutanizer.** Separates butane (the light key) from pentane (the heavy key).

In binary distillation, the light and heavy components appear in both product streams. A component that appears in both product streams is said to be a “distributed component.” In the Hengstebeck approximation, only the light and heavy keys are distributed, the assumptions being as follows:

1. All components of the feed that are lighter than the light key leave with the distillate product. Basically, these components are treated as noncondensable gases.
2. All components of the feed that are heavier than the heavy key leave with the bottoms product. Basically, these components are treated as nonvolatile liquids.

Consider the depropanizer in the separation train in Figure 1.13. The light key is propane (C_3); the heavy key is butane (C_4). All methane and ethane leave with the distillate product; all pentane and heavier components leave with the bottoms.

1.8.2. Components

When using the Hengstebeck approximation for the depropanizer in Figure 1.13, the four components are as follows:

Lighter-than-light key (LL). This includes all methane and all ethane. These components leave entirely with the distillate product. This is a pseudocomponent whose composition in the distillate product is y_{LL} . None of these components appear in the bottoms, so x_{LL} is zero.

Light key (L). This is propane. This component appears in both the distillate product and the bottoms product, and thus is a distributed component. In a multicomponent system, y_L is the composition of the light key in the distillate and x_L is the composition of the light key in the bottoms.

Heavy key (H). This is butane. This component appears in both the distillate product and the bottoms product and thus is a distributed component. In a multicomponent system, y_H is the composition of the heavy key in the distillate and x_H is the composition of the heavy key in the bottoms.

Heavier-than-heavy key (HH). This includes all pentane and heavier components. These components leave entirely with the bottoms product. This is a pseudocomponent whose composition in the bottoms product is x_{HH} . None of these components appear in the distillate, so y_{HH} is zero.

1.8.3. Component Material Balances

The light key (L) and the heavy key (H) are real components; the lighter-than-light (LL) and the heavier-than-heavy (HH) are pseudocomponents. A component material balance can be written for each:

$$\begin{aligned} F z_{LL} &= D y_{LL}, \\ F z_L &= D y_L + B x_L, \\ F z_H &= D y_H + B x_H, \\ F z_{HH} &= B x_{HH}. \end{aligned}$$

The equations for the light key and the heavy key are identical to those written for the light and heavy components of binary distillation.

1.8.4. Separation Factor

The objective of the Hengstebeck approximation is to permit the relationships developed for binary distillation to be applied to multicomponent distillation. For example, the equation for the separation factor is still

$$S = \frac{y_L}{x_L} \frac{x_H}{y_H} = \frac{y_L/x_L}{y_H/x_H}.$$

For multicomponent separations, y_L and x_L pertain to the light key; y_H and x_H pertain to the heavy key.

1.9. STAGE-BY-STAGE SEPARATION MODEL

The reality is that all components of the feed to a tower appear to some extent in both product streams. For a depropanizer, the amount of ethane in the bottoms will be extremely small, but some will be present. The amount of methane in the bottoms will be even smaller, but some will be present. Similar statements can be made with regard to the pentane in the distillate stream. Assuming the composition of these minor components to be zero is not always acceptable.

Another issue arises when isomers are present. Consider butane. In most gas plants, butane is primarily n-butane. However, some isobutane is present. Isobutane is more volatile (has a lower boiling point) than n-butane. Consequently, the ratio of isobutane to n-butane in the distillate will be higher than their ratio in the feed.

1.9.1. Separation Model

The stage-by-stage separation model is based on the following equations:

1. A component material balance is written for each component on each stage. If the feed to the column contains 10 components and there are 30 stages in the column, this gives 300 equations.
2. An energy balance is written for each stage (equimolal overflow is not assumed).
3. Realistic vapor–liquid equilibrium relationships can be used. Without such relationships, the relative volatility is assumed to be constant, which is rarely the case.

This gives a large number of nonlinear equations. Only computers can solve such equations. Today, commercial software packages are available that are specifically designed to solve the equations that arise in distillation, and most companies have standardized on one (or perhaps two) of these.

1.9.2. Issues for Control

The stage-by-stage separation models are occasionally used in on-line optimization and similar undertakings. Incorporating into regulatory control configurations poses two problems:

High dimensionality. The total number of equations is very large. For 10 components and 30 stages, the number of equations is in excess of 300. The concepts for controlling multivariable processes are well known, but high dimensionalities present a variety of problems, including numerical difficulties.

Nonlinear equations. The vapor–liquid equilibrium relationships are highly nonlinear. Unfortunately, most of the currently available control technologies are based on linear systems theory.

Difficulties such as these can certainly be overcome. However, there must be an incentive to do so. The improved accuracy of the stage-by-stage separation models led to improved column designs, which provided the incentives to develop techniques specifically for solving the model equations. But regulatory control depends primarily on repeatability, not accuracy. To date, no incentives have been identified that justify developing methods to incorporate the stage-by-stage separation models into regulatory control configurations.

1.9.3. Start with a Column Model

Even though the stage-by-stage separation model will not be used directly in the regulatory control configuration, developing such a model must be the starting point for any control effort directed to a distillation column.

When analyzing a control problem associated with a distillation column, the first step is to make sure the column is capable of doing what is desired. A common practice in production facilities is to blame all problems on the control system. Process problems often lead to the control system being unable to maintain a process variable at its target. But if the process is unable to attain the target value, control efforts directed at the problem are doomed to failure. Distillation is a complex unit operation that offers many possibilities for problems to arise. Some of these problems can be very subtle, and some problems will only arise under certain situations.

Normally, the data set for a stage-by-stage separation model is part of the “deliverables” from the design team. Ideally, the startup effort should include collecting data from the column and calibrating the model to the process, but this is not always the case. Starting with whatever is available, one proceeds as follows:

1. Collect current operational data from the tower (flows, temperatures, compositions, etc.).

2. Calibrate the separation model to the process by adjusting parameters such as stage efficiencies.

If the column performance is far different from what the model suggests, this must be resolved before proceeding with any true control work.

1.9.4. Steady-State Issues in Control

The common impression is that regulatory control is only concerned with process dynamics. This view is reinforced by the typical academic course on “process control,” which is in reality a mathematics course on linear systems theory. But in most applications, the key issues pertain to the steady-state behavior of the process, not its dynamics. Consequently, much that is relevant to regulatory control of a distillation column can be understood from its steady-state model.

One relevant characteristic is the sensitivity. If you increase the energy flows (energy in at the reboiler and energy out at the condenser), you would expect the impurities in both product streams to decrease. But will the major effect be in the distillate composition, will it be in the bottoms composition, or will the impact be about the same in both product compositions? Distillation is a complex unit operation, so questions such as these can only be answered with confidence when the answers are obtained via a good separation model. This is especially true in complex towers and towers separating nonideal mixtures.

In distillation columns, some degree of interaction always exists between the product compositions. When double-end composition control is being attempted, this interaction must be analyzed very carefully. There are two aspects of interaction—steady state and dynamic. The dynamics of the composition loops will be about the same, which makes the steady-state aspects very significant. The analysis of this interaction can be based entirely on results obtained from the separation model.

1.9.5. Limitations

Engineering involves obtaining numerical answers to numerical problems. In this regard, the stage-by-stage separation models are superb. Probably, the main concern is the quality of some of the relationships (vapor–liquid equilibrium data, heat capacity equations, etc.). Even minor changes in these relationships can give significantly different results.

But suppose one’s objective is to obtain insight into how a specific tower behaves, or possibly to improve one’s understanding of distillation in general? The stage-by-stage separation models are not very useful. One can obtain a series of solutions by changing certain parameters and examining their effect on the results. One is quickly inundated with data. The stage-by-stage separation models are very good at one thing—determining the numerical

solution to the column equations for a specific situation (feed rate, reflux ratio, etc.).

1.9.6. Depropanizer Model

In order to illustrate how information relevant to control can be obtained from the steady-state model of a distillation column, a simplified version of a depropanizer from a production facility will be used. Just to avoid carrying too many numbers, the following simplifications are made:

1. The feed (from a deethanizer) contained very little ethane and almost no methane. The methane composition is set to zero.
2. All components heavier than pentane are treated as pentane.

The feed contains only four components, their compositions being the following:

- Ethane (C_2): 0.4 mol%
- Propane (C_3): 23.0 mol% (the light key)
- Butane (C_4): 37.0 mol% (the heavy key)
- Pentane (C_5): 39.6 mol%

1.9.7. Separation Sections

The upper separation section (above the feed stage) has 11 ideal stages; the lower separation section has 9. The column has a total condenser (distillate product is liquid). Including the reboiler, the column has a total of 21 ideal stages. Stages will be numbered from the top of the tower. That is, stage 1 is at the top of the upper separation section; stage 20 is at the bottom of the lower separation section; stage 21 is the reboiler.

The column pressure is 16.0 barg. The overhead product is primarily propane, so the overhead temperature will be approximately the boiling point of propane (50.3°C at 16.0 barg). Coefficients for the relationships for vapor pressures, heat capacities, and so on, are obtained from Yaws [1].

In distillation calculations, a common approach is to base the calculations on a feed rate of 100 mol per unit time (hour, minute, etc). Herein mol/h will be used. The feed enters as a liquid under pressure at 105°C. About 10% of the feed flashes upon entry into the tower.

1.9.8. Base Case

One of the first steps in the analysis of any control problem with a distillation column is to develop a column simulation that matches the current plant

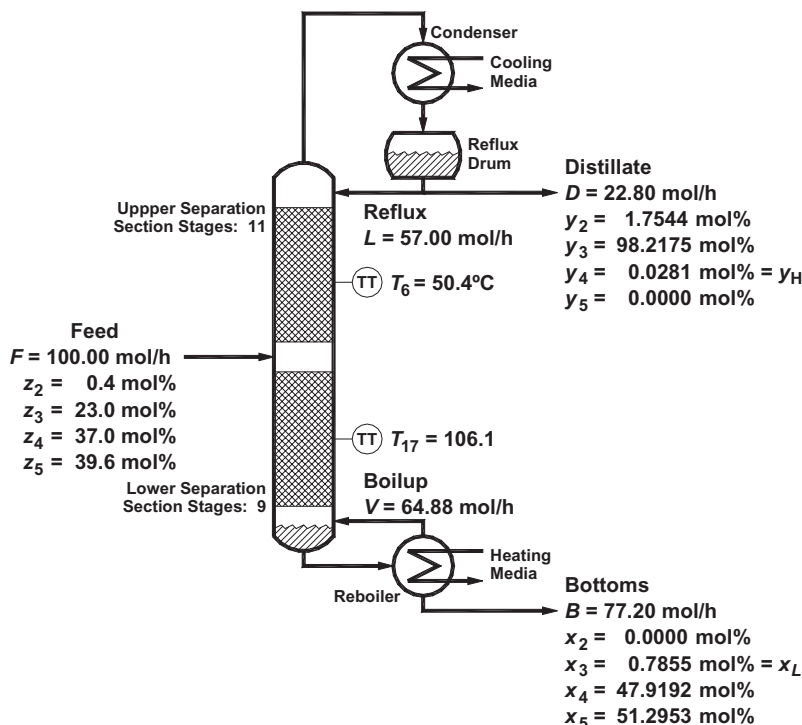


Figure 1.14. Depropanizer model base case solution.

operating conditions. This solution becomes the “base case” for subsequent analyses. For the depropanizer, the solution for the base case is computed for the following conditions:

Distillate flow $D = 22.80 \text{ mol/h}$

External reflux ratio $(L/D) = 2.5$ (since $D = 22.80$, $L = 57.00 \text{ mol/h}$).

Figure 1.14 summarizes the solution of the steady-state model for the base case.

A subsequent discussion on product compositions will explain why the column is operated in this manner, but briefly, the objective is for the propane product to contain as much ethane as the specifications permit (and consequently very little butane) and for the butane product (to the next column) to contain as much propane as the specifications permit.

In practice, one cannot assume that the column is well-designed for its current service. Although design mistakes are occasionally made, the most likely explanation is that the column is not being operated for the service for which it was designed.

1.9.9. Utility in Control Analyses

When any operational problem arises for a distillation column, the basic controls are frequently viewed as the culprit. Sometimes they are, but not always. The analysis of the problem should begin with a stage-by-stage column simulation. Especially when the column is not operating in the service for which it was designed, two aspects must be verified:

1. The column is performing in a manner consistent with its design.
2. The column can deliver the performance being demanded by current operations.

Beyond these, there are other uses of the stage-by-stage simulation, including the following:

Temperature profile. Improper location of the control stages leads to problems with temperature controls.

Internal flows. The internal vapor and liquid flows must be within the limits imposed by the tower internals.

Sensitivities. Distillation is a complex unit operation with significant interaction between the operating variables.

1.9.10. Temperature Profile

A common practice is to use temperature measurements in either or both of the following manners:

Upper control stage. The temperature of this stage (from the upper separation section) is used as the measured variable for a temperature controller that adjusts either the reflux flow L or the distillate flow D .

Lower control stage. The temperature of this stage (from the lower separation section) is used as the measured variable for a temperature controller that adjusts either the heat input to the reboiler (and consequently the boilup flow V) or the bottoms flow B .

When either or both of these are used, the temperature profile within the tower must be examined. The temperature on each stage is computed through the energy balance that is incorporated into the stage-by-stage calculations. The temperature profile is obtained by plotting these temperatures, the result being the graph presented in Figure 1.15.

The graph also indicates the location of the control stages. The issues pertaining to using stage temperatures in control configurations will be examined in the next chapter.

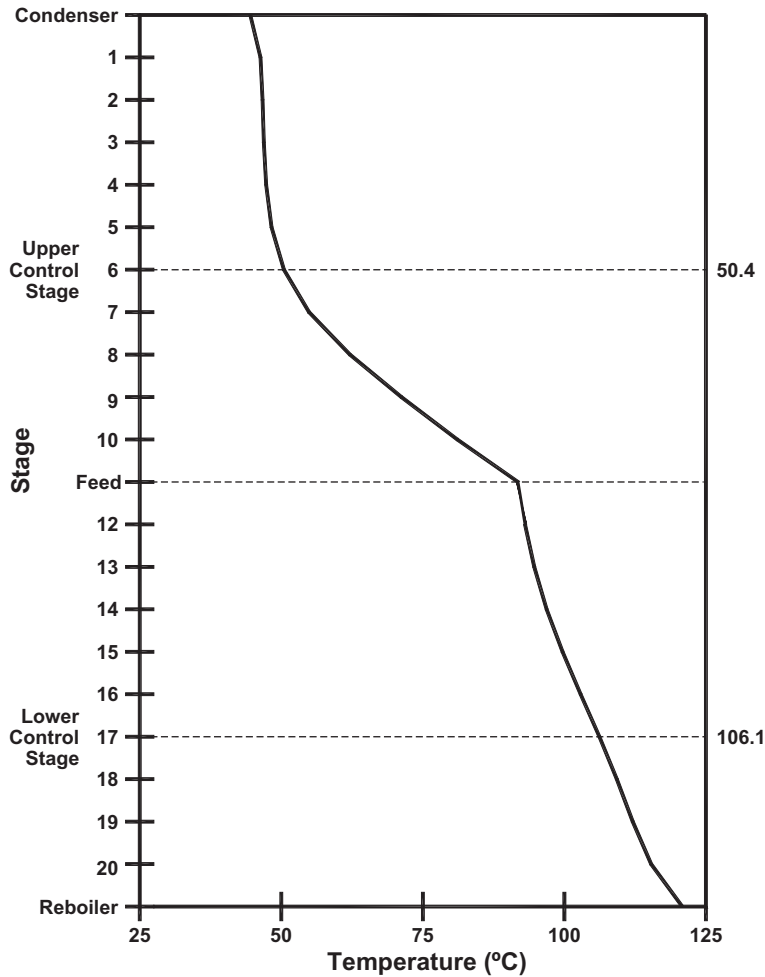


Figure 1.15. Temperature profile.

1.9.11. Internal Vapor and Liquid Flows

The stage-by-stage calculations also provide values for the vapor and liquid flows leaving each stage. Figure 1.16 presents the vapor and liquid flows for the depropanizer plotted in a manner similar to the stage temperatures. The most noticeable change is at the feed stage:

Liquid flow. About 90% of the feed contributes to the liquid flow, so the liquid flow below the feed stage is significantly higher than the liquid flow above the feed stage.

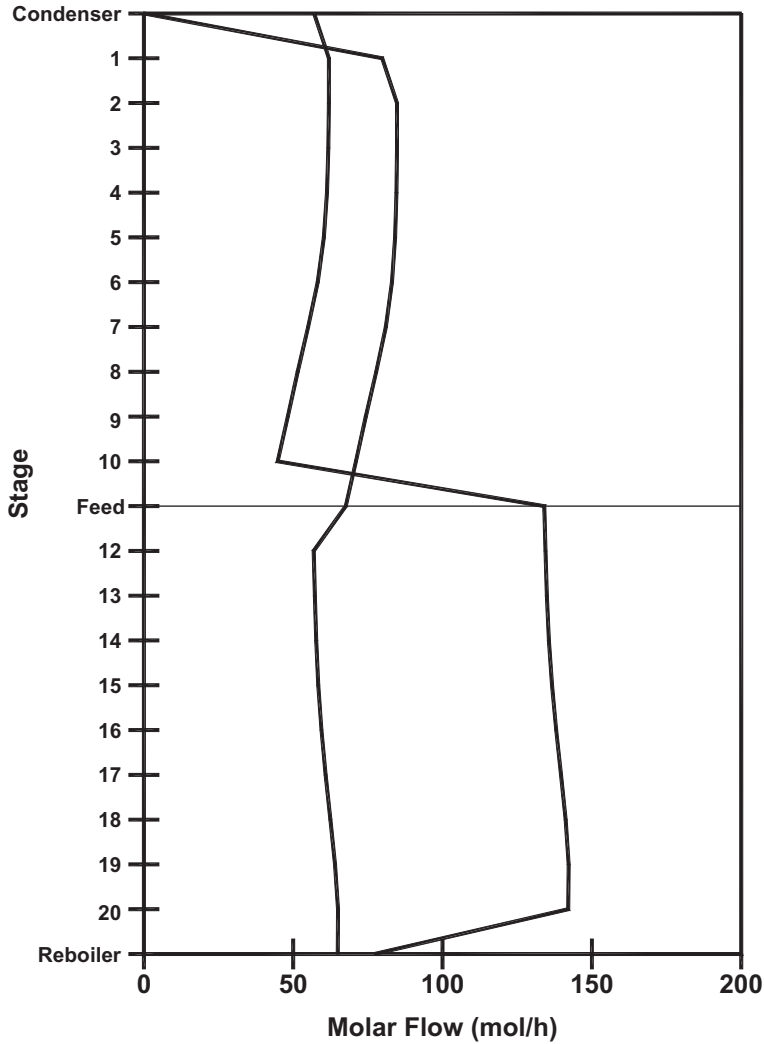


Figure 1.16. Vapor and liquid flows.

Vapor flow. About 10% of the feed contributes to the vapor flow, so the vapor flow above the feed stage is only slightly higher than the vapor flow below the feed stage.

Although not constant, the changes in liquid and vapor flows within each separation section are nominal. This is generally the case for mixtures that are close to ideal. Mixtures of hydrocarbons such as propane and butane deviate only slightly from ideal. Larger departures are normally the case

for other chemicals, and vapor and liquid flows within a separation section can easily change by a factor of 2. The effect of such changes on the tower internals can give operational problems that are sometimes interpreted as control problems.

1.9.12. Sensitivities

Suppose the bottoms composition is being controlled by adjusting the heat input to the reboiler. In the stage-by-stage calculations, this is equivalent to adjusting the boilup V . One parameter that significantly affects the performance of this composition controller is the sensitivity of the bottoms composition to a change in the boilup V . The only way to obtain a value for this sensitivity is by using the stage-by-stage separation model.

The bottoms composition controller makes adjustments in the boilup V so as to maintain the bottoms composition at its target. However, these adjustments affect other variables within the tower. Specifically, control actions taken by the bottoms composition controller become disturbances to the composition of the distillate product. The significance of these disturbances is determined by another sensitivity, specifically, the sensitivity of distillate composition to a change in the boilup V .

Using the stage-by-stage model to calculate such sensitivities for the current operating conditions is relatively easy. The more challenging task is to determine what changes in the operating conditions will have a significant effect on the value of the sensitivity. Changes in the feed rate, feed composition, recoveries, product composition targets, and so on, have the potential to affect the value of the sensitivity, which will result in performance problems in the controller.

A natural extension of the use of sensitivities is to compute the degree of interaction for a proposed control configuration. This becomes crucial for double-end composition control configurations, and an entire chapter is subsequently devoted to this subject. The availability of a stage-by-stage separation for a column permits the degree of interaction to be assessed prior to implementing a control configuration.

1.9.13. Precision

In this context, precision will be used as in C++—the number of digits after the decimal point for representing numerical values. Sensitivities are computed from the difference in two values. Distillation is nonlinear, so the difference in the two values must be the result of a small difference in variables such as boilup, reflux, and product flow. Consequently, the difference in the two values will be small.

Herein the values of flows, compositions, temperatures, and so on, will be routinely represented to a greater precision than justified by the separation

model. Computing sensitivities depends on the ability of the model to translate a small change in one variable to a small change in another. Models can generally do this better than the accuracy of the individual values. This is much like the accuracy versus repeatability of a measurement device—the repeatability is usually better than the accuracy.

1.10. FORMULATION OF THE CONTROL PROBLEM

Distillation columns are relatively complex unit operations with a large number of permutations. At this point, the column in Figure 1.17 is arbitrarily used as the starting point. The key aspects of the configuration in Figure 1.17 are as follows:

- The tower is a two-product tower.
- Both product streams are liquid (condenser is a total condenser).
- The reflux drum is partially filled (the reflux drum level must be measured and controlled).
- The condenser transfers heat to cooling water.
- The reboiler is heated with steam.

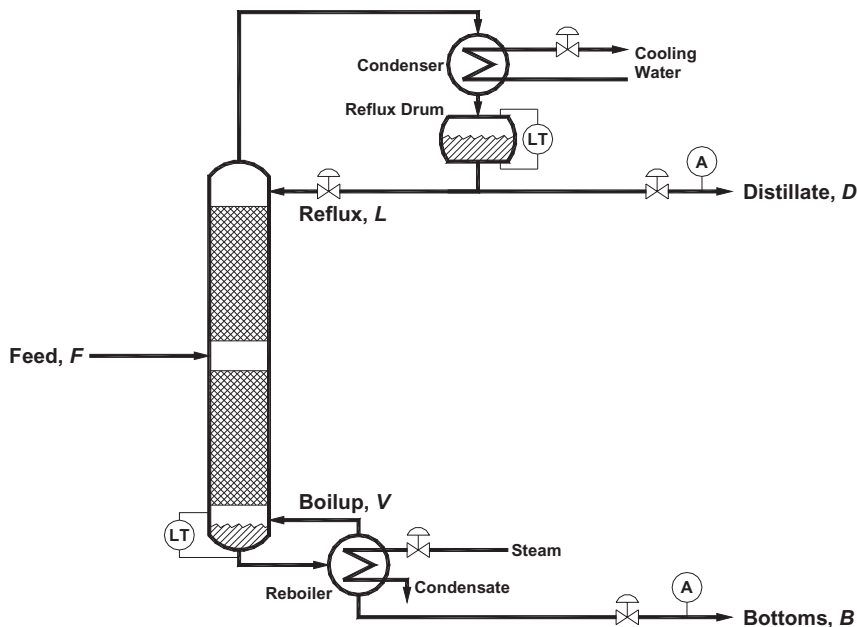


Figure 1.17. Controlled and manipulated variables for a two-product tower.

1.10.1. Reboiler and Condenser

Using a steam-heated reboiler in the P&I diagram in Figure 1.17 is reasonable—steam is the most common heating medium used in production facilities. Alternatives such as hot oil and fired heaters will be discussed along with various reboiler configurations in the subsequent chapter devoted to reboilers.

There are a couple of issues pertaining to the condenser arrangement in Figure 1.17:

- Although water-cooled condensers are probably installed most frequently, air-cooled condensers are common.
- Varying the water flow through the condenser raises lots of issues. Alternatives such as hot gas bypass arrangements are often installed, especially when natural water is used as the cooling media.

All of these are discussed in the subsequent chapter on pressure control and condensers. Some mechanism by which the control system can vary the heat transfer rate in the condenser is required, but the exact nature of that mechanism has little impact on the remaining control issues for the tower.

The “default” condenser arrangement used in most illustrations within this book is a water-cooled condenser with a control valve on the cooling water, as in Figure 1.17. The reason: this is the simplest to draw. Another simple arrangement is a control valve in the overhead vapor line, but this arrangement is not commonly installed.

1.10.2. Controlled Variables

In the language of control engineers, a controlled variable is a process variable whose value is to be maintained at or near a target (or set point). For the column illustrated in Figure 1.17, there are five controlled variables:

1. bottoms level,
2. reflux drum level,
3. column pressure,
4. distillate composition,
5. bottoms composition.

Figure 1.17 indicates composition measurements on both product streams. But as noted previously, practical considerations often locate the analyzer elsewhere or even utilize a control stage temperature in lieu of composition.

1.10.3. Manipulated Variables: Instrument Context

A manipulated variable is a variable whose value is at the discretion of the control system. At the hardware level, these are the physical outputs of the

controls, which are signals that drive a final control element. In distillation columns, the final control element is usually a valve, but occasionally is a variable speed drive, a power regulator to an electric heater, and so on.

At the instrument level, the five manipulated variables for the distillation column illustrated in Figure 1.17 are as follows:

1. distillate valve opening,
2. bottoms valve opening,
3. reflux valve opening,
4. condenser cooling water valve opening,
5. reboiler steam valve opening.

1.10.4. Manipulated Variables: Process Context

For each manipulated variable, there is a process variable that corresponds to the instrument variable. Consider the stage-by-stage separation models. One never specifies the distillate valve opening; one specifies the distillate flow. One could always consider the process variable to be the flow through the control valve, but this is not very satisfactory for the condenser or the reboiler. Instead, the choices are the following:

Condenser. The cooling water flow affects the condenser heat transfer rate Q_C , which in turn affects the condensation rate within the condenser. At steady state, the overhead vapor rate V_C must be consistent with the condensation rate. Herein, the overhead vapor rate V_C will generally be used as the manipulated variable associated with the condenser.

Reboiler. The steam flow determines the reboiler heat transfer rate Q_R , which in turn affects the vaporization rate within the reboiler. The vaporization rate within the reboiler is the boilup V . Herein, the boilup V will generally be used as the manipulated variable associated with the reboiler.

Table 1.3 lists the manipulated variables in both the instrument context and the process context for the column illustrated in Figure 1.17. For the

TABLE 1.3. Manipulated Variables for Column in Figure 1.17

Instrument Context	Process Context
Distillate valve opening	Distillate flow D
Bottoms valve opening	Bottoms flow B
Reflux valve opening	Reflux flow L
Condenser cooling water valve opening	Heat transfer rate in condenser Q_C or overhead vapor flow V_C
Reboiler steam valve opening	Heat transfer rate in reboiler Q_R or boilup V

condenser and the reboiler, the manipulated variable in the instrument context depends on the equipment, but the manipulated variable in the process context does not.

1.10.5. Multivariable Control Problem

There are five controlled variables; there are five manipulated variables. This constitutes a 5×5 multivariable control configuration:

Controlled Variable	Manipulated Variable
Bottoms level	Distillate flow D
Reflux drum level	Bottoms flow B
Column pressure	Reflux flow L
Distillate composition	Overhead vapor flow V_C
Bottoms composition	Boilup V

This is only a listing of the controlled and manipulated variables; there is no significance to the order of either the controlled or manipulated variables.

In the single-loop approach to column control, a proportional–integral–derivative (PID) controller is configured for each of the controlled variables. The output of the controller must be to one of the control valves, but which one? The term “pairing” refers to selecting the manipulated variable to be used for each controlled variable. No pairing is implied in the previous list.

1.11. TOWER INTERNALS

The purpose of the tower internals is to facilitate mass transfer between the vapor and liquid phases within the tower. The options for tower internals are as follows:

Trays. Vapor–liquid contact is enhanced by dispersing the vapor into the liquid retained on the tray. The number of trays within each separation section is determined by the number of theoretical stages required by the design and the tray efficiency.

Packing. Liquid flowing over the packing provides a large wetted surface area for vapor–liquid contact. The height of each packed section is determined by the number of theoretical stages and the height of packing equivalent to a theoretical stage.

Trays versus packing is a tower design choice. Trays were primarily used in the older towers. Until the advent of structured packing, the maximum height of a packed section was restricted (the packing crushes under its own weight).

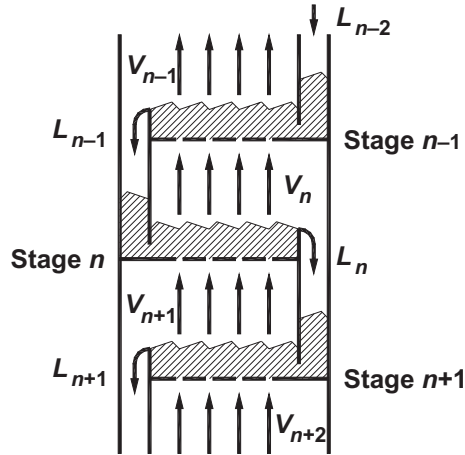


Figure 1.18. Vapor and liquid flows for trays.

But a packed tower tends to be smaller than a tray tower; in an existing tower, replacing trays with packing usually provides greater separation.

These and other design issues dictate the choice of trays versus packing. Control issues pertaining to trays versus packing are minor compared with the design issues.

1.11.1. Trays

Figure 1.18 illustrates the flows associated with trays. For a given tray, these are briefly as follows:

Liquid. Liquid flows from the tray above through a pipe called a downcomer. This liquid flows across the tray, then over a weir into the downcomer to the tray below.

Vapor. The vapor from the tray below enters through openings in the bottom of the tray and mixes with the liquid on the tray. The tray spacing provides for vapor–liquid disengagement so that, ideally, only vapor flows into the tray above.

Separation of liquid flow from vapor flow is not perfect:

Weeping. Some liquid “weeps” through the openings in the bottom of the tray, which to some extent short-circuits the vapor–liquid contact on the tray. Theoretically, valve caps and bubble caps prevent liquid weeping, but not in practice.

Entrainment. Ideally, the vapor and liquid totally disengage before the vapor enters the tray above. The more space between trays, the better the disengagement, but this adds height (and cost) to the tower.

Both are significantly affected by the vapor flow. Increasing the vapor flow reduces the weeping but increases the entrainment.

1.11.2. Vapor Flows for Trays

The limits on the vapor flows for a tray are as follows:

Minimum. The nature of the trays with regard to weeping establishes the minimum vapor flow. There must be some flow over the downcomer at all times. If not, the amount of liquid on the tray is inadequate and vapor–liquid contact is lost. If the vapor rate is too low, all liquid is lost from the trays, which rapidly increases the level in the reboiler. A shutdown on high reboiler level is likely to be initiated, which shuts off the heat to the column. But in any case, the result is a major upset to the tower.

Maximum. The vapor flow through the openings on the bottom of the tray results in a pressure differential across the tray. This pressure differential increases with the square of the vapor flow. If this pressure drop is too large, the consequence is a phenomenon known as “flooding.” This is a major upset to the tower, so flooding will be examined in detail shortly.

The consequences of both high vapor flows and low vapor flows can be painful. Flooding normally receives the most attention. Most towers are equipped with pressure drop measurements that can detect the onset of flooding, and operations personnel take high pressure drops seriously. Unfortunately, there is no convenient measurement that can draw attention to low vapor flows, so its consequences are often a surprise to operations personnel.

1.11.3. Liquid Flows for Trays

The limits on the liquid flows for a tray are as follows:

Minimum. As noted above, some liquid must flow over the downcomer at all times. The loss of liquid through weeping to the tray below and entrainment to the tray above is largely offset by liquid gained through weeping from the tray above and entrainment from the tray below. Rarely would one attempt to operate a column at such low liquid flows that these factors would be significant.

Maximum. High liquid rates contribute somewhat to the pressure drop across a tray. This pressure drop occurs primarily at the point where the liquid flows from the bottom of the downcomer onto the tray. But unless

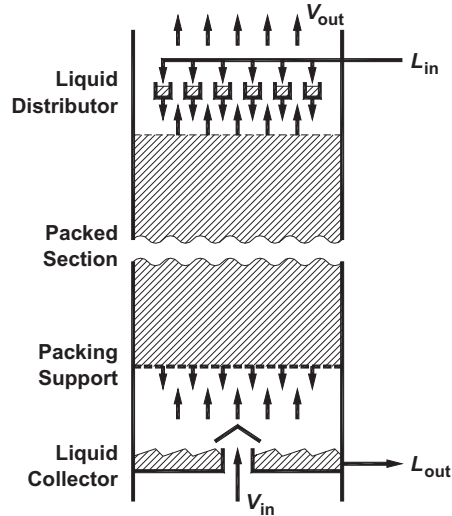


Figure 1.19. Vapor and liquid flows for packing.

this opening is unusually small, this contribution is minor as compared with the contribution from the vapor flow through the openings in the bottom of the tray.

High liquid rates are normally accompanied by high vapor rates; low liquid rates are normally accompanied by low vapor rates. For trays, the limiting conditions are normally attained due to the vapor flow, not due to the liquid flow.

1.11.4. Packing

Figure 1.19 illustrates the flows associated with a packed section. There are two items of equipment associated with each packed section:

Liquid distributor. Before entering a packed section, the liquid flows through a liquid distributor whose function is to distribute the liquid uniformly over the flow area of the packed section. For the upper packed section, the liquid flowing to the liquid distributor is the external reflux. For the lower packed section, the liquid flowing to the liquid distributor is the liquid from the upper packed section plus the liquid from the tower feed.

Liquid collector. The liquid flowing out of the packed section is collected by a liquid collector. For the upper packed section, the liquid from the liquid collector flows to the liquid distributor for the lower packed

section. For the lower packed section, the liquid from the liquid collector is the liquid flow to the reboiler.

Packing comes in a variety of shapes and designs, most of which are proprietary. The objective is to provide the maximum amount of surface area for vapor–liquid contact per unit volume of the tower. A significant advancement was the introduction of structured packing, which permitted significantly greater heights of packed sections. The older packing was referred to as random packing and was basically dumped into the tower.

1.11.5. Vapor Flows for Packing

The limits on the vapor flows for a packed section are as follows:

Minimum. The vapor flow through a packed section can be stopped entirely. For example, start-up usually begins with the liquid flow. The vapor flow remains off until sufficient liquid has been admitted to the tower to completely wet the packing.

Maximum. The packing offers resistance to vapor flow, which results in a pressure differential across the packed section that increases with the square of the vapor flow. If this pressure drop is too large, the consequence is a phenomenon known as “flooding.” The consequences of flooding on a packed section are the same as for trays.

As for tray towers, attention is directed to flooding. Most packed sections are equipped with pressure drop measurements that can detect the onset of flooding, and operations personnel take high pressure drops seriously.

1.11.6. Liquid Flows for Packing

The limits on the liquid flows for a packed section are as follows:

Minimum. If any vapor is flowing through a packed section, the liquid flow must be sufficient to keep the entire surface area of the packing wet with liquid. The consequences of a hot but dry packing surface are always adverse. The exact consequences depend on the nature of the materials being separated. In some cases, the consequence is a residue or buildup on the surface of the packing. For some materials, the dry surface is adversely affected (sometimes referred to as glazing) such that it will not be subsequently wetted by the liquid. The packing designers recommend what liquid flow is required to wet the packing.

Maximum. At high liquid rates, the flow area for the vapor is reduced, thus effectively increasing the pressure drop due to the vapor flow. However, high vapor rates usually accompany high liquid rates, and the adverse

consequences of high vapor rates appear before any adverse consequences of high liquid rates.

For packed towers, logic is required within the controls to maintain an adequate liquid flow to keep the packing wet. However, no control logic is normally required to avoid high liquid flows for a packed section.

1.12. FLOODING

For all towers, the pressure is highest at the bottom of the tower and lowest at the top. This pressure drop is primarily a function of the vapor flow (actually vapor velocity); the contribution from the liquid flow is small.

Most towers are equipped with one or more differential pressure measurements, the options being the following:

Across the entire tower. Despite what some illustrations imply, the upper connection (the low pressure connection) is not always physically at the top of the tower. When the condenser and reflux drum are physically located at grade level, the low pressure connection is normally in the overhead vapor line. This connection can be at approximately the same physical elevation as the high pressure connection, which minimizes wet leg/dry leg issues that arise in differential pressure measurements.

Across a separation section. For these differential pressure measurements, the two connections will not be at the same elevation, so the wet leg/dry leg issues must be addressed. When the tower internals are the same in both separation sections, flooding initially occurs in the separation section with the highest vapor flow. Only the pressure drop across this separation section is required. But when the vapor flows are significantly different, the tower may have different tower internals or even different tower diameters (which affects the vapor velocity). In such cases, differential pressure measurements across both separation sections are recommended.

1.12.1. Pressures on Trays

Flooding is basically the same phenomenon for both tray towers and packed towers. Most find it easier to understand for trays, hence the illustration of a tray tower in Figure 1.20. The notation is as follows:

P_n = pressure on stage n (cm H₂O);

$\Delta P_n = P_n - P_{n-1}$ = pressure drop across stage n (cm H₂O);

H_n = height of liquid in the downcomer on stage n , relative to the weir (cm);

H_T = tray spacing (height between trays) (cm);

G = specific gravity of the liquid in the tower relative to water.

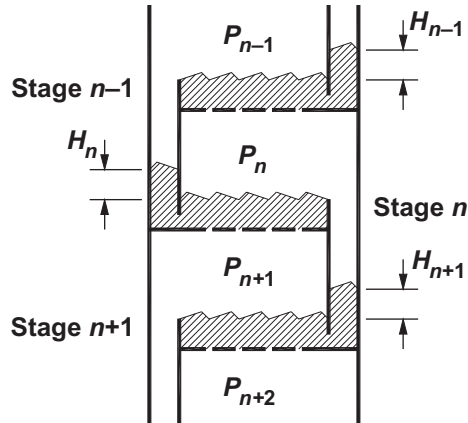


Figure 1.20. Pressures on trays.

The pressure on stage n is greater than the pressure on stage $n - 1$; that is, $\Delta P_n > 0$. In order for the liquid to flow from the downcomer to the tray, sufficient hydrostatic head is required in the downcomer to overcome this pressure differential. Consequently, the height of liquid in the downcomer depends on the pressure drop across the stage.

1.12.2. Hydrostatic Head in Downcomer

The height of liquid in the downcomer is given by the following expression:

$$H_n = G \Delta P_n.$$

The height of the liquid in the downcomer increases linearly with the pressure drop across the tray. But there is a limit on the available height in the downcomer. This limit is determined by the tray spacing; that is, the maximum allowable value for H_n is the tray spacing H_T . Consequently, there is a maximum allowable pressure drop for a stage:

$$\Delta P_n \leq G H_T.$$

If this pressure drop is exceeded, the liquid cannot flow from the downcomer onto the tray. This causes liquid to accumulate on the upper tray, which is said to “flood.”

1.12.3. Contribution of Vapor Flow

The pressure drop across a tray is determined largely by the vapor flow, or more precisely, the velocity of the vapor as it flows through the orifices on the tray. This depends on both design and operational parameters:

Number and size of the openings in the tray. In towers where there are significant changes in the vapor flow between one separation section and another, the number and size of the openings may not be the same in both sections.

Vapor flow. The pressure drop increases with the square of the vapor flow. Logic is required in the control system to keep the vapor flow below its flooding limit.

Pressure. For a given vapor mass or molar flow, reducing the tower pressure increases the vapor velocity. Lowering the pressure in a tower could lead to flooding.

The tray spacing determines the maximum allowable pressure drop across a tray; this relationship is very simple and was presented previously. The maximum allowable pressure drop across a tray determines the maximum allowable vapor flow. However, accurately calculating the vapor flow that corresponds to the maximum allowable pressure drop is not generally possible.

1.12.4. Contribution of Liquid Flow

The hydrostatic head in the downcomer must also overcome any resistance to liquid flow. If the tray is designed properly, this resistance should be very small. The most likely location of any resistance is at the clearance at the bottom of the downcomer, which is illustrated in Figure 1.21.

There are two components to the hydrostatic head in the downcomer:

Vapor contribution. This is essentially $\Delta P_n / G$, with ΔP_n increasing with the square of the vapor velocity.

Liquid contribution. Assuming a proper clearance at the bottom of the downcomer, this component would only be significant at very high liquid flows.

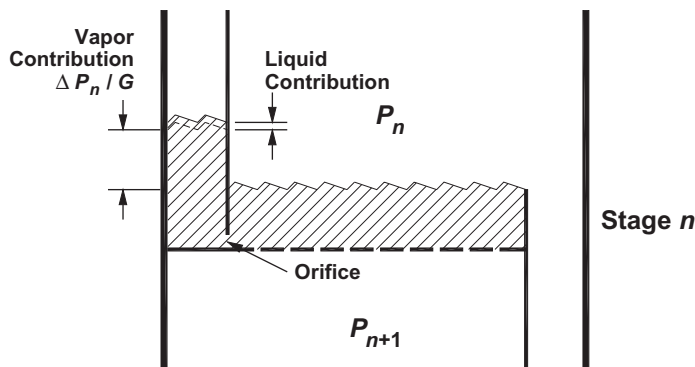


Figure 1.21. Contribution of liquid flow to tray pressure drop.

1.12.5. Maximum Pressure Drop for a Separation Section

Gravity provides the driving force for liquid to flow down a tower. The maximum driving force for liquid flow is the hydrostatic head provided by a column of liquid of the same height as the separation section. The maximum pressure drop for a separation section is

$$\Delta P_{\max} = G H_S,$$

where

- H_S = height of the separation section (cm);
- ΔP_{\max} = pressure drop across the separation section (cm H₂O);
- G = liquid specific gravity relative to water.

It is essential that the pressure drop over a separation section never exceed this pressure drop.

The differential pressure measurement for a separation section or for the entire tower provides the basis for preventing flooding. Sometimes alarms are defined on the differential pressure measurement, and the operators are responsible for taking the appropriate action. However, it is also possible to incorporate logic so that the controls will take the necessary actions to avoid pressure drops that cause flooding.

1.12.6. Separation Sections with Packing

Figure 1.22 illustrates the pressure drop across a packed section. Flooding in a packed tower is basically the same as flooding in a tray tower. Gravity pro-

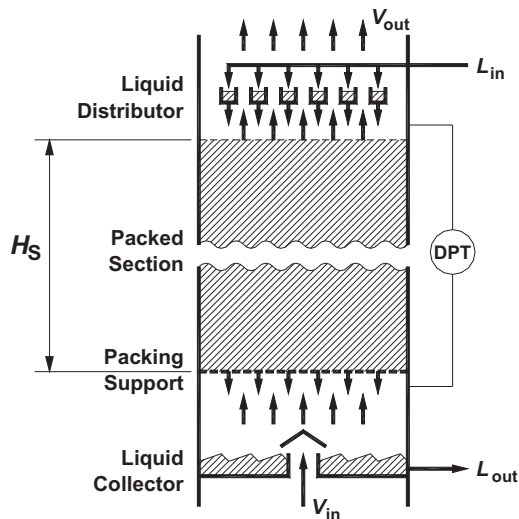


Figure 1.22. Pressure drop across a packed section.

vides the driving force for fluid flow down the tower. The packing provides resistance to vapor flow, which leads to a pressure drop across the separation section. As for tray towers, this pressure drop is proportional to the square of the vapor velocity. Should the pressure drop caused by the vapor flow exceed the hydrostatic head provided by a column of liquid of the same height as the packed section, liquid cannot flow down the packed section. Instead, it accumulates within the tower, which constitutes flooding.

Within the liquid phase in packed towers, there is very little resistance to fluid flow. This resistance would only be significant at very high liquid rates. Therefore, the major contribution to the pressure drop across a packed section is due to the vapor flow.

1.12.7. Issues Pertaining to Flooding

Flooding means the tower is filling with liquid. Consequently, reducing the liquid flow would seem to be an appropriate action to take. However, the tower is filling with liquid because the liquid cannot flow down the tower, not because too much liquid is being fed to the tower. The appropriate response to a flooding situation is to reduce the vapor flow. This is true for both tray and packed towers.

If (1) the tower is properly designed and (2) the tower is operating under the conditions for which it was designed, the limit on tower operations should be imposed by the most expensive component of the tower. In most cases, this is the tower internals. Consequently, encountering the flooding limit during production operations should be expected, and the controls must be configured accordingly.

The optimum operating point is often at a constraint. For a distillation column, this constraint is often associated with flooding. A subsequent chapter considers control configurations that will operate a tower close to the constraint imposed by flooding.

1.13. TRAY HYDRAULICS

Only those aspects of tray hydraulics that are of interest from a control perspective are examined herein. Specifically, the amount of liquid (the holdup) retained on a tray is affected by both the liquid flow and the vapor flow, the manner being as follows:

Liquid flow. The liquid holdup on a tray increases with liquid flow.

Vapor flow. The liquid holdup on a tray decreases with vapor flow.

This section examines the effect of the liquid flow; the next section examines the effect of the vapor flow.

1.13.1. Flow over Weirs

The flow of liquid over weirs has been extensively studied, one result being relationships between the height of liquid over a weir and the flow over the weir. These are routinely used in the water and wastewater industry, but their applicability to trays within a tower could certainly be questioned. Water flows in flumes and channels are relatively calm, whereas the liquid flowing across a tray in a tower is in a violent state of agitation. However, these relationships are used in various analyses, including tray efficiencies and tray dynamics.

The key relationship is the Francis weir formula, which is expressed as follows:

$$f = k w h^{3/2},$$

$$h = \left[\frac{f}{k w} \right]^{2/3},$$

where

h = height of liquid above weir (cm);

w = width of weir (cm);

f = volumetric flow (cc/s);

k = coefficient, 18.4 cm^{3/2}/s (3.33 ft^{3/2}/s).

1.13.2. Effect of Liquid Flow on Tray Holdup

On an increase in the liquid flow onto a tray, the height of the liquid above the weir must increase sufficiently so that the outlet flow is the same as the inlet flow. This means that some of the liquid flowing onto the tray is retained on the tray to cause the height over the weir to increase. On a decrease in liquid flow, the effect is the opposite. The effect is that of a first-order lag with the time constant being the hydraulic time constant τ_h .

The volume of liquid V_h above the weir is the product of the tray area A and the height above the weir h . Substituting the Francis weir formula for h gives a relationship for the effect of the liquid flow f on the volume of liquid above the weir:

$$V_h = A h = A \left[\frac{f}{k w} \right]^{2/3}.$$

The hydraulic time constant τ_h is the rate of change of the volume V of liquid above the weir with respect to the flow f :

$$\tau_h = \frac{dV_h}{df} = \frac{2A}{3 k w} \left[\frac{f}{k w} \right]^{-1/3}.$$

The hydraulic time constant depends on the following parameters:

Tray area A (a design parameter). The larger the tray area A , the larger the hydraulic time constant.

Length of the weir w (a design parameter). The longer the length of the weir w , the smaller the hydraulic time constant.

Flow f (an operating parameter). The hydraulic time constant is largest at low liquid flows.

1.13.3. Dynamic Effect of the Hydraulic Time Constant

The hydraulic time constant is manifested as a lag in the liquid flow within a separation section. For a single tray, the relationship between the outlet liquid flow and the inlet liquid flow is the characteristic first-order lag response. However, a separation section consists of some number of trays in series. The overall behavior of a large number of time constants in series is very similar to the behavior of dead time or transportation lag.

Figure 1.23 presents the response to a step increase in the liquid flow into a separation section that consists of 20 trays. The hydraulic lag on each tray is 6 seconds. The flow from tray 1 is the response of a 6-second lag to a step change in its input. The flow from tray 20 is the result of 20 lags in series, each lag being 6 seconds. This response is closer to the response of a dead time of 120 seconds (20 trays with a 6-second lag on each tray).

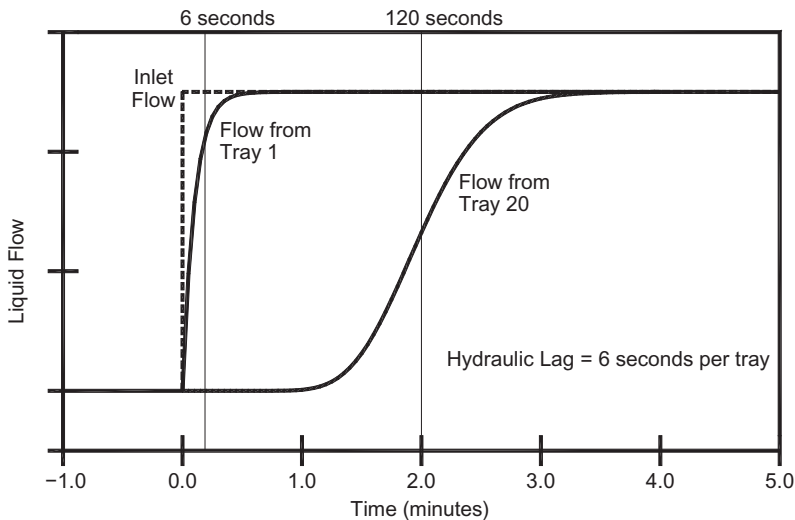


Figure 1.23. Effect of hydraulic lag on an increase in the liquid flow to a separation section.

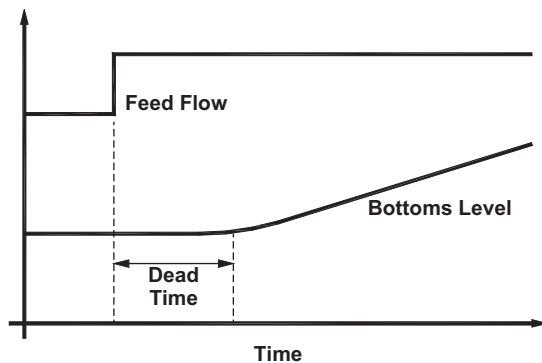


Figure 1.24. Response of bottoms level to increase in feed flow.

The hydraulic time constant is typically in the range of 5–10 seconds, but since internal flows cannot be measured, an accurate value is not generally available.

1.13.4. Response in Reboiler Level

The consequences of the hydraulic time constant are often observed in the response of the bottoms level to changes in either the feed flow or reflux flow. The ensuing discussion assumes all controls are on manual, so the response will be that of the tower alone.

If the feed is mostly liquid, an increase in the feed rate means an increase in the liquid flow to the lower separation section of the tower. This leads to an increase in the liquid flow out of the lower separation section and into the reboiler. With no controls in operation, this causes the reboiler level to increase.

Figure 1.24 illustrates the response of reboiler level to an increase in the feed rate. The reboiler level does increase, but not immediately following the increase in the feed rate. There is a delay, after which the reboiler level increases in the expected manner. The value for the delay, often referred to as dead time, is determined by the number of trays in the lower separation section and the hydraulic time constant of each tray. If the lower separation section has 10 trays and the hydraulic time constant is 6 seconds, the dead time in the reboiler level response is approximately 1 minute.

1.14. INVERSE RESPONSE IN BOTTOMS LEVEL

On an increase in the heat input to a reboiler, the bottoms level should decrease. Indeed, this is always the long-term effect; however, the short-term effect can be that illustrated in Figure 1.25. The bottoms level initially increases,

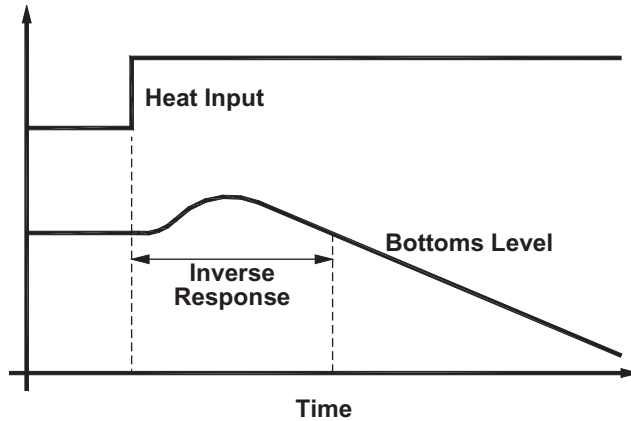


Figure 1.25. Inverse response in bottoms level.

but eventually decreases as expected. This type of behavior is known as “inverse response.”

Towers exhibit inverse response to varying degrees:

- Some exhibit little or none. Given the noise generally present in level measurements associated with a boiling liquid, a small amount of inverse response would be difficult to detect. Any smoothing on the level measurement would also obscure the inverse response.
- Some exhibit what appears to be dead time; following the increase in the heat input, some time elapses before the bottoms level begins to drop.
- Some exhibit inverse response to a minor degree.
- Some exhibit inverse response to a very noticeable degree. Such a tower was reported by Buckley et al. [2].

Inverse responses can be extreme. However, few, if any, columns exhibit inverse response to this degree.

1.14.1. Effect of Vapor Flow on Tray Holdup

Vapor enters the tray through small openings and is dispersed into the liquid on the tray. If the liquid on the tray is in a quiescent state as illustrated in Figure 1.26, the volume of liquid on the tray is the tray volume (tray area times weir height) less the volume of liquid displaced by the vapor bubbles.

What happens when the vapor flow increases? More vapor bubbles are dispersed into the liquid, displacing a greater volume of liquid. The trays are said to “dump liquid.” In the tower, an increase in vapor flow is felt on all trays in the tower. Therefore, each tray is “dumping liquid” into the downcomer and

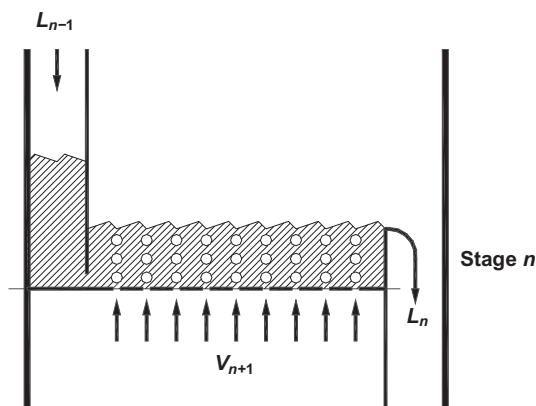


Figure 1.26. Effect of vapor flow on tray holdup.

onto the tray below. All of this liquid eventually ends up in the tower bottoms, resulting in an increase in bottoms level.

This can lead to an interesting sequence of events. If one increases the heat input to the reboiler, the expectation would be for the vapor flow to increase and the bottoms level to drop. This is indeed the long-term effect. However, the short-term effect is potentially quite different. The increase in vapor flow causes the trays to dump liquid, which causes the bottoms level to increase. The initial response is in a direction opposite of the long-term response, hence resulting in what is known as an “inverse response.”

In columns that exhibit dead time in the response of bottoms level to an increase in heat input, there is no true transportation lag. But during what appears to be the dead time, the decrease in bottoms level due to the increased boilup is basically offset by the liquid being dumped by the trays.

1.14.2. Mean “Liquid” Density

The liquid on a tray is in a rather violent state of agitation caused by the dispersion of vapor bubbles into the liquid. The concept of a quiescent pool of liquid on a tray is not accurate (nor is the concept of liquid calmly flowing over a weir).

Consider the vapor–liquid mixture on the tray to be the “liquid phase.” This mixture generally extends well above the weir height, but does not extend to the tray above (should it extend to the next tray up, significant liquid entrainment would occur and the tray efficiency drops dramatically). What is the effect of vapor flow on the mean density of the “liquid phase” (in reality, the vapor–liquid mixture)?

Most information on the effect of the vapor flow on the mean “liquid phase” density comes from tray efficiency studies. At low vapor flows, changes in the

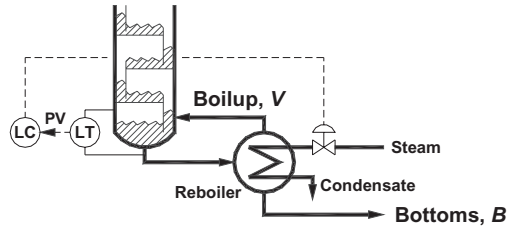


Figure 1.27. Control bottoms level with boilup.

vapor flow affect the mean density and consequently the liquid holdup on the tray. But at high vapor flows, changes in the vapor flow have little effect on the mean density and the liquid holdup. This suggests that towers with low vapor flow rates would exhibit a more pronounced inverse response than towers with high vapor flow rates.

1.14.3. Bottoms Level

If the bottoms level exhibits inverse response, will the level control loop illustrated in Figure 1.27 deliver satisfactory performance? When the level controller increases its output, the expected response is a decrease in bottoms level. But when inverse response is present, the short-term result is an increase in the bottoms level, which causes the controller to further increase its output.

Inverse response always has a negative impact on the performance of a loop, even more than dead time. In the presence of dead times, controller gains must be reduced. Inverse response necessitates even lower values of the controller gain. The inverse response in bottoms level is generally mild to at most moderate. While level control performance suffers, the result is usually acceptable.

1.15. COMPOSITION DYNAMICS

For the same number of stages, the separation provided by a tray tower and by a packed tower is exactly the same. Whether the tower is trays or packed has little impact on the steady-state solution. However, this is not the case for the dynamics.

1.15.1. Vapor and Liquid Dynamics

The vapor dynamics are the same for both tray and packed towers. Vapor dynamics are also the simplest: any change in vapor flow is propagated instantly throughout the tower. That is, a change in the vapor flow from the reboiler is

immediately felt at the top of the tower. As compared with liquid holdups, the vapor holdup in a tower is very small.

The liquid holdup in a separation section of a packed tower is essentially constant provided the following two criteria are met:

- The liquid flow is above that required to wet the packing.
- The pressure drop across the separation section is less than about 80% of the flooding limit.

A constant liquid holdup means that a change in the liquid flow into the separation section immediately appears in the liquid flow out.

The liquid holdup on trays is far more complex. The liquid holdup increases with the liquid flow, but decreases with the vapor flow. Both of these were examined previously, so no further discussion is required.

1.15.2. Composition Dynamics

The composition dynamics are largely determined by the liquid holdups within the tower. Liquid holdups include the following:

Tower internals. The liquid holdup in a tray tower is generally larger than the liquid holdup in a packed tower.

Reflux drum. A few towers do not have a reflux drum. In small vacuum towers, the condenser is often physically located at the top of the tower, with the condensate returned directly to the upper separation section.

Bottoms holdup. This depends on the type of reboiler. For thermosyphons, it is in the bottom of the tower; for kettle reboilers, it is in the reboiler itself. A later section examines various types of reboilers.

A dynamic simulation must encompass all of these. Since the liquid holdup is less in a packed tower, the contribution of the condenser and reboiler to the overall tower dynamics is larger for a packed tower than for a tray tower.

1.15.3. Stage Dynamics

The simplest approach to simulating stage dynamics is to assume that the liquid within the stage is perfectly mixed. This assumption is generally made for both packed and tray towers.

For packed towers, this assumption is clearly not correct. Theoretically, there should be no composition gradients in the horizontal direction. However, there is little vertical mixing in the liquid within a packed tower, so vertical gradients are present within the stage. Such composition gradients can be simulated, but do the improved results justify the extra effort? Assuming that

the liquid in a stage is perfectly mixed permits the simulation program for a tray tower to be also used for a packed tower.

In developing the equations for the steady-state simulation, the liquid on a tray is assumed to be perfectly mixed. Would the liquid on a tray in a 10-m-diameter tower be perfectly mixed? Of course not. For the steady-state simulations, the tray efficiency compensates for the errors. For the dynamic simulation, assuming the tray consists of two perfectly mixed sections would give slightly different results than assuming the entire tray is perfectly mixed. However, the results are not drastically different. Dynamic simulations can be developed for any assumption regarding the mixing on the tray, but what assumption should be made? Until this question can be answered, assuming the entire tray is perfectly mixed will continue.

1.15.4. Dynamic Simulations

Before undertaking a dynamic simulation, a steady-state simulation is a must. But a dynamic simulation requires many additional parameters. For example, stage holdups and reflux drum capacity have no effect on a steady-state simulation, but must be known for a dynamic simulation. The additional parameters include the following:

Flows. Steady-state simulations can be done on the basis of a feed rate of 100 mol/h. However, dynamic simulations require actual flow rates.

Tower size. The dynamic simulation requires the number of actual stages, the column diameter, column height, and so on.

Tower internals. To determine the liquid holdup for each stage, the nature of the separation sections must be known.

Condenser. At a minimum, the capacity of the reflux drum must be known. The type of condenser must be known, and for some, the size must also be known.

Reboiler. The type of reboiler must be known. The capacity of the liquid holdup in the bottoms must be known, and for some types of reboilers (e.g., kettle reboilers), their size must also be known.

Programs to simulate column dynamics are widely available. However, much effort is required to obtain the additional parameters.

1.15.5. Simulation Detail

The degree of detail for a simulation must be consistent with the intended use of the simulation. The greater the degree of detail, the greater the number of parameters that will be required.

The simplest situation is when the primary requirement is to simulate the composition or temperature dynamics. The purpose for such a simulation may

be to verify that the temperature or composition control configuration will function properly. The composition and temperature loops are the slowest loops. Such simulations can be simplified by assuming that the level loops, the flow loops, and the column pressure loop are much faster. Assuming perfect performance from a loop has several attractive consequences:

- A constant value to be used for its controlled variable (the level, flow, or pressure).
- Values of several parameters (measurement device characteristics, control valve characteristics, etc.) are not required.
- The flow through the control valve can be calculated from a steady-state equation, thereby eliminating additional parameters.

Compared with the temperature and composition loops, flow loops will be instantaneous. The column pressure is very likely to be instantaneous (except possibly for a condenser arrangement known as the “flooded” condenser).

Assuming constant level deserves somewhat more attention. Using the reflux drum as the example, there are two possibilities:

Control level by adjusting the distillate flow. With perfect level control, any change in the overhead vapor flow is immediately translated to a change in the distillate flow. As the distillate flow is to external equipment, errors in this assumption would not significantly affect the results of the dynamic simulation.

Control level by adjusting the reflux flow. With perfect level control, any change in the overhead vapor flow is immediately translated to a change in the reflux flow. As the reflux flow is returned to the tower, significant errors in this assumption would affect the results of the dynamic simulation.

Similar issues arise for the reboiler.

1.15.6. Dynamic Simulation as Part of Design

Most process designs are based largely on steady-state relationships. Arguments have been advanced for dynamic simulations to become an integral part of process design. The progress, if any, has been slow.

Driven by financial reasons, shortening the design and construction cycle is of major interest. One consequence is that major items of equipment must be ordered as early as possible, often before all aspects of the design are finalized. Distillation columns are usually major items of equipment.

One claim for dynamic simulation is that it will uncover flaws in the design. Experience seems to support this claim. But to take advantage of this for a major item of equipment, placing the order must be delayed to provide the

time to make the simulation. But this is counter to the objective of shortening the design cycle. Doing the simulation in parallel is an option. However, the simulation must be delayed until detailed designs are available for the condenser, the reboiler, and the tower internals. Undertaking the simulation at that time is too late; by the time the results are available, the equipment fabrication and plant construction are too far along.

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- 2 Buckley, P. S., R. K. Cox, and D. L. Rollins, Inverse response in a distillation column, *Chemical Engineering Progress*, 71(6), June 1975, 83–84.