Strategy for Distillation-Column Control

In chemical plants and petroleum refineries, there are, today, many distillation columns that are working well. There are also many others that are not working well, and at least a few that function very poorly, or not at all. Failure to obtain performance specified by the column design engineer is due, in many cases, to faulty or inadequate control system design. Troubleshooting of columns that are already in operation is frequently necessary, but practical considerations usually limit corrective measures to relatively minor items. Proper original design is by far the best way to guarantee satisfactory operation and control. Therefore, in this book we will approach the design of integrated distillation-column control systems as a systems problem in process design. The application of feedforward, feedback, and protective controls will be coordinated with the sizing and proper location of process holdups to achieve both automatic start-up and shutdown and smooth, noninteracting control of column product compositions.

1.1 DISTILLATION CONTROL OBJECTIVES

The starting point of any design project is a definition of objectives. For distillation there are many possible approaches, but the one chosen here is one the authors have found broadly useful in virtually all kinds of processes. It has three main facets: (1) material-balance control, (2) product quality control, and (3) satisfaction of constraints. As applied specifically to distillation columns, this philosophy suggests the following:

1. Material-balance control*

* This term is sometimes used by others to mean a control system in which reflux is set by reflux drum level control, and distillate/feed ratio is set manually or by a composition (temperature) controller. The authors of this book have been unable to find any special merit for this scheme except for some high reflux ratio columns.
The column control system must cause the average sum of the product streams to be exactly equal to the average feed rate. Harbert has called this requirement that of keeping the column in "balance."

The resulting adjustments in process flows must be smooth and gradual to avoid upsetting either the column or downstream process equipment fed by the column.

Column holdup and overhead and bottoms inventories should be maintained between maximum and minimum limits.

It is important to note that the material-balance controls on any given column must be consistent with the material-balance controls on adjacent process equipment. In most cases material balance will be controlled by so-called "averaging" liquid-level or pressure controls.

2. Product quality control

The control system for a binary distillation in most cases must:

- Maintain the concentration of one component in either the overhead or bottoms at a specified value.
- Maintain the composition at the other end of the column as close as possible to a desired composition.†

It is usually true that minimum operating cost is achieved when the products are controlled at minimum acceptable purities. This is so because the relationship between thermodynamic work of separation and purity is nonlinear. For some columns compositions are allowed to vary at one end, and sometimes both ends, to satisfy certain economic constraints.

Both material-balance and composition controls must function satisfactorily in the face of possible disturbances in:

- Feed flow rate
- Feed composition
- Feed thermal condition
- Steam supply pressure
- Cooling-water supply temperature
- Cooling-water header pressure
- Ambient temperature, such as that caused by rainstorms

3. Satisfaction of constraints

For safe, satisfactory operation of the column, certain constraints must be observed. For example:

- The column shall not flood.
- Column pressure drop should be high enough to maintain effective column operation, that is, to prevent serious weeping or dumping.

† For multicomponent columns subjected to feed composition changes, it is not possible to hold exactly constant the compositions at both ends of the column; the composition at one end must change a little. If feed variations are in the nonkey components, composition may vary somewhat at both ends of the column. With only two drawoffs, we can control two keys, but not the nonkey components.
- The temperature difference in the reboiler should not exceed the critical temperature difference.
- Column feed rate should not be so high as to overload reboiler or condenser heat-transfer capacity.
- Boilup should not be so high that an increase will cause a decrease in product purity at the top of the column.
- Column pressure should not exceed a maximum permissible value.

There are, in addition, several other facets of column control.

**Startup and Shutdown**

Column controls should facilitate startup and shutdown and, by implication, should make it easy to achieve total reflux operation when desired.

**Transitions**

When it is desired to change product compositions, the column controls should facilitate doing so. This is particularly important when feed stock composition varies widely and it is desired to optimize column or train operation, as, for example, with a computer.

**Heat Recovery**

Increasingly there is an interest in recovering as much heat as possible. The petroleum industry has frequently used the sensible heat in a column bottom product to preheat column feed. Recently more ambitious schemes have been employed in which the reboiler for one column is used as the condenser for another. Such schemes magnify control problems and sometimes limit process turndown.

**Testing**

There should be enough instrumentation so that testing may be carried out for tray efficiency, heat and material balances, flooding, and so on.

**Miscellaneous**

In addition to the above, column control should be designed with human engineering in mind. For example:

- The operator's work station, whether a cathode-ray-tube/keyboard console or a panelboard with gages, switches, recorders, and so on, should be carefully designed according to human engineering principles for easy use.
- The controls should be so designed as to require minimum maintenance. The need for frequent or critical "tuning" should be avoided. The hardware should be designed and arranged for convenient access and quick repair or replacement.
- The control system design engineer should use the simplest possible design procedures, not only to hold design costs down, but also so that the work can be readily discussed with other design and plant personnel. This will facilitate, at a later date, any necessary minor redesign at the plant site. Failure
of design and plant personnel to achieve a mutual understanding of design objectives, concepts, and methods is one of the frequent causes of unsatisfactory plant operation.

1.2 ARRANGEMENTS FOR MATERIAL-BALANCE CONTROL

As shown elsewhere, the size and location of tanks and the concept of overall material-balance control used can have a great influence on plant investment and process control. If the design engineer uses the concept of control in the direction opposite to flow, tanks may be smaller and plant fixed investment and working capital can be lower than if control in the direction of flow is used. The meaning of these expressions is illustrated, for simple tanks with level controllers, in Figures 1.1 and 1.2.

For the last tank in a series (final product storage), the demand flow is always the shipments to a customer. As shown by Figures 1.3 and 1.4, the choice in control strategy is between adjusting the flow into the last tank (control in direction opposite to flow) or adjusting flow into the first process step (control in direction of flow). In the first case, we can easily use simple automatic controls. In the second case, it is more common to have an operator make the adjustment.

When more than one tank is involved, other advantages of control in the direction opposite to flow are (1) less difficulty with stability problems, and (2) reduced internal turndown requirements. "Turndown," as used here, is the ratio of maximum required flow rate to minimum required flow rate. In this instance the meaning is that, in response to a given change in demand flow, the required change in the manipulated flows will be smaller in one case than in the other.

Once the basic concept of material-balance control has been selected for a process, one must apply the same concept to all process steps. It is for this reason that the first step in designing column controls is to determine the material-balance control arrangement. Control in the direction of flow is the most commonly used concept (although the least desirable), and a frequently encountered arrangement is shown on Figure 1.5. Here level in the condensate receiver (also commonly called reflux drum or accumulator) sets the top product, or distillate flow, while the level in the base of the column sets the bottom product flow; in other columns base level sets steam or other heat-transfer media to the reboiler, in which case the condensate receiver level sets top product flow.

Generally speaking the direction of material-balance control is determined by the demand stream. In recycle systems we may find some material-balance controls in the direction of flow while others are in the direction opposite to flow.

Material-balance control, in the direction opposite to flow, can lead to many interesting level-control and flow-ratio options. These are discussed in detail in Chapter 6.
1.2 Arrangements for Material-Balance Control

FIGURE 1.1
Material balance control in direction opposite to flow

FIGURE 1.2
Material balance control in direction of flow
FIGURE 1.3
Overall material balance control in direction opposite to flow
FIGURE 1.4
Overall material balance control with intermediate material balance control in direction of flow
FIGURE 1.5
Distillation column with material balance control in direction of flow
1.3 **FUNDAMENTALS OF COMPOSITION CONTROL**

Let us consider briefly what must be done to a column to keep terminal compositions constant on a steady-state basis when the column is subjected to sustained changes in feed flow rate or feed composition. Methods of handling other disturbances will be discussed later.

To simplify the analysis, let us limit our attention to an ideal, binary distillation. This is somewhat restrictive, although the results will be applicable in a general way to multicomponent systems, particularly those that may be treated as quasi-binary or pseudobinary systems.

As will be shown in Chapter 2, if feed composition and feed thermal condition are constant, then we want the "operating lines" on a McCabe-Thiele diagram to remain constant when feed flow changes. The operating lines (defined in the next chapter) will not change as long as the distillate-to-feed, reflux-to-feed, boilup-to-feed, and bottom-product-to-feed ratios are held constant. Practically speaking one may hold all four ratios constant by fixing any one of the three pairs: (1) the reflux-to-feed ratio and the boilup-to-feed ratio, (2) the reflux-to-feed ratio and the bottom-product-to-feed ratio, and (3) the boilup-to-feed ratio and the distillate-to-feed ratio. In considering case (1), for example, we see that if the rectification section vapor-to-feed ratio is fixed (and it will be if the boilup-to-feed ratio is fixed) and the reflux-to-feed ratio is fixed, then the distillate-to-feed ratio will be fixed, since the distillate flow is the difference between the reflux flow and the vapor flow. Similarly, the bottom-product-to-feed ratio will be fixed, since the bottom product flow is the difference between the stripping section liquid and the boilup.

If feed rate and feed composition are not constant, Rippin and Lamb have shown that, for small perturbations, one should change the boilup and reflux according to the following equations:

\[
\Delta V = K_{f1} \Delta z_F + K_{f2} \Delta F \\
\Delta L_R = K_{f3} \Delta z_F + K_{f4} \Delta F
\]

where

- \( V \) = vapor flow from the reboiler
- \( L_R \) = internal reflux flow at the top of the column

The \( \Delta \)'s represent departures from average operating conditions. The constants \( K_{f1} \cdots K_{f4} \) may be calculated approximately by the column designer. Luyben has shown that it is necessary to be quite careful in designing feedforward compensation for feed composition changes, particularly when the column is not making a sharp separation.

Visualization of column operation, in terms of reflux-to-feed and boilup-to-feed ratios, was suggested by Uitti and has since been proposed in varying degrees by many others. Throughout the rest of this book, it will be used as the primary basis for column composition control.
It should be noted, however, that this "feedforward" approach to column control has a particular limitation: In general one cannot calculate the constants $K_{f1}$ ... $K_{f4}$ with great accuracy. For columns that are not operating too close to either upper or lower limits of capacity, small changes in feed rate, and consequent changes in boilup and reflux, will not change tray efficiency appreciably. The terms $K_{f2}$ and $K_{f4}$ therefore will be constants. If the feed composition changes are not too large (as will usually be the case), then $K_{f1}$ and $K_{f3}$ may also be treated as constants. To determine the control accuracy obtainable by this approach, one should make the necessary calculations or tests for each individual column. Where really close control is required, one must supplement feedforward control with measurement of the column terminal compositions and subsequent feedback control, at least at one end of the column. The usual philosophy will be to use feedforward for fast, approximate control and feedback for long-term, accurate control of composition.

It should be noted, too, that feedforward from feed composition may not be needed if the feed comes from a process step with discharge composition control. Feedforward compensation for other process variables, such as bottom product or distillate demand flow, is discussed in Chapter 6.

As will be shown later (Chapters 5 and 20), a properly designed column feed system can play a very important role in filtering our disturbances in feed rate, feed composition, and feed enthalpy, thereby making composition control much easier.

1.4 COMPENSATION FOR VARIOUS DISTURBANCES

Feed Thermal Condition

Feed should enter the column with a constant enthalpy. When significant changes are anticipated, a heat exchanger and feed-enthalpy control system should be provided. This is discussed in more detail in Chapters 5 and 11.

Steam Supply Pressure

Changes here can cause changes in boiling rate. The best preventive is a steam-to-feed ratio control system combined with temperature and pressure compensation of steam flow. A high pressure drop across the steam valve favors smooth control but velocity-limiting trim may be required to minimize noise and plug and seat wear. A high pressure drop is also undesirable for energy conservation. If the pressure drop is high enough, sonic flow through the valve results, and reboiler steam-side pressure has no effect on flow rate. The design should avoid having sonic flow corresponding to low feed rates and nonsonic flow corresponding to high feed rates, since required controller gain changes make tuning very difficult.
Cooling-Water Supply Temperature

Cooling-water temperature changes are usually seasonal, and will require no specific correction. If, for some reason, they are large and rapid, then it may be desirable to provide an enthalpy control system for the condenser. By measuring the temperature rise of the cooling water across the condenser, and multiplying it by the cooling-water flow rate, one has a measure of the heat transferred, $q_c$. This calculated value of $q_c$ can serve as the measured variable in an enthalpy control system. For column pressure control, the enthalpy control system can serve as the secondary loop in a cascade system.

Cooling-Water Header Pressure

One of the best ways to make a flow system immune to pressure changes is to provide a high system pressure drop. This, however, is costly. Another way, which should be satisfactory for some distillation columns, is to use a cooling-water flow control system. It does, however, have limitations, which will be discussed later.

Ambient Temperature

If the column, auxiliaries, and piping are properly insulated, and if the column is properly controlled, ambient changes should cause little difficulty unless the condenser is of the air-cooled type. In this event it may be desirable to use an internal reflux computer, discussed in Chapter 11. If, as is often the case, the vapor piping from the top of the column to the condenser is both uninsulated and long, ambient temperature changes may cause fluctuations in pressure and the rate of condensation.

1.5 STARTUP AND SHUTDOWN

Startup and shutdown are often dismissed as relatively unimportant, since they happen so seldom that it is not economical to spend much time and money on improvements. This may be true in a petroleum refinery where shutdowns may occur at intervals of two or three years. In the chemical industry, however, where process equipment is often plagued by severe corrosion or by plugging process materials, startups and shutdowns are far more common—perhaps monthly, weekly, or even daily. Further, elaborate heat and material recycle schemes may require intricate startup/shutdown procedures as part of the original design. To put it another way, columns and their control systems may have to be designed specifically to accommodate a particular startup/shutdown procedure.

Columns are commonly started up in total reflux—no product is taken off at top or bottom. Limited experience, however, suggests that faster and smoother
startups may be achieved by recycling top and bottom products back to feed during part of the startup sequence.

Startup/shutdown will be discussed in more detail in Chapter 9.

1.6 CONTROL SYSTEM DESIGN PHILOSOPHY

Current Design Practices

In the preface we noted that we would try, in this book, to present a multivariable control approach to distillation column control. Before discussing this, however, let us look at typical controls in existing plants.

There is a traditional pattern of what is called “instrumentation” in the chemical and petroleum industries based on single-loop control (sometimes called SISO—single input, single output). Each process operation has a number of independent or single loops for feedback control of temperatures, pressures, flows, liquid levels, and sometimes compositions. The term “single loop” means there is one measurement, one controller, and one final control element, usually a valve. Many, but not all, of these loops are represented in the central control room (CCR) by control stations.

The controllers usually have neither antireset windup nor automatic tracking, and there is little or no logic circuitry to tie the many loops together. This statement is true for both analog and some digital hardware. As a consequence the operators must perform startup operations with the control stations switched to “manual,” and must implement process logic by switching in and out of “automatic.” Since the original design procedures are usually qualitative and intuitive, with heavy emphasis on conformity to past practices, it is not surprising that some loops never work in “automatic.” Others, although stable in “automatic,” are so sluggish that they are ineffective in dealing with typical disturbances. For newer plants with higher throughputs, operation closer to hard constraints, smaller and fewer holdups, energy-recovery systems, and elaborate material-recycle systems, the traditional “instrumentation” approach to control is often seriously inadequate.

For some years now, progress in single-loop design has been essentially at a standstill. As recently as 1950, process control was hardware limited. Since then primary measuring elements, controllers, computing devices, control stations, and control valves have improved greatly in reliability, sensitivity, and speed of response. Consequently these characteristics less and less frequently pose limitations to the single-loop designer. Further, the quality of control achieved by single-loop systems is not greatly affected by type of hardware; it makes little difference, for a given algorithm, whether one uses analog pneumatic or electronic gear, microprocessor controls, or a digital computer.* In addition,

* Digital computers and microprocessor controls, however, usually offer a wider range of controller parameter adjustments and facilitate the design of control systems more sophisticated than most of those discussed in this book.
optimized tuning procedures for unaided feedback controllers have limited practical value for continuous processes; they yield results that are far inferior to those obtainable with well-damped feedback controls with simple feedforward and override control.\(^8\)

For plants being designed today (late 1983), it is increasingly common to use microprocessor controls instead of analog (see discussion under “Hardware Conventions and Considerations”).

**Multivariable Control**

To avoid the limitations of single-loop design and to provide a more flexible and sophisticated process operating logic than can be implemented by human operators, we use an approach we call multivariable control.\(^9\) Many definitions of this term may be found in the literature, but most of them are expressed in mathematical terms rather than in terms of process functions. For our purposes we define a multivariable control system as one that has the built-in intelligence to look simultaneously at two or more process variables and to choose, in a given situation, the best of several preprogrammed strategies (algorithms) for manipulating one or more control valves (or other final control elements).

For example, the steam valve for a distillation column reboiler, depending on circumstances, may respond to controllers for:

- Steam flow rate
- Column \(\Delta P\)
- Column pressure
- Base temperature
- Column feed rate
- Column base level
- Column bottom-product rate

The seven variables listed may also exert control on five or six other valves.

To provide automatic control of this sort, we make extensive use of “variable configuration” controls that are usually implemented by overrides. If, for instance, base composition is normally controlled by steam flow that can be taken over or overridden by high column \(\Delta P\), this is a variable configuration. If base level control is normally achieved by a PI controller that can be overridden by high or low base level proportional-only controls, we call that “variable structure.” Multivariable control may involve both variable configuration and variable structure controls. Hardware permits us to automate this kind of control with a speed, precision, and reliability that are completely beyond the capabilities of human operators.

It is common to think of process control functions stacked one above another in a pyramid or hierarchic arrangement as with traditional business or military organization structures, particularly when computers are involved. Multivariable control structures, however, with their extensive lateral and diagonal crossovers, are functionally more like the modern “matrix” concept of management. From
a control engineering standpoint, the shorter lines of communication and de-
centralized control functions permit more rapid and stable control, and more
reliable, troublefree operation.

By now it is probably apparent that we are striving for control system
designs whose performance and design parameters are specified in advance of
plant startup. In practice we furnish calibration data for controller parameters
and computational devices for the majority of control loops prior to startup. We
calculate these from simulations or simple linear models. For microprocessor
computer controls, we calculate scaling parameters for computation blocks
(either in software or hardware). Our design procedures are accurate enough
that only a modest amount of empirical controller tuning is required at startup.

**Stability, Speed of Response, and Interactions**

Most existing literature on automatic control is concerned with the stability
and speed of response of single loops. The traditional objectives of feedback
control system design are:

1. To get the fastest possible response to set point changes.
2. To compensate for or to attenuate disturbances as much and as quickly
   as possible. These must be accomplished with a reasonable degree of closed-
   loop stability.*

In process control the objectives are often quite different. The objectives
of averaging level control, for example, clearly are different from those just
mentioned. A typical chemical plant or refinery has hundreds of single loops
with many interactions among them. It is usually far more important to design
for a dynamic balance among these loops with a minimum of interaction than
to strive for maximum speed of response. Further, it is usually undesirable
to make rapid changes in manipulated variables since these may upset the process.
For example, distillation column reflux flow and boilup should not be changed
too rapidly since these might cause transient flooding or weeping in part of
the column. Our preferred philosophy of controller tuning is discussed in a
book¹ and a paper."¹

There are five simple methods by which to make a system noninteracting:

1. Design material-balance control loops to be at least a factor of 10 slower
   than related composition control loops. Similarly, in cascade systems, make the
   secondary or slave loop at least a factor of 10 faster than the master loop.
2. Avoid designs that are intrinsically interacting, such as pressure control
   and flow control at the same point in a pipeline. One of the two controllers
   must be detuned.
3. Select process designs that eliminate or minimize interaction. For example,

* Some processes are open-loop unstable, which means that controls must be added and
must be kept in "automatic" for stable operation.
if a tubular reactor is fed at several points along its length from a common header, flow-rate interactions may be reduced to an arbitrary level by providing a very high pressure drop across each feed valve in comparison with the reactor pressure drop.

4. Use override circuits. Although not specifically intended for this purpose, override circuits, by permitting only one controller at a time to control a given valve, eliminate interactions.

5. Use interaction compensators (decouplers). If two control loops, such as top and bottom temperature controls on a distillation column, interact, we can eliminate the interactions by installing two compensators. One compensates for the action of the top temperature controller on the bottom loop while the other compensates for the action of the bottom temperature control loop on the top one. This is discussed in Chapter 20.

In addition, there are some very sophisticated mathematical methods for dealing with interactions. Some are intended for noninteracting design while others seek a design that provides an optimum amount of interaction.

**Hardware Conventions and Considerations***

For control loops represented in the CCR (central control room), it is normal practice (as mentioned earlier) to furnish “control stations.” These may be analog pneumatic, analog electronic, or microprocessor based. In the last case, the station may be physically distinct, like an analog station, or may be represented on a CRT display as a “faceplate.” Each provides an indication of the process variable (flows, level, temperature, etc.), the desired value (set point), and the valve loading signal (controller output, really). There is also a “manual-automatic” switch, which some vendors label “hand-automatic.” In the “manual” mode, the feedback controller is disconnected and there is a knob that enables the operator remotely to set the valve position. This may or may not be subject to restrictions imposed by feedforward compensation, overrides, and so on, depending on the design philosophy for a particular project.

If cascade control is involved, as, for example, liquid level control cascaded to flow control, the secondary station not only has manual-automatic switching, but also another function—“remote-local.” In the “remote” position, the secondary controller set point comes from the output of the primary controller. In the “local” position, valve position may be set manually or the controller set point may be set by the operator (“local-auto”). Although cascade functions are sometimes combined into one station (or CRT “faceplate”) for space and money-saving reasons, we recommend dual stations. Most single-station designs with which we are familiar are very inflexible and complicated; they do not permit ready implementation of feedforward, overrides, and so forth.

*Much of this section may seem superfluous to instrument and plant personnel. Our experience, however, with process design engineers, column designers, and so forth is that this discussion may be very helpful. *
As of this writing (late 1983), the two biggest hardware needs appear to be:

1. Better measurements, especially of compositions.
2. Better control valves. With regard to these, more progress has been made in the design of valve bodies and trim than in the design of actuators. Valve positioners should always be used (it is assumed in this book that they will be).* Piston actuators with double-acting, two-stage positioners are recommended.

As far as CCR hardware is concerned, we have a decided preference for microprocessor controls. They are technically more versatile and are less expensive (some versions) than analog. As of late 1983, many are featuring satisfactory antireset windup and override capabilities. In addition, they provide more advanced logic capability, dead-time simulation, and adaptive tuning. Some of the last named achieve self-tuning via stochastic techniques or by pattern recognition. Others have gain scheduling, where reset time and proportional gain are functions of some process variable or the controller error signal.

Microprocessor controls usually have a sampling time of a fraction of a second. Although slightly slower than analog controls, their performance can generally be approximated by analog control algorithms.

Other advantages include freedom from drift, and the fact that they can be calibrated more precisely, can be reconfigured or restructured without wiring changes, have a larger range of tuning parameters, and contain more control algorithms.

For most projects today, it is possible to find worthwhile applications for a supervisory digital computer with a good data historian, regardless of the type of basic controls selected (pneumatic, electronic, or microprocessor). Digital readouts for important variables are worthwhile because they permit seeing their magnitudes with sensitivity approaching that of the original analog measurements. Most typical analog measurements have a sensitivity ranging from one part per 1000 to one part per 10,000. Most analog readout devices, however, are limited to 0.5–2 percent.

For maximum advantage a supervisory computer should be programmed to have the control algorithms discussed in Chapter 12. These are position rather than velocity algorithms. It is our opinion that using such a computer to imitate unenhanced two- and three-mode analog controls is poor practice. Some worthwhile applications for computers will be discussed later.

Computer consoles were originally provided in the CCR for the convenience of the operators. Sometimes a consolidated console is also provided for production supervision. Engineers’ consoles, perhaps at another location, facilitate technical studies. Separate consoles for maintenance personnel (the computer can be a powerful maintenance tool) are highly desirable.

* Most control valves today are either single-seated global types or rotary types. For both there are substantial, coercive stem forces when the valves are in flowing streams. Positioners compensate for this and maintain the valve’s inherent flow characteristic expressed as a function of controller output signal. It should be noted, however, that some users prefer not to use positioners.
1.7 PROCEDURE FOR OVERALL CONTROL SYSTEM DESIGN

We now have enough information to suggest a sequence of design steps to follow that will lead to a quantitative definition of column and hardware performance.

1. After careful discussion with the process engineer and column designer, and after careful review of the overall process flow sheet, prepare a simplified flow sheet that defines the control concepts:
   a. Select the overall material-balance control scheme first, preferably proceeding from final product inventory to raw material inventory. All individual equipment-piece material-balance controls must be consistent with this scheme.
   b. Select composition control schemes.
   c. Add feedforward and interaction compensators as required.
   d. Add protective controls and automatic startup/shutdown circuits.
   e. Add miscellaneous temperature and pressure controls.
2. Prepare material-balance and composition-control signal flow diagrams.
3. Determine holdup volumes required for smooth material-balance control and for liquid-level override controls at each end of the column.
4. With holdups determined calculate column-composition transfer functions.
5. Select measurement spans and calculate control valve sizes.
6. Calculate feedforward and interaction compensators.
7. Calculate all other overrides.
8. Calculate feedback-controller gain and reset settings and control-loop natural frequencies. Check feed-tank material balance and mixing time constants for adequacy.
9. Use simulation for some columns, particularly those in critical service or with a new untested control system or process configuration. Simulation of the column and its control system will be useful in confirming control concepts and controller tuning parameters. It may also save startup time.

Hardware vendors may now be selected and the measurement and control equipment may be ordered.

Part II of this book deals with the qualitative and heuristic aspects of steps 1 and 3. Quantitative information for the other steps is presented in Part III.

1.8 COLUMN DESIGN PHILOSOPHY AND CONTROL SYSTEM DESIGN

Experience on many projects shows that even small, simple columns benefit from a modest application of feedforward compensation and overrides. This is due to the trend toward increasingly tight column design. The number of trays is held down because smaller allowances are made for uncertainty in tray efficiency. Column diameter and tray spacing are now kept to smaller values, with the
result that columns typically operate closer to flooding. The combined effect of these design policies is to make columns much touchier and harder to control.

Experience indicates that typical incremental instrument investment over that required for unenhanced feedback controls is 5–10 percent (large projects tend toward the lower figure). This incremental investment not only provides better normal control, but it also helps to avoid inadvertent shutdowns. It is, therefore, our opinion that these controls should be used to some extent on almost every column.

Suppose, however, that the customer insists on minimum application of feedback controls with no feedforward compensation or overrides. What column design philosophy should be followed? Having had considerable adverse experience with columns with primitive controls, particularly sidestream drawoff columns, and columns with heat recovery schemes, we suggest the following:

1. Design for normal operation at 60 percent of the rate for flooding.*
2. Provide five extra trays or increase the number of trays by 10 percent, whichever is larger.
3. Provide increased tray spacing.
4. Provide larger condensers. If water cooled, use tempered water.
5. Provide larger reboilers (lower heat flux).
6. Control column ΔP by boilup, or flow control steam.
7. Provide 100 percent reserve capacity in heat-recovery schemes or avoid them altogether.
8. Avoid sidestream drawoff designs.
9. Provide surge tanks between columns with at least 30–60 minutes of holdup each.

1.9 EXISTING COLUMNS—TYPICAL PRACTICES AND TROUBLESHOOTING

Although this book deals primarily with distillation control in design projects, it is pertinent to consider briefly the controls of typical, existing columns, the opportunities for their improvement, and how to troubleshoot them when necessary. Frequently encountered problems include unstable or ineffective controls, off-specification product or products, and flooding or dumping. In addition, it is fairly common practice to use excessive boilup and reflux to make sure of meeting or exceeding product specifications. This not only wastes energy; it also reduces column capacity. To provide a perspective on energy savings, one may note that 100 lbm/hr of steam is worth $3200 per year (basis $4.00/1000 pounds, 8000 hours per year). To save this amount of steam would probably be only a modest accomplishment for most columns.

If composition control of each product stream is desired (and this is usually

* In some companies columns are designed by the probabilistic methods recommended by Fractionation Research, Incorporated.
the most obvious deficiency of most existing columns is the lack of appropriate composition measurements. Most commonly temperature in the upper or lower section of the column (or both) is used in lieu of true composition measurements. Frequently composition control is attempted at only one end of the column, and sometimes at neither end.

Another shortcoming frequently observed is the use of fixed flow controls for steam, reflux, or product drawoff. Any such unaided flow control should be regarded with suspicion. With rare exceptions flow-control set points must be changed to accomplish either composition control or material balance control.

In view of the preceding comments about problem areas and likely opportunities for improvement of composition control and reduction in energy consumption, the following guidelines are suggested:

1. Make sure that column material-balance controls are properly designed and tuned, and that hardware, especially level and flow transmitters and control valves, is in good working condition. If PI level controllers are used, follow the tuning procedures of Chapter 16; auto overrides or nonlinear controllers should be used. It is usually desirable to cascade level control to flow control, in which case flow measurement should be linear.

2. Provide averaging level control of column feed. Column feed rate should also be held between maximum and minimum limits.

3. Provide a linear flow measurement for steam flow control. Also provide temperature and pressure compensation.

4. If condenser controls are a problem, review the control schemes in Chapter 3.

5. At this point with flows established, smoothed, and in some cases limited, it will probably be possible to see some improvement in composition control, at least for part of the time. For further improvement provide steam-to-feed ratio control, internal reflux-to-feed or distillate-to-feed ratio control, high ΔP override on steam to protect against flooding, and a minimum steam flow limiter to protect against dumping.

6. If composition is measured and is cascade controlled via reflux or boilup, or both, ratio controls should be replaced by impulse feedforward compensation (see Chapter 12) if feed flow turndown is greater than 2:1.

7. For pressurized or vacuum columns, make sure an adequate scheme is provided for maintaining an inert balance.

8. If temperature by itself is not an adequate measure of composition, consider one of the schemes in Chapter 10 for using temperature, pressure, and, in some cases, flow measurements to deduce compositions.

1.10 CONVENTIONS FOLLOWED IN THIS BOOK

Throughout the remainder of this book, we will frequently illustrate control schemes with pneumatic components. This is done for convenience. Pneumatics,
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with few exceptions, features a standard signal span of 3–15 psig. All pneumatic controllers, as far as we know, can achieve antireset windup by the same external reset feedback method. This appears to be the most universally useful method (we use it extensively), and has been adopted by one manufacturer of analog electronic instruments and by several vendors of microprocessor-based distributed controls. Other vendors of electronic analog and digital controls feature a wide variety of techniques; some of these work fairly well, while some are quite inflexible. A brief discussion is presented in Chapter 12.

Units used in this book are those commonly employed in chemical engineering: pounds, feet, degrees Celsius, mols, and so on. To facilitate the calculation of control engineers' "time constants," we have mostly used time units of minutes or seconds [e.g., lbm/sec, lbm/min, (pcu/sec)/°C ft²]. For projects that make partial use of metric or SI units, we found it convenient to convert them to the above units. Generally speaking, we have found no advantage in writing equations with SI units. Instead, in programs for computers or programmable calculators, we mostly write the equations in the older units and add subroutines for going back and forth to metric or SI units.

The most common type of controller used in the chemical and petroleum industries was once called "proportional plus automatic reset," later shortened to "proportional reset." Today it is more common to use "PI," which stands for proportional-integral. It is also becoming common today to speak of controller gain rather than proportional band \((\text{PB} = 100/K_c)\). We will also use "reset time," usually in minutes, rather than its older reciprocal, "repeats per minute."

For drawings in which we are trying to present a perspective of control concepts (configurations), we use very simplified symbols. In drawings where we are trying to illustrate concepts of structure (overrides, feedforward, etc.), we use a more detailed symbolism that we have found useful in our design work.

In pneumatics it is common to refer to most signal-conditioning devices other than controllers as "relays." Included are adders, subtractors, multipliers, and so on. For electronic analog and digital controls, it is more common to use terms such as signal scalers and multipliers.

A table of nomenclature and symbols will be found at the end of this book.

1.11 LITERATURE

For anyone seriously interested in distillation control, two books are highly recommended. The first is an easy-to-read, nontheoretic (as far as control is concerned) work by F. G. Shinskey.\(^1\) The treatment of energy conservation alone is worth the price of the book. The second book is by Rademaker, Rijnsdorp, and Maarleveld.\(^2\) It relies heavily on conventional, single-loop control theory, and explores painstakingly a large number of possible control systems. It also contains an extensive bibliography.
For basic reference books on distillation, we have made much use of those by Van Winkle and King. Others by Holland, Trebyl, and Hengstebeck also have been useful.

For basic books on control, we recommend two written by the authors, one by Harriott, and one by Murrill. For more advanced treatments, we suggest texts by Koppel, Douglas, and Gould. The last contains some perceptive comments about the difficulties of applying advanced control theory developed by electrical and mechanical engineers to chemical processes. Recent books by Ray and by Stephanopolous discuss applications of "modern control theory" in chemical processes. McAvo has addressed the specific subject of interaction analysis.

Nonchemical engineers with no background in distillation may find an introductory text by Nisenfeld and Seemann useful. It is clearly written and easy to read.

REFERENCES