COMMON PLUMBING AND CONTROL ERRORS IN PLANTWIDE FLOWSHEETS

WILLIAM L. LUYBEN
Lehigh University • Bethlehem, PA 19015

Almost all senior design courses discuss only the steady-state economic aspects of process design and exclude any consideration of dynamic behavior. Very few design textbooks even mention dynamics and control.[1,2] Given this tendency, the senior design course at Lehigh University is apparently quite distinctive in that it emphasizes "simultaneous design," *i.e.,* the consideration of both steady-state economics *and* dynamic controllability at the early stages of conceptual design. A detailed discussion of the need for and the importance of this simultaneous approach has been presented in a recent book.[3]

The Lehigh design course requires two semesters. In the fall, traditional steady-state synthesis covers steady-state computer flowsheet simulation, engineering economics, equipment sizing, reactor selection, energy systems, distillation separation sequences, azeotropic distillation, and heuristic optimization. In the spring, dynamic plantwide control covers dynamic computer simulation, pressure-driven plumbing, control structure development, and controller tuning.

Commercial flowsheet simulation software is now sufficiently user friendly that undergraduates can produce steady-state and dynamic simulations of fairly complex processes. Computer speed has increased to the point that dynamic simulations of fairly complex flowsheets can be run in reasonable times. Figure 1 presents an example of a flowsheet generated by a senior design group. Note that all the plumbing details are not given in the flowsheet, particularly the overhead piping, valves, reflux drum, and pump.

The organization of the Lehigh course has three-person groups, with each group working on a different design project. These projects are supplied by an industrial consultant who works with the group throughout the year. Active and retired engineers from industry graciously volunteer their time and years of practical experience to this effort. Engineers have participated from Air Products, DuPont, Exxon-Mobil, FMC, Praxair, Rohm&Haas, and Sun Oil.

As educational aids in the area of plantwide control and in the use of commercial dynamic simulators, two textbooks have been written.[4,5] Two basic types of errors are made by many students: inoperable plumbing arrangements and unworkable control structures. We consider these in the following sections.

COMMON PLUMBING ERRORS

The lack of physical understanding of practical fluid mechanics by many students is somewhat alarming. They have learned momentum balances, boundary-layer theory, the Navier-Stokes Equation, etc., in their fluid mechanics course. But when it comes to putting together a piping system to get material to flow around in a process, many students have great difficulty in coming up with a reasonable plumbing system.

The commercial process simulators have contributed to this weakness by permitting *flow-driven* dynamic simulations in which material "magically" flows from one unit to another despite the fact that the first unit is at a lower pressure than the second.

Fortunately *pressure-driven* dynamic simulations are also available. These are much closer representations of reality. Pumps, valves, and compressors must be inserted in the flowsheet in the required locations so that the principle "water flows downhill" is satisfied.

William L. Luyben *earned degrees in chemical engineering from Penn State (B.S., 1955) and Delaware (Ph.D., 1963). His industrial experience includes four years with Exxon, four years with DuPont, and three decades of consulting with chemical and petroleum companies. He has taught at Lehigh University since 1967 and has participated in the development of several innovative undergraduate courses.*

In my experience about 50% of the problems in designing and operating a real chemical plant involve hydraulics. Students need to have a solid understanding of practical fluid mechanics. Pressure-driven dynamic simulations provide a useful platform for developing this vital plumbing know-how.

The following is a brief compilation of some of the most common plumbing errors that students make in developing flowsheets. It might be useful to also state that I have seen many of these same errors made by presumably experienced engineers on real plants. So perhaps they are not quite as obvious as one might think.

No Valve Installed

Perhaps the most serious plumbing error, and one that is alarmingly common in student flowsheets, is to not have any valve in a line connecting process units that are operating at different pressures. This is illustrated in Figure 2 where a process stream flows from a vessel operating at a pressure of 10 bar into a vessel operating at 2 bar. There must be a valve in this line to take the pressure drop and regulate the flow. The valve can be set by an upstream controller (*e.g.*, level or pressure controllers), or it can be set by a downstream controller. But a valve is required.

Students often state that the pressure can be reduced by just cooling the stream. They confuse a "closed" system having a fixed amount of material with the "open" flow system encountered in a continuous-process flowsheet.

Stream Flowing "Uphill"

Equally distressing is to see a flowsheet in which a process stream is shown as flowing from a low-pressure location into a unit at higher pressure. Students often forget to put in the necessary pumps or compressors.

Two Valves in Liquid-Filled Line

This is probably the most frequently made error. Since a liquid is essentially incompressible, its flowrate is the same at any point in a liquid-filled line. Therefore the flowrate can be manipulated at only one location.

This means there should be only one valve in the line that is regulating the flowrate of liquid. It is physically possible to install two valves in series in a line, but these two valves cannot function independently.

Figure 3 shows several examples of this type of "forbidden" plumbing arrangement. When a stream is split into two streams at a tee in the line, the flow through each branch can be independently set by two valves. The same is true when two streams are combined.

Note that we are talking about *liquid*-filled lines. For gas systems, valves can be used in a line at several locations.

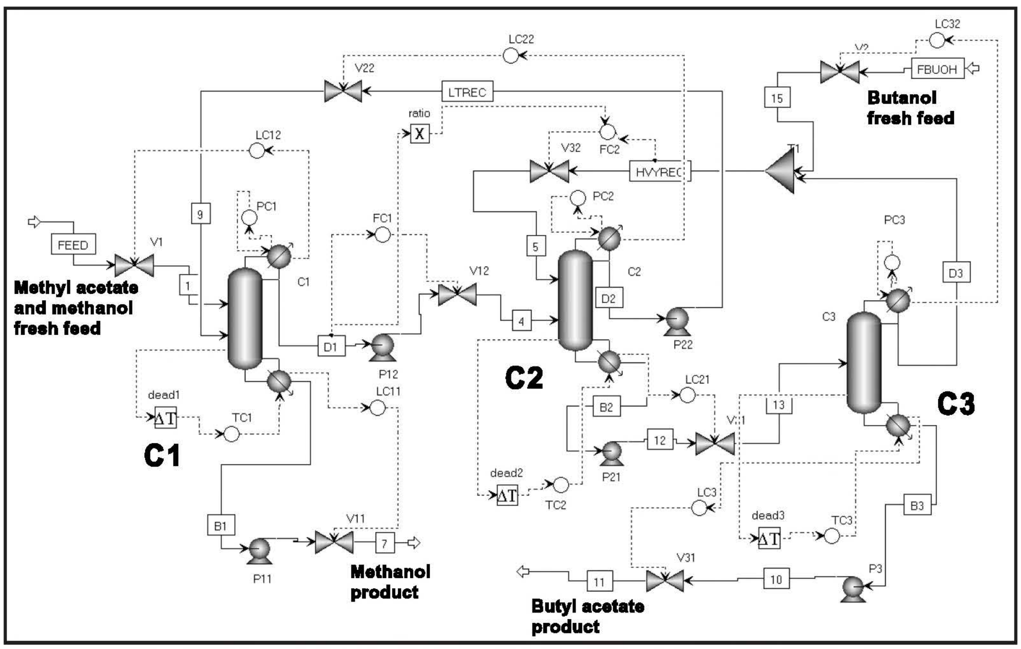

Figure 1. *Example of plantwide control structure.*

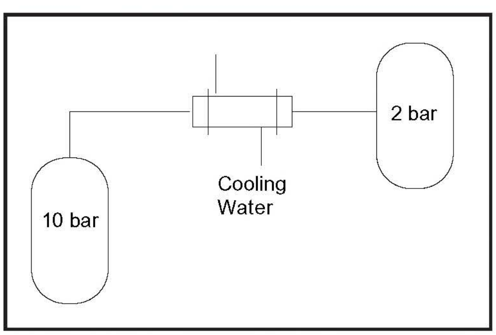

Figure 2. *Missing valve.*

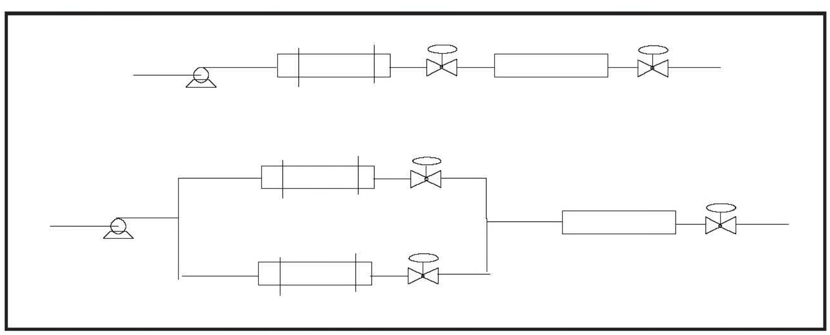

Figure 3. *Forbidden plumbing: two valves in liquid-filled line.*

Figure 4 illustrates this situation. The pressure in the first vessel is regulated by valve V1. The pressure in the second vessel is regulated by valve V2. This is workable because gas is compressible, so the instantaneous flowrates through the two valves do not have to be equal as is the case with liquids. The gas pressure in the process units can vary between the two valves.

Valve in Suction of Pump

Pumps are used to raise the pressure of a liquid stream. Compressors are used for the same purpose in gas systems. In this section we are considering liquid flows using *centrifugal* pumps.

Although students have learned about net positive suction head (NPSH) requirements for pumps, they frequently forget about this concept and install a control valve in the suction of a pump. Figure 5 illustrates this forbidden plumbing. Suppose the liquid is coming from the base of a distillation column. This liquid is at its bubblepoint under the conditions in the column. The base of the column must be located at an elevation high enough to provide adequate pressure at the pump suction to prevent the formation of vapor in the pump. This is the NPSH requirement.

If a control valve is installed between the column and the pump suction, the pressure drop over the valve will create a pump suction pressure that violates the NPSH requirements. So control valves in liquid systems should be located downstream of centrifugal pumps. The exact opposite is true for gas systems with compressors, as discussed in the next section.

It should also be remembered that no valves should be used for *positive displacement* pumps. The flowrate of the liquid can only be regulated by changing the stroke or speed of the pump or by bypassing liquid from the pump discharge back to some upstream location. The lower part of Figure 5 illustrates this forbidden plumbing with a positive displacement pump. Throttling a valve in the pump discharge will not change the flowrate of liquid through the pump. It will just increase the pump discharge pressure and raise the power requirement of the motor driving the pump.

Valve Downstream of Centrifugal Compressor

Centrifugal rotary compressors are positive displacement devices. At a fixed speed they compress a fixed volume of gas per time (ft^3/minute).

The mass flowrate of gas depends on the density of the gas at the compressor suction, so changing the suction pressure will change the mass flowrate.

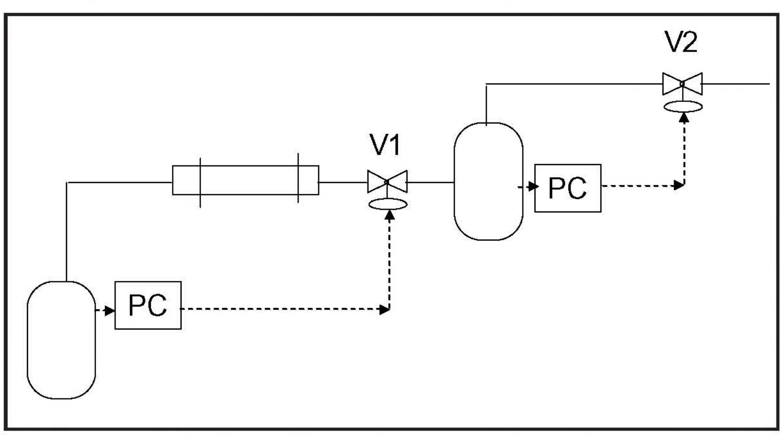

Figure 4. *Two valves in gas-filled line.*

Throttling a valve in the compressor *suction* changes the compressor suction pressure, so it can be used to control the gas flowrate. But throttling a valve in the compressor *discharge,* as shown in Figure 6, does not change the gas flowrate. It just increases the compressor discharge pressure and power requirements.

There are three viable ways to regulate the flowrate of gas in a compression system:

1. *Suction throttling*
2. *Bypass or spill-back from discharge to suction*
3. *Change compressor speed*

The last option is the most energy efficient but requires a variable-speed drive, which is typically a steam turbine if high-pressure steam is available in the plant. Variable-speed electric motors are also available. In compressor simulations this variable-speed option can be easily simulated by manipulating compressor work.

In the discussion above we have considered centrifugal compressors. Regulation of flow through a *reciprocating* compressor can be adjusted by throttling a valve in the suction, by changing

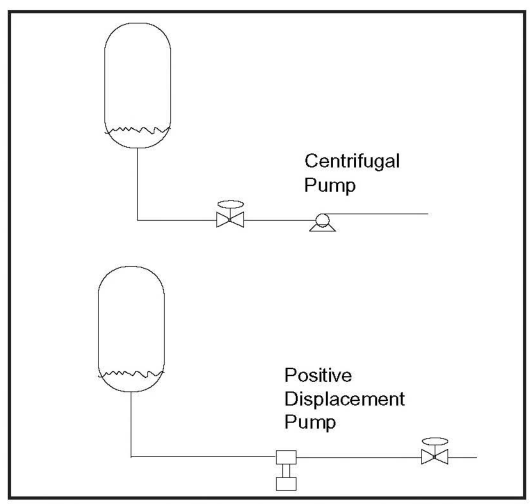

Figure 5. *Forbidden pump plumbing.*

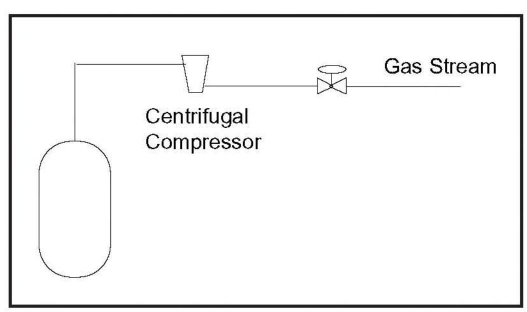

Figure 6. *Forbidden compressor plumbing.*

speed, or by changing the length of the stroke—but not by throttling a valve in the discharge.

Reciprocating gas compressors usually have clearance pockets that change the flowrate slightly, and therefore only provide minor adjustments in flow.

COMMON CONTROL STRUCTURE ERRORS

Most students in a senior design course have had a course in control fundamentals. They have been exposed to the mathematics and to the tuning of single-input, single-output feedback control loops with specified variables to be controlled and manipulated.

To develop a control scheme for a typical process, however, many control loops are required. Decisions must be made about what to control and what to manipulate. Students have had little exposure to this more complex *and* more realistic situation.

The most practical way to learn how to develop a plantwide control system is to examine several realistic examples and step through a logical plantwide design procedure.[5] At Lehigh, several lectures are given early in the second semester discussing reactor control, distillation control, and plantwide control. Then the design groups attempt to develop a control structure for their individual flowsheets. Despite these lectures and reading assignments in the textbook, the students' first efforts at developing a plantwide control system often contain many control-structure errors. Some of the more common are listed below.

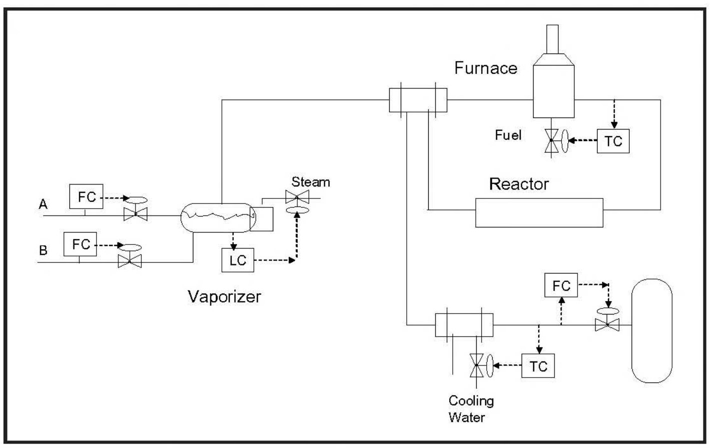

Figure 7. *Flows fixed in and out.*

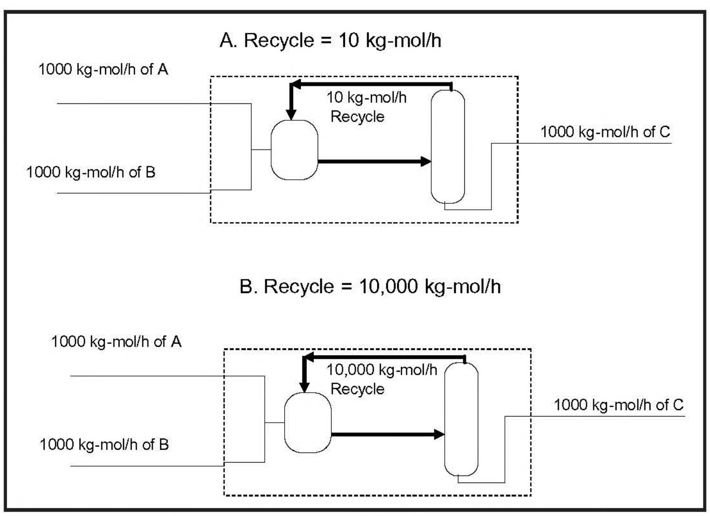

Figure 8. *Recycle independent of fresh feed.*

Fixing Flows Both In and Out

Figure 7 shows a process in which two liquid streams, containing reactants A and B, are fed into a vaporizer. Each stream is flow controlled.

The liquid feeds are vaporized and preheated before entering an adiabatic tubular reactor. Reactor effluent is cooled and fed into a downstream distillation column. The flowrate to the distillation column is flow controlled.

It is obvious that this structure is unworkable. But control schemes like this are proposed year after year by several groups of very capable students. They get wrapped up in the individual unit operations and neglect to look at the big picture.

Similar conceptual issues often occur in specifying recycle streams. Students often have trouble realizing that the flowrate of a recycle stream is completely independent of the flowrate of a fresh-feed stream. Fresh-feed flowrates are set by the production requirements. To produce 1000 kg-mol/h of a product C in a process with the reaction A + B $\rightarrow$ C, the fresh feed of each of the reactants must be 1000 kg-mol/h. Of course, if any reactants are lost as impurities in the streams leaving the unit, the fresh feeds must be appropriately larger. But inside the process we could have a recycle stream of reactant A, for example. As illustrated in Figure 8, the flowrate of this recycle can be anything we want it to be: 10 kg-mol/h or 100,000 kg-mol/h.

Recycle flowrate is completely independent of fresh-feed flowrate.

Liquid Levels and Gas Pressures Not Controlled

Students frequently submit flowsheets in which there is no control of liquid levels in vessels or no control of pressure in gas-filled systems. All liquid levels must be controlled in some way. They can be controlled by manipulating a downstream valve or by manipulating an upstream valve. Of course, the level control schemes for the individual units must be consistent with the plantwide inventory control scheme that connects all the units.

There are very few exceptions to this requirement for controlling all levels. The most common exception is when a solvent is circulating around inside a process and there are no losses of this solvent. In this case there will be a liquid level somewhere in the process that "floats" up and down as the solvent circulation-rate changes. This level is not controlled.

The pressure in a gas-filled system must also be controlled. Gas pressure can be controlled by regulating the flow of gas into or out of the system. It can also be controlled by regulating the rate of generation of gas (*e.g.*, in a vaporizer, in a distillation column reboiler, or in a boiling exothermic reactor). Pressure can also be controlled by regulating the rate of condensation of gas (*e.g.*, in the condenser of a distillation column).

The system can consist of several gas-filled vessels with vapor flowing in series through the vessels. Figure 9 illustrates some of these ideas. In this flowsheet the pressure in the gas loop is controlled by the rate of addition of a gas fresh-feed stream. The pressures in all of the vessels float up and down together, but differ slightly due to pressure drops (which are typically kept quite small to reduce compression costs). The flowrate of the gas recycle stream is flow controlled, using a cascade system: Flow controller output adjusts the setpoint of the turbine speed controller, whose output manipulates high-pressure steam to the turbine.

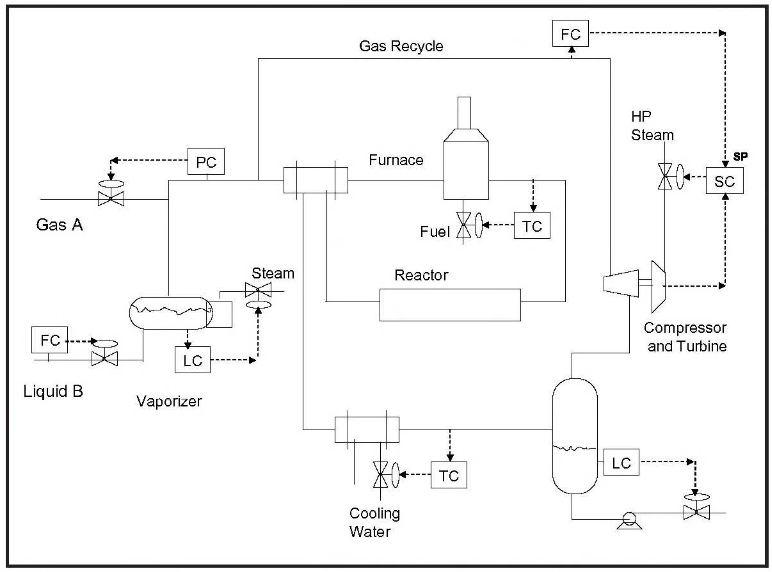

Figure 9. *Pressure in gas loop.*

There are rare occasions when pressure is allowed to float. These occur when it is desirable to keep pressure as low as possible for some process optimization reason (*e.g.*, in some distillation columns where relative volatilities increase with decreasing pressure). In these systems heat removal is maximized to keep pressure as low as possible.

Distillation Columns with a Fixed Product Flowrate

The first law of distillation control says that you cannot fix the distillate-to-feed ratio in a distillation column and also control any composition (or temperature) in the column. This law is a result of the very strong impact of the overall component balance on compositions and the relatively smaller effect of fractionation (reflux ratio, steam-to-feed ratio, etc.) on compositions.

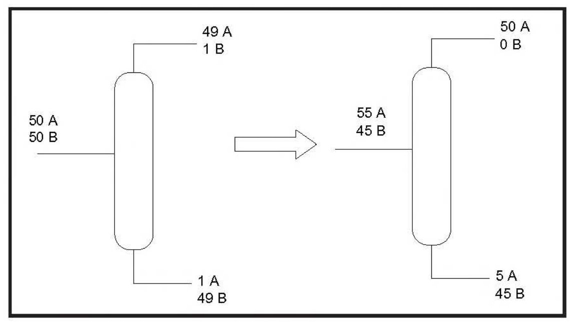

Figure 10. *Fixing product stream in distillation column.*

Figure 10 illustrates the effect of fixing the distillate and bottoms flowrates when changes in feed composition occur. Initially the feed contains 50 mol/h of A and 50 mol/h of B. The distillate contains 49 mol/h of A and 1 mol/h of B, and the bottoms contains 1 mol/h of A and 49 mol/h of B. So product purities are 98 mol%. Then the feed composition is changed so there are 55 mol/h of A and 45 mol/h of B. The distillate and bottoms flowrates are kept constant at 50 mol/hr. Now the distillate will be essentially 50 mol/h of A, and the bottoms will be 5 mol/h of A and 45 mol/h of B. Thus the bottoms purity will drop from 98 mol% B to 90 mol% B. No matter what reflux ratio or reboiler heat input is used, this purity cannot be changed. Controlling a composition or a temperature in the column is not possible.

There are columns in which a product stream is fixed. These are called "purge columns" because the purpose is to remove a small amount of some component in the feed. In these columns, temperature or composition is not controlled. The flowrate of the purge stream is simply ratioed to the feed flowrate.

A somewhat more complex situation occurs when the purging is done in a sidestream column that has three product streams. Consider the sidestream columns shown in Figure 11. The feed stream is a ternary mixture. Two cases are shown. In the column on the left the feed contains a small amount of the lightest component, and it is purged in the distillate stream. The intermediate component is removed in the liquid sidestream.

The distillate is flow controlled, and reflux-drum level is controlled by manipulating reflux flowrate. The issue here is how to manipulate the sidestream flowrate. It cannot be fixed but must change in response to feed composition and flowrate disturbances. The scheme shown in the left of Figure 11 achieves this by ratioing the sidestream flowrate to the reflux flowrate. Temperature or composition can be controlled in this column because the separation between the intermediate and heavy components can be adjusted.

In the column on the right in Figure 11, the feed contains a small amount of the heaviest component, and it is purged in the bottoms stream.

The intermediate component is removed in the vapor sidestream. The bottoms stream is flow controlled, and base level is controlled by manipulating

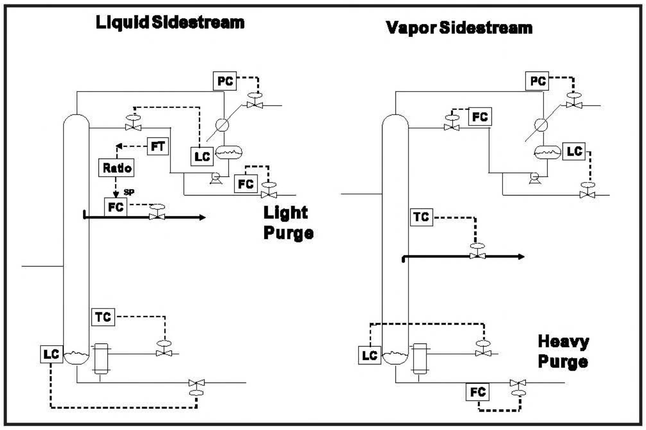

Figure 11. *Purge column with sidestream.*

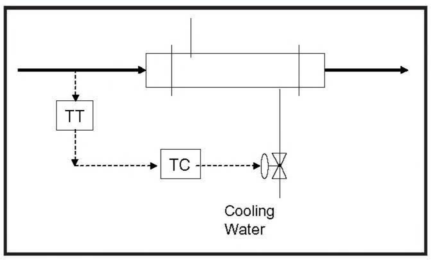

Figure 12. *Herron Heresy.*

reboiler heat input. The vapor sidestream flowrate, which cannot be fixed, is manipulated to control a temperature in the column. Note that when a small amount of light impurity is present in the ternary feed, a liquid sidestream of the intermediate component is used with its drawoff location above the feed location. This configuration is used because the liquid at the sidestream tray has a lower concentration of the lightest component than the vapor. When a small amount of heavy impurity is present in the ternary feed, a *vapor* sidestream of the intermediate component is used with its drawoff location *below* the feed location because the vapor at the sidestream tray has a lower concentration of the heaviest component than the liquid.

Incorrect Sensor Location and Valves Without Input Signals

Figure 12 shows what we call at Lehigh the "Herron Heresy" (after a senior student in the design course who made the same mistake twice). The diagram shows that the temperature upstream of the cooler is controlled by the flowrate of cooling water to the heat exchanger. This, of course, is impossible and should be obvious. Yet this type of error crops up on several flowsheets every year.

Sometimes students correctly insert a valve in a line to satisfy plumbing requirements, but fail to connect it to a controller. All valves must be positioned by some controller.

Ratioing Reactant Feeds

One of the most important aspects of plantwide control is the manipulation of the fresh-feed streams. A common error is to simply ratio the flowrates of the reactants so as to satisfy the reaction stoichiometry. Although this will work in a simulation study, it will not work in reality.

Flowrates cannot be measured accurately enough to guarantee an absolute matching of the number of molecules of the various reactants. The separation section typically prevents the loss of any of the reactants. Therefore simply ratioing reactants inevitably results in a gradual buildup inside the process of the reactant that is in slight excess.

Some indication of the inventory of the reactants inside the system must be found so that the flowrates of the fresh-feed streams can be appropriately adjusted. Ultimately these flows must satisfy the reaction stoichiometry down to the last molecule. But this much accuracy is way beyond our ability to measure flowrates.

The plantwide control structure in Figure 1 illustrates this principle. The chemistry in this example is the reaction of methyl acetate and butanol to produce butyl acetate and methanol. The reaction occurs in a reactive distillation column (C2). There are two recycle streams. The "LTREC"—the distillate D2 from the reactive column—is an azeotropic mixture of methyl acetate and methanol. The "HVYREC" is the distillate D3 from the third column, which is mostly recycled butanol.

The fresh butanol is added to this recycle stream to control the reflux-drum level in the third column (level controller LC32). This level gives an accurate measurement of the amount of butanol in the system. If more butanol is reacting than is being fed, this level will decrease. On the methyl acetate side, the level in the reflux drum of the first column is controlled by manipulating the fresh-feed stream, which contains methyl acetate and methanol (level controller LC12). This level provides a measurement of the methyl acetate in the system.

Note that the production rate in this plant is set by the flow controller FC1, which controls the feed flowrate D1 to the second column. If more production is desired, the operator increases the setpoint of this flow controller. The increase in D1 also results in an increase in the flowrate of the heavy recycle because of the ratio.

CONCLUSION

Common plumbing and control concept errors have been discussed and illustrated. It is hoped that this paper will help students and engineers avoid these problems in their design projects, and more importantly, in real life. Most of these errors are obvious and can be avoided by using some common sense and not getting all wrapped up in the computer simulation aspects of the problem.

REFERENCES

1. Seider, W.D., J.D. Seader, and D.R. Lewin, *Product and Process Design Principles*, Wiley (2004)
2. Dimian, A.C., *Integrated Design and Simulation of Chemical Processes*, Elsivier (2003)
3. Luyben, W.L., Chapter A1, "The need for simultaneous design education," in *The Integration of Process Design and Control*, P. Seferlis and M.C. Georgiadis, editors, Wiley (2004)
4. Luyben, W.L., *Plantwide Dynamic Simulators for Chemical Processing and Control*, Marcel Dekker (2002)
5. Luyben, W.L., B.D. Tyreus, and M.L. Luyben, *Plantwide Process Control*, McGraw-Hill (1999) ❒