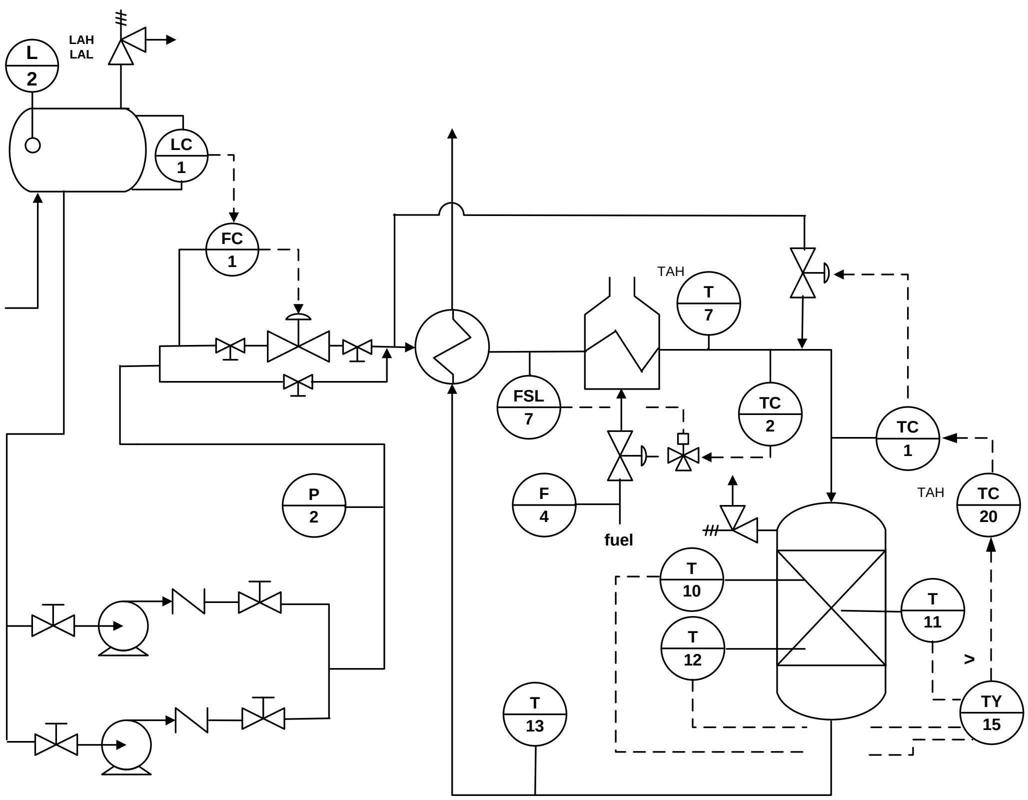
Operability in Process Design: Achieving Safe, Profitable, and Robust Process Operations

Chapter 6 Process Control



Thomas Marlin

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- Tariq Samad and Russ Rhinehart for assistance with permissions from the American Control Conference
- Everyone who posts materials with the Creative Commons license

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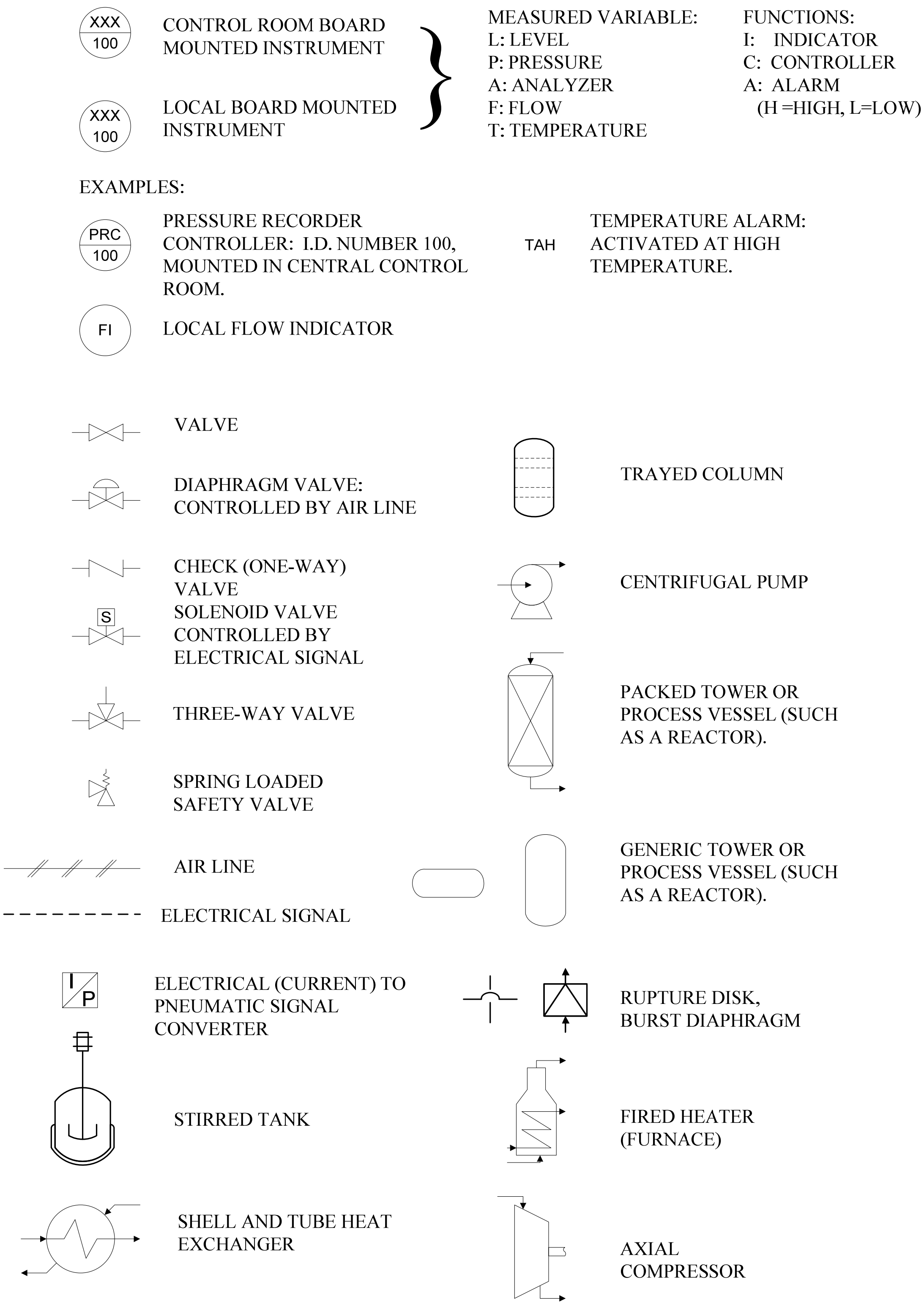
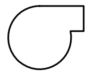
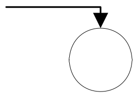
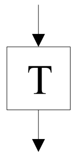
Symbols


XXX 100
CONTROL ROOM BOARD MOUNTED INSTRUMENT
XXX 100
LOCAL BOARD MOUNTED INSTRUMENT
MEASURED VARIABLE:
L: LEVEL
P: PRESSURE
A: ANALYZER
F: FLOW
T: TEMPERATURE
FUNCTIONS:
I: INDICATOR
C: CONTROLLER
A: ALARM
(H =HIGH, L=LOW)
EXAMPLES:
PRC 100
PRESSURE RECORDER CONTROLLER: I.D. NUMBER 100, MOUNTED IN CENTRAL CONTROL ROOM.
TAH
TEMPERATURE ALARM: ACTIVATED AT HIGH TEMPERATURE.
FI
LOCAL FLOW INDICATOR
VALVE
DIAPHRAGM VALVE: CONTROLLED BY AIR LINE
CHECK (ONE-WAY) VALVE
S
SOLENOID VALVE CONTROLLED BY ELECTRICAL SIGNAL
THREE-WAY VALVE
SPRING LOADED SAFETY VALVE
AIR LINE
ELECTRICAL SIGNAL
I
P
ELECTRICAL (CURRENT) TO PNEUMATIC SIGNAL CONVERTER
STIRRED TANK
SHELL AND TUBE HEAT EXCHANGER
TRAYED COLUMN
CENTRIFUGAL PUMP
PACKED TOWER OR PROCESS VESSEL (SUCH AS A REACTOR).
GENERIC TOWER OR PROCESS VESSEL (SUCH AS A REACTOR).
RUPTURE DISK, BURST DIAPHRAGM
FIRED HEATER (FURNACE)
AXIAL COMPRESSOR


CENTRIFUGAL FAN


HEADER (STEAM OR CONDENSATE)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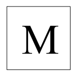
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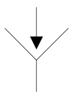
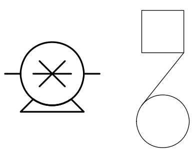
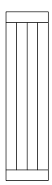

DRAIN TO SEWER


POSITIVE DISPLACEMENT PUMP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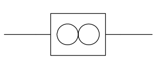
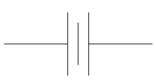

TURBINE FLOW METER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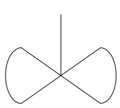
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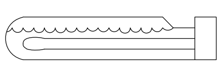
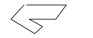
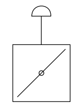
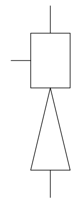

VAPORIZER OR REBOILER


FILTER AND DRAIN


CYCLONE


DAMPER OR BUTTERFLY VALVE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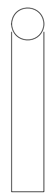
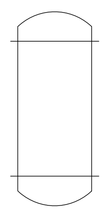
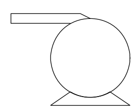
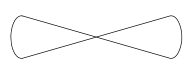
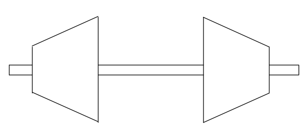

FLARE STACK


SHELL AND TUBE HEAT EXCHANGER


FAN OR BLOWER


FAN


TURBINE USED TO PROVIDE WORK TO DRIVE COMPRESSOR

Nomenclature

A	Area for heat transfer
B	Benefits for the change in dynamic performance ($/y)
BPCS	Basic process control system
C_P	Heat capacity
CSTR	Continuous (flow) stirred tank reactor
CV	Controlled variable
D	Disturbance
D_1	Distribution of the controlled variable under new control strategy (histogram)
D_0	original (base case) distribution in controlled variable (histogram)
E	Error = SP - CV
F_j	Fraction of data in histogram in range j
h	Film heat transfer coefficient
HAZOP	Hazard and Operability Study
I	Initialization constant in PID control algorithm (also termed “bias”)
I_v	Incremental value of change in process performance
IAE	Integral of absolute value of error $IAE = \int_0^\infty \lvert SP(t) - CV(t)\rvert \, dt$
ISE	Integral of the squared error $ISE = \int_0^\infty (SP(t) - CV(t))^2 \, dt$
K	Process gain
K_C	Controller gain
K_d	Disturbance gain
LOPA	Layer of Protective Analysis
M	Correction factor in benefits calculation for other operating conditions, e.g., production rate, (dimensionless)
MV	Manipulated variable
P&ID	Piping and instrumentation drawing
PID	Proportional-integral-derivative controller
PR	Process performance appearing in benefit calculation
Q	Heat transferred
SF	Service factor, the fraction of time that the control strategy is improving the process performance (dimensionless)
SIS	Safety instrumented system (sometimes referred to as “safety interlock system”)
SP	Set point

T	Time
T	Time when the control strategy should be in service (h/y)
T_d	Controller derivative time
T_I	Controller integral time
U	Overall heat transfer coefficient
$\Delta V(*,*)$	Improved economic process performance at the base case operating conditions ($/h), based on two arguments (D_0, D)
Greek symbols	
λ	Relative gain array element
θ	Dead time
ρ	Density
τ	Time constant
ω	frequency

Chapter 6. Process Control

6.0 To the Student

Imagine that you are the operator of a complex chemical process without process control. You introduce a change to the raw material feed rate to the process. In response, you would have to implement changes to 10's to 100's of additional valve openings, affecting condenser cooling and reboiler heating, solvent flows, boiler fuel and airflows, recycle flows, and many more. You would not only have to introduce these changes in the correct magnitude and direction but also at the correct times as the production rate change coursed through the process. This demanding task would consume all of your time, would lead to periodic human error, would lead to frequent large deviations in product quality, and could occasionally result in unsafe operations.

Fortunately, there is an alternative provided by process control! Process control is essential for achieving safely, reliably, and desired process conditions as disturbances occur in the plant. While engineers and operators understand the needed actions, only precise automation through process control can implement actions as rapidly and reliably as required in demanding chemical processes. In fact, the entire development of the technology of automatic control has been in response to demands for controlling steam engines, airplanes, electronic circuits, chemical plants, and many other complex systems.

Since you have already completed a course covering process control, this chapter presents complementary materials that address some key design concepts. You may be surprised (and pleased) to find that this coverage has little mathematical sophistication. This decision is not made because dynamic modeling, simulation and controller calculations are unimportant but because you have already mastered these topics in your process control course.

This is process control without the mathematics, but it requires deep understanding and careful thought to successfully apply the topics to control design.

Little math ≠ Easy to do without preparation, skills, and knowledge!

This material emphasizes designing process and control structures. It explains process characteristics that make process control challenging (or easy), and it presents control structures to achieve the best dynamic performance possible given the process design and disturbances occurring. Naturally, process control is realized with physical equipment for sensing, final elements, signal transmission, human interfaces, and computation. Some of the key aspects of control equipment are addressed in Appendix A.

This material supports all topics in operability, including product quality, safety, reliability, and troubleshooting. So, let's learn some more process control!

6.1 Basics of Process Control

Why is process control necessary in a chemical process? Some reasons are given in the following; can you think of others?

- Plants are physically large, so that adjustments and data collection must be managed from centralized locations
- Materials can be hazardous and are often maintained at extreme conditions, e.g., high pressures and temperatures
- Equipment functions successfully without damage over only a limited range of conditions, so that excursions outside of acceptable ranges must be avoided
- Demands for product quality and safety require rapid and precise process adjustments that are often beyond the capability of plant personnel
- People must be relieved of high frequency decision making, so that they can perform more complex analyses that are better performed by people

Process control involves a large and continuously expanding array of technology. Here, we will address the technology that is implemented in a typical process control design. This technology is based on one basic principle, feedback.

Feedback uses information from system outputs for deciding adjustments to system inputs.

The use of system outputs requires measurements of process variables that are influenced or caused by adjustable variables. The selection of output variables for measurement is critical to success and will be addressed throughout the chapter. Manipulated input variables can be adjusted by a person or computer; for example, a valve opening is an acceptable input variable. In contrast, disturbance input variables cannot be adjusted; an example would be raw material composition.

The schematic in Figure 6.1 shows a feedback control loop with limited detail, containing the essential three elements of sensor, control calculation and final element. The loop requires inputs from plant personnel in the form of controller tuning constants and the set point that defines the desired value for the variable. Then, the controller functions essentially continuously by adjusting the valve to bring the controlled variable to its set point. However, process control does not result in a plant running on "automatic pilot". Since control systems involve complex equipment that can fail to operate properly, plant personnel monitor the performance of process and control equipment and intervene when they diagnose a fault.

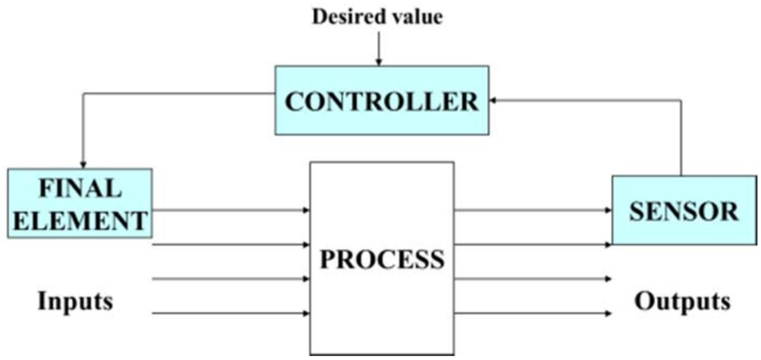


Figure 6.1 Schematic of Feedback control loop

Figure 6.1 pictures the concepts of control, but engineers need to design equipment to realize these concepts. Let's begin with a single control loop shown in Figure 6.2 to determine the key equipment common to all control systems. The sensor generates a signal that is proportional to the measured variable; in the example, a thermocouple produces a millivolt signal that is (approximately) proportional to the temperature. The sensor signal is converted in a transmitter to an alternative signal that can be accurately transmitted longer distances and is compatible with other equipment in the loop; the transmitted signal would typically be an analog current (4-20 mA) or a digital signal. The transmitted signal is converted for use in the calculation equipment; this could be voltage for analog equipment or a decimal number for digital control. In addition to the controller calculation, the measurement signal is used for display and historical storage

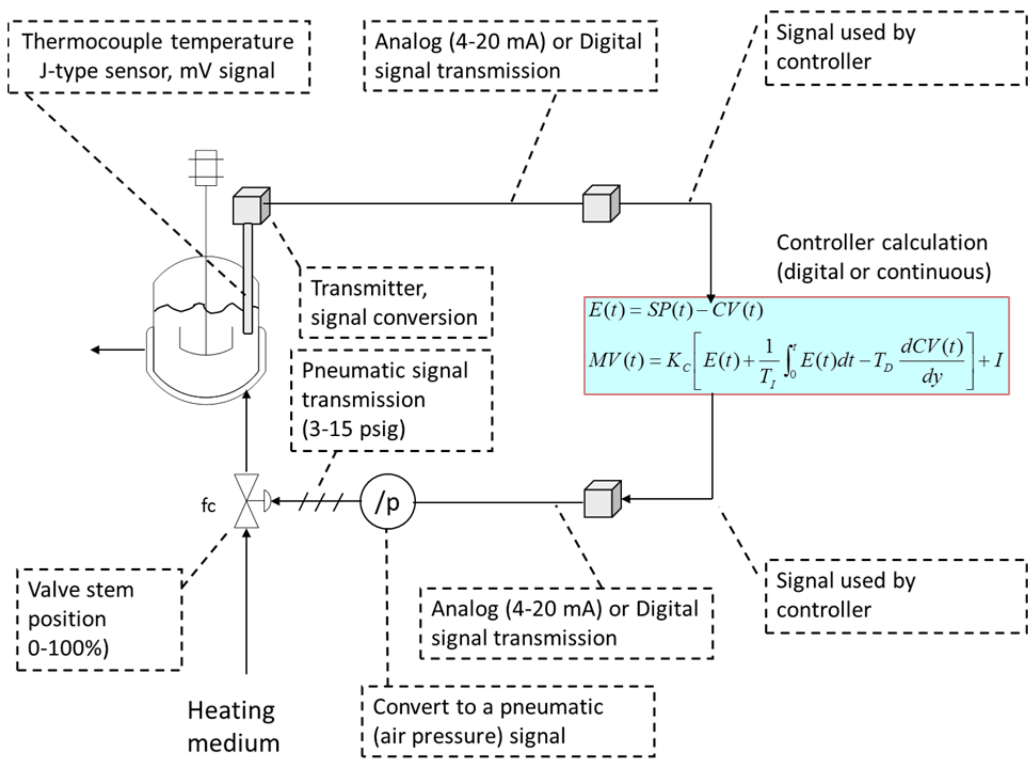


Figure 6.2. Typical equipment for a single-loop controller, with human-machine interface is not shown. Note that signal transmission and control calculation control equipment can be based on analog (shown here) or digital technology.

and can be used for calculations. The controller output is converted for transmission to the final element. Near the final element, the transmitted signal is converted to affect the final element. In many process applications, the final element is a control valve, and the signal must determine the force provided by compressed air. Thus, the transmitted signal is converted to an air pressure. The pneumatic signal is applied to the final element that changes its stem position and opening for flow. Every control loop in a plant has its own individual local equipment, sensor, pneumatic converter, and final element. The control calculation is performed digitally in equipment manufacturer in the last thirty years. (Control equipment has a long life, so you might encounter analog control equipment that performs the PID calculations via an electronic circuit.) The transmission until recently has been achieved using an individual wire carrying an amperage for each signal. More recently, digital transmission has been introduced for transmission.

A large segment (or the entirety) of a process plant is controlled in a centralized location, where a few people can observe all measurements and make adjustments throughout the plant. The centralized control house enables coordinated actions, but it requires long transmission. Fortunately, electronic and digital signals can be transmitted with essentially no delay. A picture of a typical centralized control room is shown in Figure 6.3.

All sensors are located at the process equipment, while displays of the measurements can be located either at the equipment or centrally, or both. Local displays of measurements are essential for plant personnel who are performing maintenance and are monitoring the equipment. For example, when an operator starts a pump, s/he wants to observe the outlet pressure and perhaps, the flow as well, to ensure that the equipment is working properly. However, coordinated analysis and control of the entire plant requires that most measurements be transmitted and displayed in the centralized control room. Most of these will be recorded on trend plots to provide a display of the recent dynamic behavior. When appropriate, a measurement can be displayed both locally and remotely.

Figure 6.3 Picture of a typical centralized control room. (Photo courtesy of Worsley Alumina)

Modern control equipment in the centralized facility includes a network of computers. The network has the following advantages (over a single, centralized computer).

- Parallel computation ensures minimal computing delay in control loops.
- The behavior for a failure is superior. Even though the probability of failures is higher, because of the increased number of computers, the impact of a failure is much lower since only a small section of the plant would be affected. To further improve reliability, most digital equipment is redundant with automatic switching upon failure detection.
- Computer software and hardware can be tailored in each module for specific functions, like process control, complex and flexible computations, history storage and display, safety functions, and so forth.
- The computing system can be designed to match a plant, without excess capacity, while allowing subsequent modular expansion.

It is important to recognize that control systems have many preprogrammed functions, so that plant engineers do not program PID control algorithms, details of graphical displays, and so forth. Most control systems require "configuring" calculations, displays and history storage using existing functions.

We conclude this section with a brief discussion of drawings that are used to document designs. We have to recognize that complex designs could not be documented using written descriptions alone. Drawings are widely used as a basis for construction, and there are many forms of drawings, including Block Flow, Mechanical Detail, Piping and Instrumentation Drawing (P&ID), and Isometric (3-D) Layout. A clear explanation of process drawings is provided with examples by Turton et. al. (2012). Here, we will concentrate on the P&ID, whose major characteristics are given in Table 6.1. The P&IDs are used during day-to-day operation and for safety studies; therefore, the P&ID must be maintained up-to-date as changes are made to the original design and construction.

Various levels of detail are presented in a P&ID. Limited information is available during the preliminary design of the process, so that the P&ID does not include as much detail concerning the sensors and control implementation. The drawings in this educational material will tend to follow the preliminary P&ID level of detail. Preparing P&IDs is facilitated by special-purpose software that includes a library of process-related symbols. Perhaps the best, low-cost software for use by university students is MS Visio™.

Prior to addressing control technology, we need to refresh our understanding of two aspects of basic process control. The first aspect involves classical control methods, which are used widely in chemical processes; this aspect is addressed in the next section. The second aspect is process control equipment, i.e., instrumentation; the reader is referred to Appendix A for a review of instrumentation. This information provides the platform for designing process and control structures.

Table 6.1. Typical features of Piping and Instrumentation Drawings (P&ID)*

P&ID contain the following	P&ID does not contain the following
• All piping and equipment connections	• The distance between objects. The drawing is not to scale.
• An approximate location for connections (e.g., top or bottom of tank, tray location, etc.)	• The vertical or horizontal (or 3D) position of objects
• Equipment identification (numbers)	• The sizes of objects(e.g., vessel), not even the relative size
• The size of piping	• The exact design for piping connections, including those to vessels
• All sensors (whether locally or remotely displayed and recorded) and whether used for an alarm, with priority	• Sensor details such as physical principle (e.g., thermocouple) and measurement range
• All valves (whether automated or manual), including failure position if remotely operated valves	• Details of the control calculations when involving complex logic and/or calculations
• Control strategies, as much detail as possible graphically. These can be regulatory and safety-related	• Any detail about the human interface display or the type of historical data
• Whether signals and control calculations are implemented using analog or digital equipment	• Operating policy (which appears in a separate operations manual)

* Various levels of detail are presented in a P&ID, depending on the status of the design (preliminary to definitive) and company practices.

6.2 Classical Control Methods

Classical control refers to a collection of technology that was developed over many decades and applied successfully in the process industries. Because of limited computing during the early decades of the twentieth century, classical control is founded on the PID feedback algorithm. While powerful and flexible, simple single-loop feedback is not alone adequate to achieved required dynamic performance. Therefore, a number of enhancements were developed that complemented single-loop PID; the enhancements addressed in this section are cascade control, feedforward control, non-square system control with PIDs, and inferential control. The methods in this section are required to design “industrial strength” classical control. The classical control methods are introduced briefly in this section, since most of the material refreshes a typical process control course. (Even though it was a great course, we do not want to review when we can learn new topics!) The topics on non-square systems and inferential control are presented in more detail, because they are not usually included in the undergraduate course and are essential for industrial control designs.

6.2.1 Single-loop Control

Single-loop control involves controlling one measured process variable using a controller that adjusts one final control element. Process control can be achieved through the application of many single-loop controllers, one for each important controlled variable. Generally, each control calculation shares no information with any other control calculation; it has "blinders on". Single-loop control has been the paradigm for process control for over two hundred years, because it provided adequate dynamic performance with simple real-time calculations, and it continues to be the dominant approach to industrial control. Later in the chapter, enhancements to the single-loop paradigm will be introduced. These enhancements are essential for the challenges posed by complex processes.

Nearly all single-loop feedback control is implemented with the Proportional-Integral – Derivative (PID) algorithm. This algorithm was developed originally to give good performance while requiring only simple computation. Remember that process control was needed and was applied long before digital computation became practical. Perhaps surprisingly, the algorithm has proved to be very good for most single-loop applications and remains the most widely employed in spite of today's powerful digital control systems.

The continuous PID algorithm is implemented in pneumatic and electronic calculation systems, and the discrete PID algorithm, which approximates the continuous using standard numerical methods, is implemented in digital control systems. Both of the algorithms are given in the following.

$$E(t) = SP(t) - CV(t) \tag{6.1}$$

Continuous:

$$MV(t) = K_c\left[E(t) + \frac{1}{T_I}\int_0^t E(t')dt' - T_d\frac{d\,CV}{dt}\right] + I \tag{6.2}$$

$$E_n = SP_n - CV_n \tag{6.3}$$

Discrete:

$$MV_n = K_C\left[E_n + \frac{1}{T_I}\sum_{i=0}^{n}(E_i * \Delta t) - T_d\left(\frac{CV_n - CV_{n-1}}{\Delta t}\right)\right] + I \tag{6.4}$$

where

CV = controlled variable
MV = manipulated variable
SP = set point
E = error
I = constant of initialization (bias)

t = time
Δt = execution period
n = current controller execution counter (integer number)
K_c = controller gain
T_I = controller integral time
T_d = controller derivative time

Many variations to these algorithms are used in practice. Some of the variations have (a) proportional on measurement, CV, (not error), (b) filtered measurement for the derivative calculation, and (c) interaction between the proportional and integral modes (Witt and Waggoner, 1990).

The "tuning constants" in the PID algorithm (K_C, T_I, and T_d) are adjusted to achieve good performance for each control loop. Many correlations are available to determine the constants based on the dynamic model for the loop, i.e., the response of the CV to a change in the MV. When selecting tuning methods, the engineer should be aware of the following.

- There is no one method that will give good performance for all loops because each loop has unique performance goals, measurement noise and model uncertainty.
- All tuning correlations should be considered an "opening gambit" that helps the engineer in an iterative procedure.
- Proper tuning depends on the form of the PID algorithm, and as previously mentioned, many forms are in use industrially. Therefore, the engineer must match the appropriate tuning correlation with the applicable version of the PID algorithm.

Since control performance is the ultimate goal, let's discuss loop performance for a moment. Some of the key performance factors are discussed in the following.

- **Controlled variable behavior** – We want the controlled variable to remain close to (ideally, exactly equal to) the set point. Performance is often measured by the IAE (integral of absolute value of error) for single step disturbances or variance for a long sample of data. Other issues can also be important; for example, large deviations (single overshoot or oscillations) can be especially deleterious to performance, so that overshooting a set point by a large amount during a disturbance should be avoided.
- **Manipulated variable behavior** – The controller must adjust the manipulated variable to compensate for disturbances, but very aggressive adjustments are undesirable in some processes. For example, boiler pressure is typically controlled by adjusting the fuel flow rate to the combustion flame. If the fuel is adjusted rapidly and frequently, the boiler is subjected to thermal stresses through rapid expansions and contractions that over time that can damage the equipment and require expensive shutdown and repairs. In addition, measurement noise is propagated by the controller to the manipulated variable, so that tuning constant values need to be selected to moderate the propagation of noise.
- **Robustness** – The process dynamics are not known exactly, and the dynamics change, sometimes substantially, because of changes in the process operating conditions. For example, the changes in production rate affect the dynamics on all control loops in the plant. When the controller tuning is constant but the loop dynamics changes, the control loop performance can degrade and even become unstable. Therefore, the controller tuning must be robust to process dynamics changes. Tuning with good robustness performs well (or acceptably) over a range of loop dynamics, while poor robustness yields rapid degradation as the dynamics change from the base case.

We see the key conflict in feedback control. Keeping the controlled variable near its set point requires aggressive adjustment of the manipulated variable, while robustness and limitations to MV variability favor more moderate adjustment of the manipulated variable.

Achieving good control performance involves a tradeoff among competing performance objectives. The importance of each goal and the circumstances in which it is achieved (e.g., amount of measurement noise and model mismatch) differs. Therefore, the engineer must evaluate each loop to determine the proper process and control design, control algorithm, and controller tuning.

Example 6.1 Good Loop Performance - The process considered in this example involves mixing two streams and has a pure dead time due to transportation delay and a first-order mixing tank. It is shown in Figure 6.4. A typical dynamic response plot for this single-loop PI control is shown in Figure 6.5.

The dynamic response is given in Figure 6.5 for a step change in the controller set point at 2.5 minutes. We note the following.

- ***The immediate response is a step in the manipulated variable due to the proportional mode.***
- ***After the set point change, the controlled variable does not respond for the duration of the dead time. During this time, the manipulated variable changes due to the integral mode***
- ***For typical loop tuning, the manipulated variable will slightly overshoot its final value and the controlled variable will slightly overshoot the set point.***
- ***Finally, the controlled and manipulated variables achieve steady state. The controlled variable is equal to the set point because of the integral mode.***

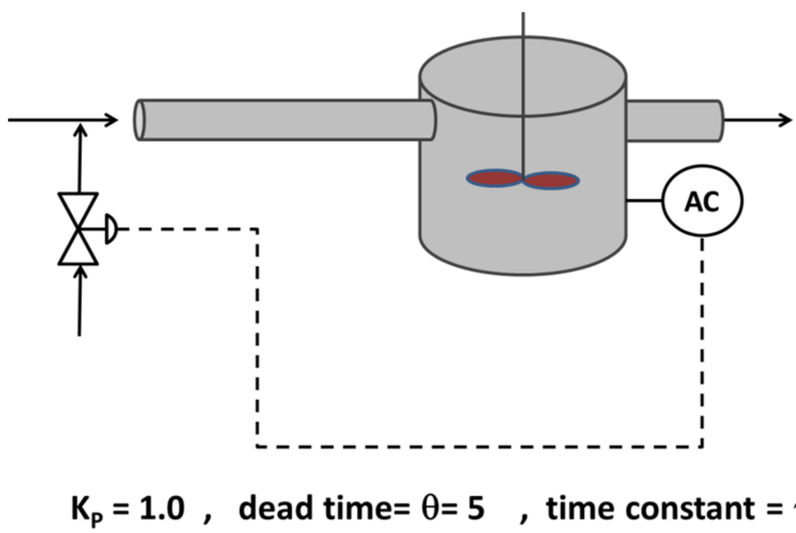


Figure 6.4 Single loop PI control for Example 6.1.

(This is not a good process design; we should minimize dead time and time constants in the feedback process, but it is used here for demonstration purposes because of the clear relationship between the pipe length and tank volume and the feedback dynamics.)

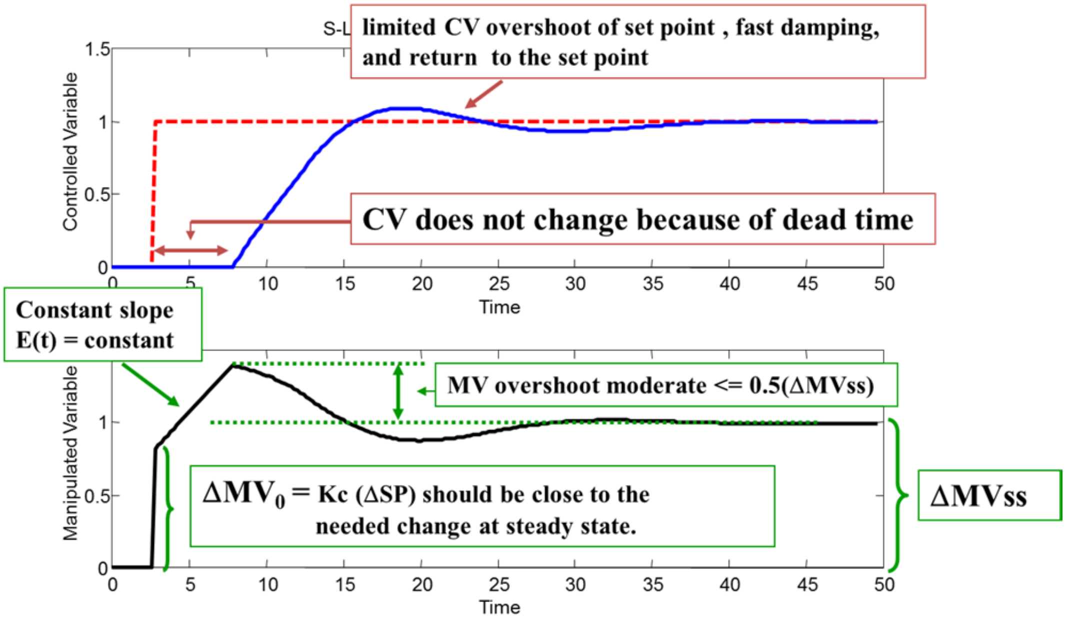


Figure 6.5 Typical good feedback response to a step set point change. (Variables plotted as deviation from initial steady state.)

Example 6.2. Control performance diagnostics - Let's apply the insights gained in the previous example to diagnose control loop behavior. The situation involves a loop that has typical performance goals, so that a pattern similar to that in Figure 6.5 is desired. The performance with current tuning is given in Figure 6.6. Decide (a) whether the performance is acceptable or not and (b) if not acceptable, determine as much possible, the cause of the unacceptable performance. Can you suggest a corrective modification?

In diagnosing a control loop, we first evaluate the performance of the instrumentation. For example, valve stiction and hysteresis could seriously degrade control performance. We will assume that this has been done and that that the instrumentation is functioning well. Our diagnose proceeds as follows.

- ***The control loop appears to be stable, which is essential.***
- ***The controlled variables behavior is too oscillatory, as is the manipulated variables behavior. This suggests that the controller is too aggressive for the process dynamics. "Too aggressive" would be caused by the controlled gain that is too large, the integral time that is too small, or both. (Remember that the integral time is in the denominator.)***
- ***We would like to evaluate the tuning.***
- ***We see that the "proportional kick" when the set point is initially changed is not too large; the manipulated variable changes near to but less than its final steady-state value. Therefore, we conclude that the proportional gain (K_C) has a reasonable value.***
- ***The manipulated variable has excessive overshoot. Since the controller gain is OK, we conclude that the integral time is too small. (Remember that the integral time is in the denominator.***
- ***We would increase the integral time (by about 50%) and perform another set point step test.***
- ***Continue until good performance is achieved.***

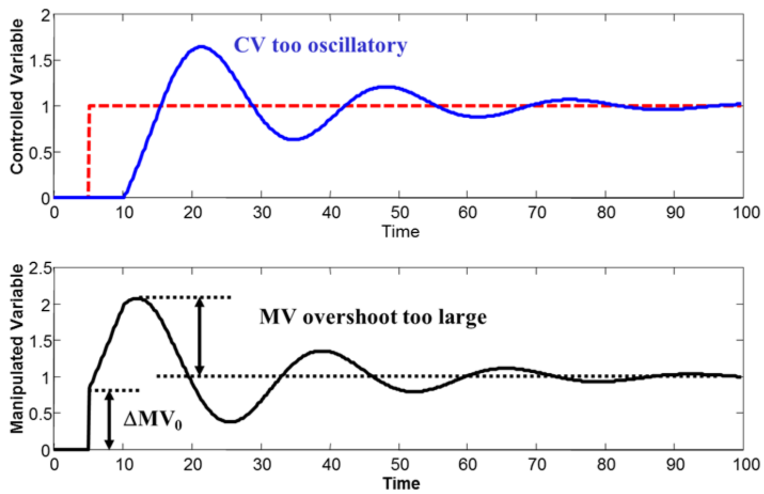


Figure 6.6 Initial control loop performance for Example 6.2

Example 6.3 Tuning for plant with variable dynamics – Generally, the PID controller will operate for months or years with one set of tuning constants. During this time, the process dynamics will change because of changes to operating conditions, like the production rate. Therefore, the tuning constant values should be the best performance for a range of operating conditions. The goal will be to minimize the IAE for a range of plant realizations (samples from the range of parameters), while observing a limitation on the MV variation.

Let's consider the following example.

Nominal process model:

$$\frac{CV(s)}{MV(s)} = \frac{K_P e^{-\theta s}}{\tau_P s + 1}$$

with the following nominal values

process gain	$= K_P$	$= 1.0$
dead time	$= \theta$	***= 5.0 minutes***
time constant	$= \tau_P$	***= 5.0 minutes***

Assume for a minute that the process dynamics were constant and known exactly. Then, the tuning constants yielding minimum IAE and the transient responses are given in Figure 6.7.

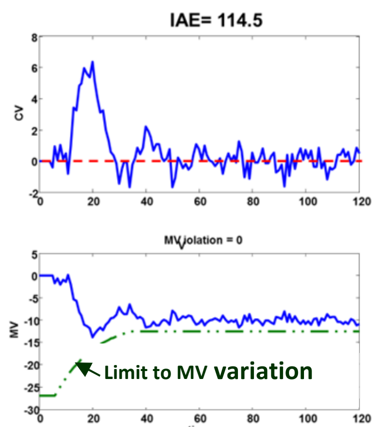
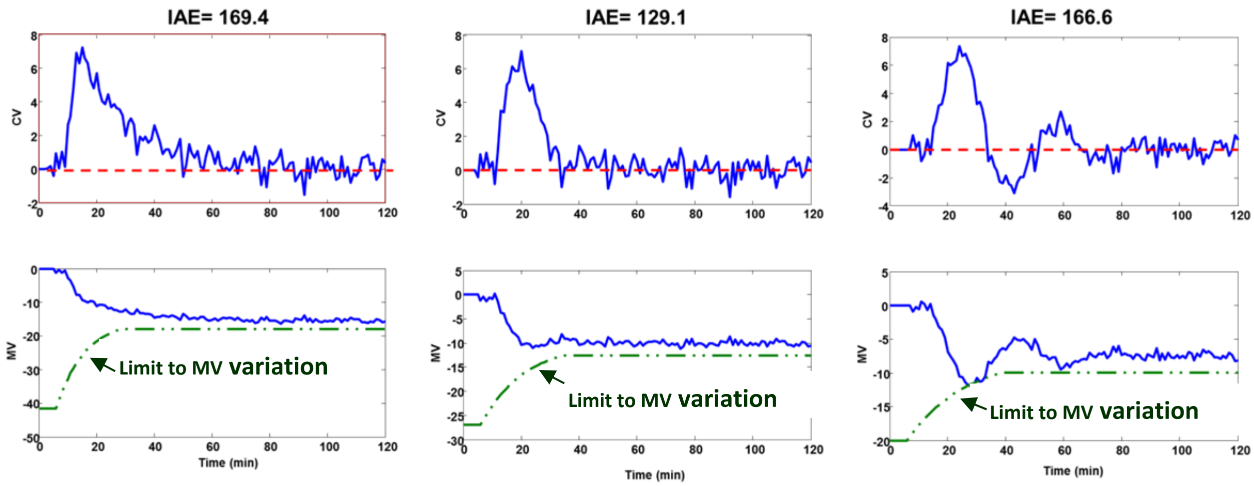


Figure 6.7 Transient response for nominal model and K_C= 1.18 and T_I = 8.59 min.

Realistically, the process conditions change. From plant operating experience, we expect that the parameters change about 35% from their nominal values and in a very highly correlated manner, as would often be the case when the process change is production rate. The best tuning and transient responses are given in Figure 6.8.



a. Plant has - 35% all parameters b. Plant has nominal parameters c. Plant has + 35% all parameters

Figure 6.8 The best tuning when ±35% dynamic model parameters are possible. Variables expressed as deviation from initial steady state. Tuning: $K_C = 0.77$ and $T_I = 7.65$ min

We observe the following from the previous three examples.

- Perfect control, maintaining the controlled variable at its set point is not possible, because of feedback dynamic delays.
- The controller can be adjusted to be more or less aggressive in its feedback compensation. In general, more aggressive feedback improves the controlled variable performance and degrades the manipulated variable performance.
- The controller tuning must be selected to match the nominal process dynamics, variability in dynamics, and control objectives.
- Very aggressive feedback can lead to undesirable oscillations and potentially, instability.

Although we will not demonstrate it here, these results are generally true for all feedback controllers.

These examples demonstrate that the major factor influencing control performance is the process dynamics and that proper controller tuning is an important factor in enabling the control loop to perform as well as possible for the process dynamics.

6.2.2 Cascade Control

When the feedback dynamics are slow and a single-loop controller cannot achieve the desired performance, single-loop control can often be enhanced to achieve much better performance with low-cost and simple technology. The two most common are cascade and feedforward control; cascade is introduced briefly in this section, with much more detail available in, for example, Marlin (2000).

In many processes, an intermediate variable exists that provides an early indication of an important disturbance, and we would like to use the early indication of a disturbance to improve control performance. Cascade accomplishes this using a hierarchical, feedback approach. Let's consider the stirred tank heating process in Figure 6.9 where the outlet temperature is controlled by adjusting the heating temperature. In this process, common disturbances include heating medium pressure variation and lack of precise valve stem positioning. Is there a measurable variable that would indicate that these disturbances have occurred? The answer is clearly "yes"; the heating medium flow. We are half way to designing a cascade controller.

Since cascade involves a hierarchy of feedback controllers, a causal relationship must exist between the control valve and the intermediate or "secondary" measured variable. By observation, we confirm the causal effect, so that a feedback controller can be implemented controlling heating flow by adjusting the valve. This controller will quickly correct for disturbances in the heating medium pressure and valve stem position errors, but it will not achieve the desired temperature control. Therefore, we direct the output of the temperature controller to the set point of the heating flow controller. This is a hierarchy and is called a cascade control system. The design is sketched in Figure 6.10, and a comparison of single-loop and cascade control performance is given in Figure 6.11; clearly, cascade control performs better, with an IAE reduction of over 90% from single-loop feedback.

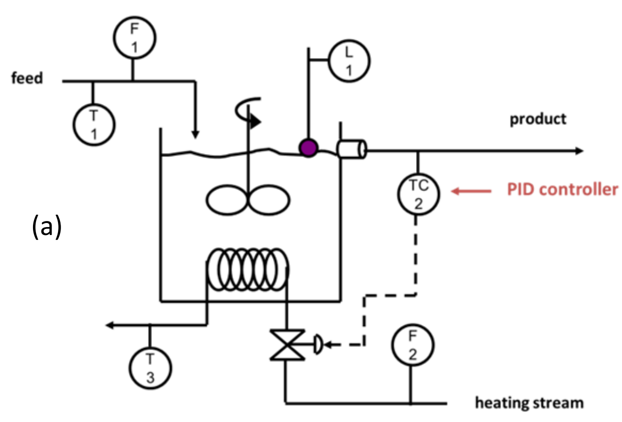


Figure 6.9. Single-loop feedback.

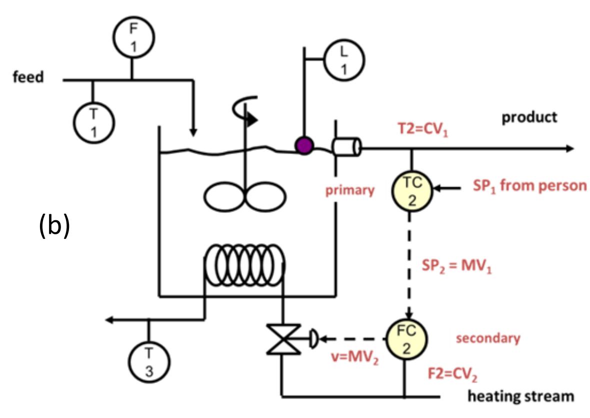


Figure 6.10 Cascade control of the stirred tank heater outlet temperature

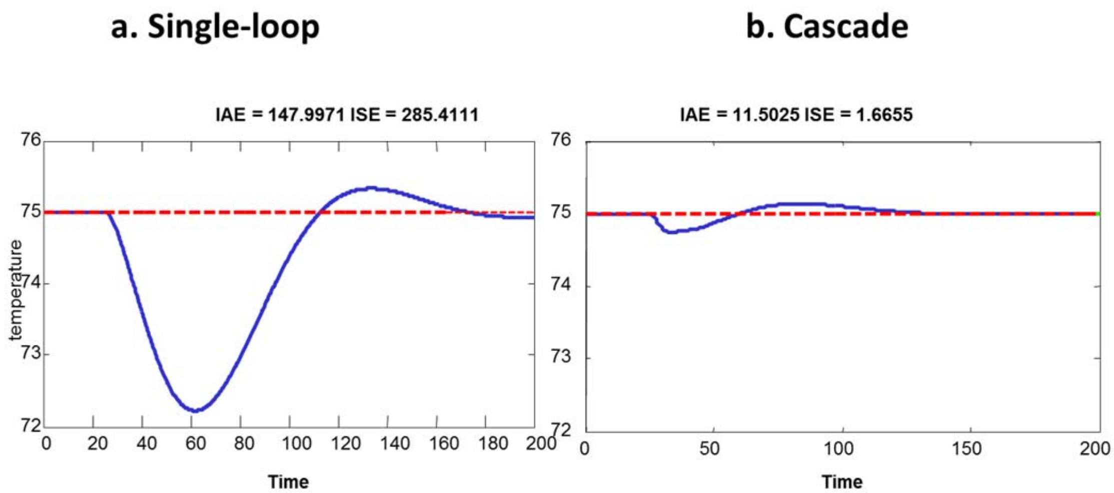


Figure 6.11 Comparison of (a) single-loop and (b) cascade control for the same process and disturbance. Cascade performs better for the same disturbance.

A properly designed cascade control system requires that certain criteria be satisfied. These five criteria are summarized in Table 6.2 that the engineer can apply when deciding whether cascade is applicable. Standard PID controllers can be used in cascade control.

6.2.3 Feedforward Control

When the feedback dynamics are slow and a single-loop controller cannot achieve the desired performance, single-loop control can often be enhanced to achieve much better performance with low-cost and simple technology. Feedforward control is introduced briefly in this section, with much more detail available in, for example, Marlin (2000).

Feedforward uses another approach for taking advantage of an early indication of an important disturbance to improve control performance. Let's again consider the stirred tank heating process in Figure 6.9 where the outlet temperature is controlled by adjusting the heating temperature. Another common disturbance involves variation in the feed temperature.

Table 6.2. Cascade design criteria
(Affirmative answer required for every entry)

	Cascade is desired when
1.	Single-loop performance unacceptable
2.	A measured secondary variable is available
	A secondary variable must
3.	Indicate the occurrence of an important disturbance
4.	Have a causal relationship from valve to secondary (cause $\rightarrow$ effect)
5.	Have a much faster response than the primary

Is there a measurable variable that would indicate that this has occurred? The answer is clearly, "yes, the feed temperature". We are half way to designing a feedforward controller. Since feedforward involves a different principle from feedback, a causal relationship must not exist between the manipulated control valve and the surrogate or "feedforward" measured disturbance variable. By observation, we confirm the absence of a causal effect.

Feedforward control uses process models to determine the valve adjustment that will exactly compensate the measured disturbance. (Naturally, "exact compensation" is the goal of the calculation, but imperfect compensation is expected because of plant-model mismatch.)

The feedforward controller compensates for the measured disturbance, but it will not achieve the desired temperature control because of model mismatch and other unmeasured disturbances. Therefore, we retain the feedback temperature controller to correct for all other disturbances and model error in the feedforward scheme. This is called a feedforward/feedback control system. The design is sketched in Figure 6.12, and a comparison of single-loop and feedforward/feedback control performances is given in Figure 6.13. Clearly, feedforward achieves a substantial improvement in control performance.

A properly designed feedforward control system requires that certain criteria be satisfied. These criteria are summarized in Table 6.3. The engineer can check a proposed design using the five criteria before completing the design. A standard PID controller can be used for the feedback control, and a special algorithm must be used for the feedforward controller. For details, see Marlin (2000).

The engineer should recognize the similarities and differences in cascade and feedforward. Use of the criteria in Tables 6.2 and 6.3 is highly recommended to ensure that a proper design is selected.

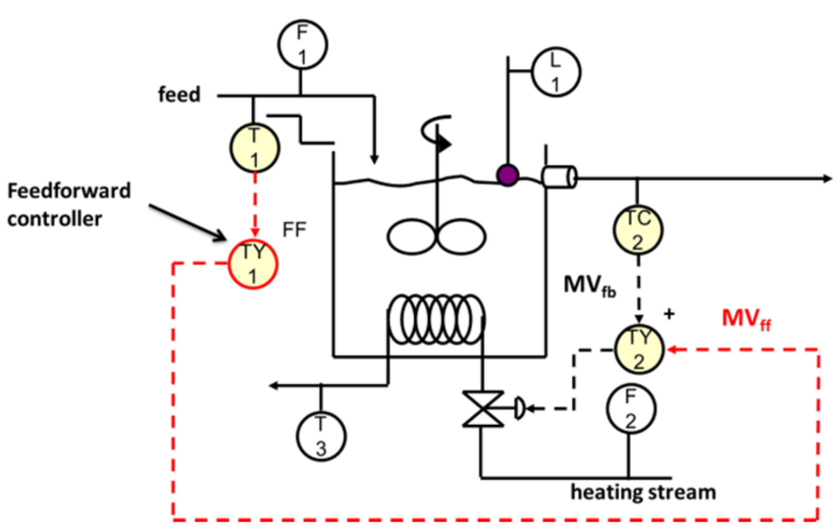


Figure 6.12 Feedforward/feedback control design

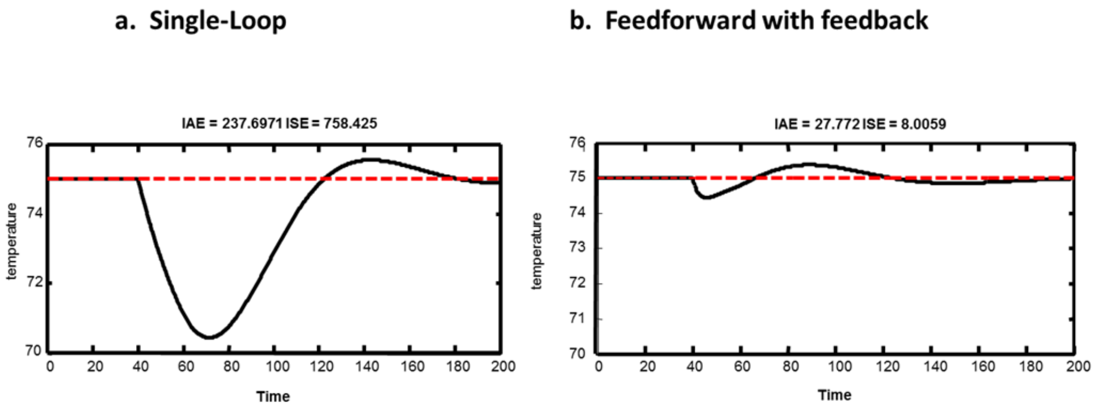


Figure 6.13 (a) Feedback and (b) Feedforward /feedback control performance

Table 6.3. Feedforward design criteria
(Affirmative answer required for every entry)

	<u>Feedforward is desired when</u>
1.	Single-loop performance unacceptable
2.	A measured disturbance variable is available
	<u>A measured disturbance variable must</u>
3.	Indicate the occurrence of an important disturbance
4.	Not have a causal relationship from valve to measured disturbance sensor
5.	Not have a much faster effect on the CV than the MV

6.2.4 Non-square process systems

Many process systems have an unequal number of controlled and manipulated variables. In this section, we will see why these situations occur and learn the basics for controlling non-square systems. We will restrict ourselves to the simplest systems with either (a) several controlled and one manipulated variables or (b) one controlled and several manipulated variables.

More Controlled variables: We encounter incidences of multiple controlled variables in everyday life. For example, we might be driving to a city and aim to arrive at exactly 3:00, but we do not want to drive faster than 65 miles per hour (about 105 km per hour). When more controlled than manipulated variables exist, it is not possible to maintain all controlled variables at their set points. Fortunately, a hierarchy of goal importances often exists, as it does in the driving example. In the driving example, we would set our speed to achieve our arrival time, unless the speed exceeded the maximum; then, we would select the maximum limit and accept a late arrival. Lets' consider a process application of this concept.

Example 6.4. Signal select The chemical reactor in Figure 6.14a has a cooling coil. The conversion in the reactor should be controlled by adjusting the cooling, but the temperature should never exceed its maximum limit (to prevent side reactions or equipment damage). We note that the coolant valve is "fail open" to provide the safest condition, maximum cooling, if the signal to the valve should go to zero. Design a control system.

We can achieve these control objectives using PID controllers. We can start by designing the composition controller, using the outlet analyzer for the measured variable and the coolant flow valve as the manipulated variable. This should work well, but it can reduce the cooling and allow the reactor temperature to exceed its maximum. Therefore, we also design a temperature controller to adjust the coolant valve. Clearly, the valve cannot obey both commands simultaneously, so when should it obey each? We note that the control objectives call for selecting the controller output that demands the most cooling and ignoring the other controller output. This is easily achieved by sending both controller outputs to a "signal select" device (or algorithm) that reads all inputs and sends an output that is the highest or lowest of the inputs. Here, we use a low signal select, because the valve is fail open, so that the smallest signal gives the largest valve opening. The control system is shown in Figure 6.14b. The dynamic performance of signal select is given in Figure 6.15.

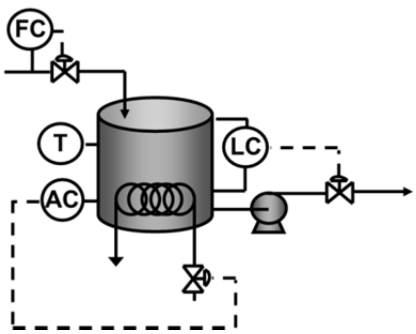


Figure 6.14a Reactor with cooling coil and single-loop control.

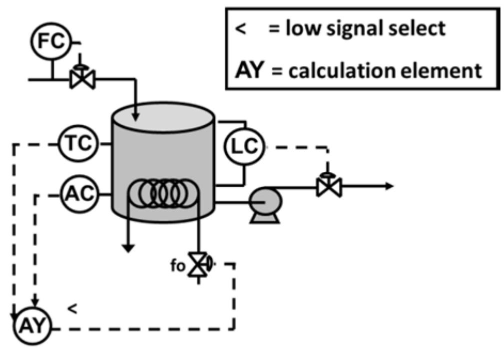


Figure 6.14b Reactor with signal-select control.

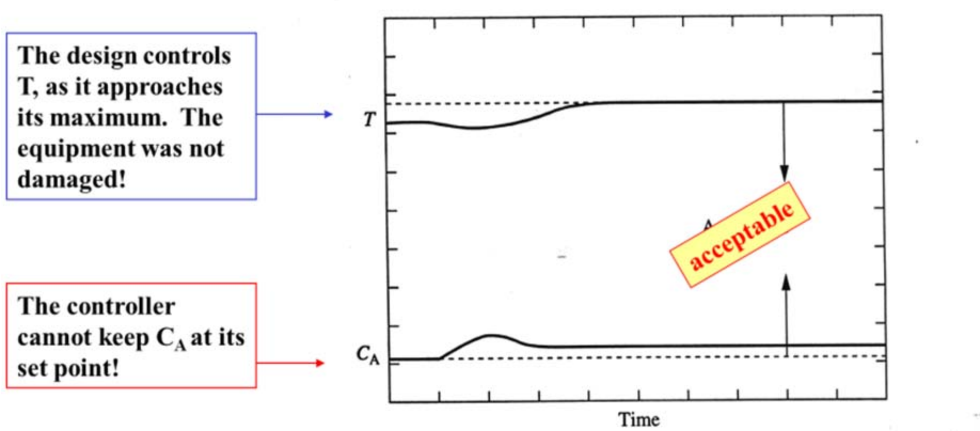


Figure 6.15 Dynamic response of the signal select control design to an unmeasured disturbance, a feed impurity that reduced the rate of reaction.

More manipulated variables: We encounter incidences of multiple manipulated variables in everyday life. For example, we want to regulate our speed at (near) 65 miles per hour by using the accelerator and the brake pedal. When more manipulated than controlled variables exist, many combinations of manipulations can regulate the controlled variable to the same values. Fortunately, an ancillary goal usually enables us to select the best operating conditions by defining an order of manipulation. In the driving example, the ancillary goal of energy conservation would indicate that we should never use the accelerator and brakes simultaneously; therefore, one would use the brake only after the accelerator is not being depressed, and one would begin to accelerate only after braking has ceased. Lets' consider a process application of this concept.

Example 6.5. Split Range The fuel gas system in Figure 6.16a has two sources of fuel and many consumers. The goal is to supply the ever-varying consumers by purchasing the appropriate amount of fuel at the lowest price. We note that Fuel A is less costly than Fuel B.

We decide to control pressure to balance the consumption and purchase of fuel gas. Since all consumers are independently determined to meet the moment-to-moment fuel needs of the process units, the purchase must be adjusted to control pressure. We would like to purchase the less expensive fuel and purchase the more expensive fuel only when required, i.e., when the less expensive is at its maximum. The pressure PID controller would split its output and send signals to both valves, which is called "split range" control. The exact details depend on the hardware implementation; here, we will assume that the signal splitting occurs in the control computer and that the fuel valves are fail closed. With the controller output scaled to 0-100 percent, the 0-50 percent values will affect only the lower cost Fuel A, with the Fuel B valve fully closed. When the controller output is 50-100 percent, the Fuel A valve will be fully opened. When the controller output is 50-100 percent, the Fuel B valve will be opened. The control strategy is shown in Figure 6.16b. The dynamic response of the system to two disturbances is shown in Figure 6.17.

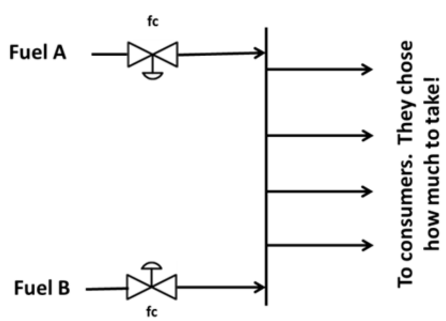


Figure 6.16a Fuel distribution system with two sources.

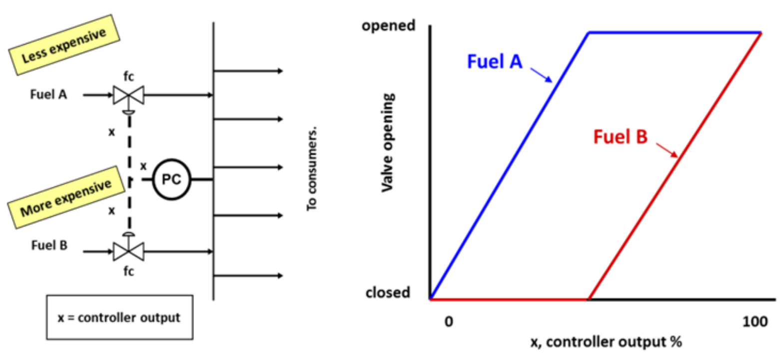


Figure 6.16b Fuel distribution system with split range pressure control.

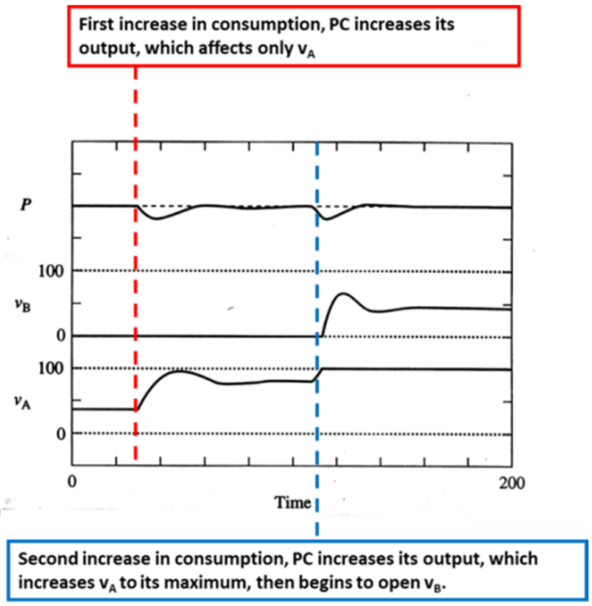


Figure 6.17 Dynamic response of the pressure split range control to two disturbances. Marlin (2000)

A general rule in control design is "never control the same variable with two controllers", which in generally true. Nevertheless, let's consider an example where controlling the same variable with two PID controllers is correct and serves as an alternative for split range control.

Example 6.6. Two controllers for one CV - The tank in Figure 6.18a accepts a flow that is controlled elsewhere in the plant. The level is to be controlled using valve v101 that regulates flow to a profitable downstream process. If the flow through v101 is inadequate to prevent overflow, flow through v102 may be used, but a severe economic penalty is incurred. Design a control strategy.

We can start with a straightforward PI (or P-only) level controller, LC101, with a set point in the middle of the vessel height; this controller will manipulate v101. If the flow through v101 is not sufficient to control the level because of a very large increase in the flow to the vessel, v102 must be adjusted. Therefore, we introduce PI level controller LC102 that could use the same or redundant level sensor. LC102 has a set point of 90% and manipulates v102. The key to this design is the difference in the level controller set points; the two controllers must have significantly different set point values. The completed design is shown in Figure 6.18b.

In most introductory process control courses non-square process systems are not discussed. Is this because they are rare? No, non-square systems are common and are designed for the following reasons.

- Extra controlled variables provide actions to avoid violating important limitations. This can prevent equipment damage and severe process disturbances.
- Extra manipulated variables provide an "expanded operating window", so that the process can function well over a larger range of set point changes and disturbances.
- Extra sensors and manipulated variables increase the reliability of the process and control system.

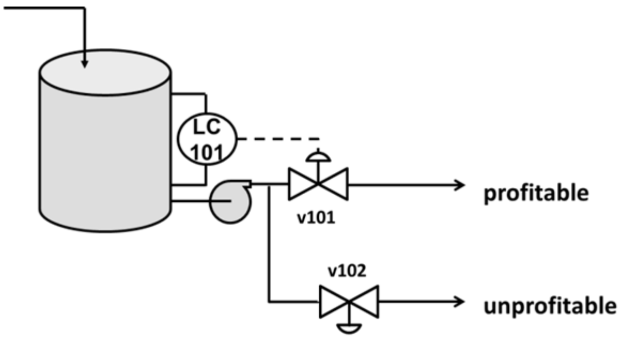


Figure 6.18a Single level controller. Operator action is required when valve v101 is fully open. (Split range control could be applied.)

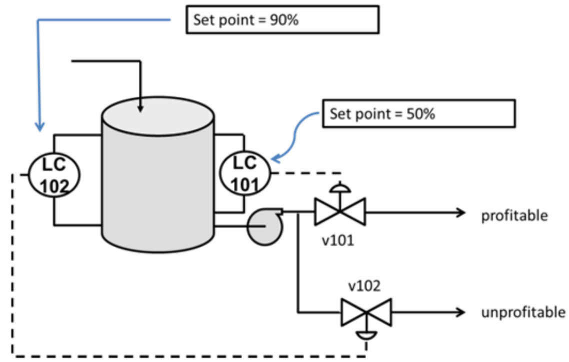


Figure 6.18b Two level controllers

When should the process and control system provide the additional flexibility? Designs provide the additional flexibility when the economics and safety benefits justify the added investment. These features are not included in most units in a plant, but large plants will typically have many instances of non-square systems.

6.2.4 Inferential control systems

In all of the control methods considered to this point, the important variables have been measured, a situation that is desirable and most often possible. However, not every important variable can be measured in real time, i.e., fast enough so that timely control actions can be based on their measurements. There are various reasons for the lack of key measurements. First, some sensitive analyses have not been sufficiently automated to provide accurate, reliable measurements without human management; thus, these measurements can be obtained only infrequently in a laboratory. Second, if the real-time measurement is possible the cost of installing a sensor in the plant may not be justified by the potential benefits derived from the additional sensor, especially considering the alternative methods in this section. The cost is not typically high for conventional sensors for measuring temperature, pressure, flow, and level but may be prohibitive for an expensive analyzer with sample system and on-going maintenance. Third, the sensor may not provide information in a timely manner if it must be located far downstream or it may have a long delay due to processing time.

The lack of measurements of key variables in a timely manner certainly makes automated feedback control more difficult, but not always impossible. In some situations, we can add extra measured variables that, while not giving a perfect indication of the key unmeasured variable, provide a valuable inference. For example, a temperature rise along a packed bed reactor might be very useful in determining the conversion of a single reactant to a single product. However, the temperature rise might not be useful in a packed bed reactor for determining the distribution of many reaction products. Let's consider a typical example of inferential control.

Example 6.7. Inferential composition. Suppose that we wish to regulate the multicomponent flash separation process in Figure 6.19. In the process, a liquid has its temperature increased by heat exchange and subsequently has its pressure reduced across a valve. The resulting two-phase flow enters a drum, where the liquid and vapor are separated and exit in by the bottoms and overhead pipes, respectively. The goals for control are given in the following.

- Safety, which we will achieve by preventing excess pressure
- Equipment protection, which we will achieving by ensuring that liquid flows through the pump
- Production rate by controlling the feed flow
- Composition of the light key ethane in the liquid product. The desired value is 10% ethane, and the allowable range is from 9 to 12 %.
- Relevant disturbances are the temperature measurement error bounded by ± 0.50 °C and changes in feed butane composition of ± 5 mole% occurring simultaneously with a change in methane of equal magnitude and opposite sign. (While we consider two disturbances here, many more might be relevant in an industrial process.)

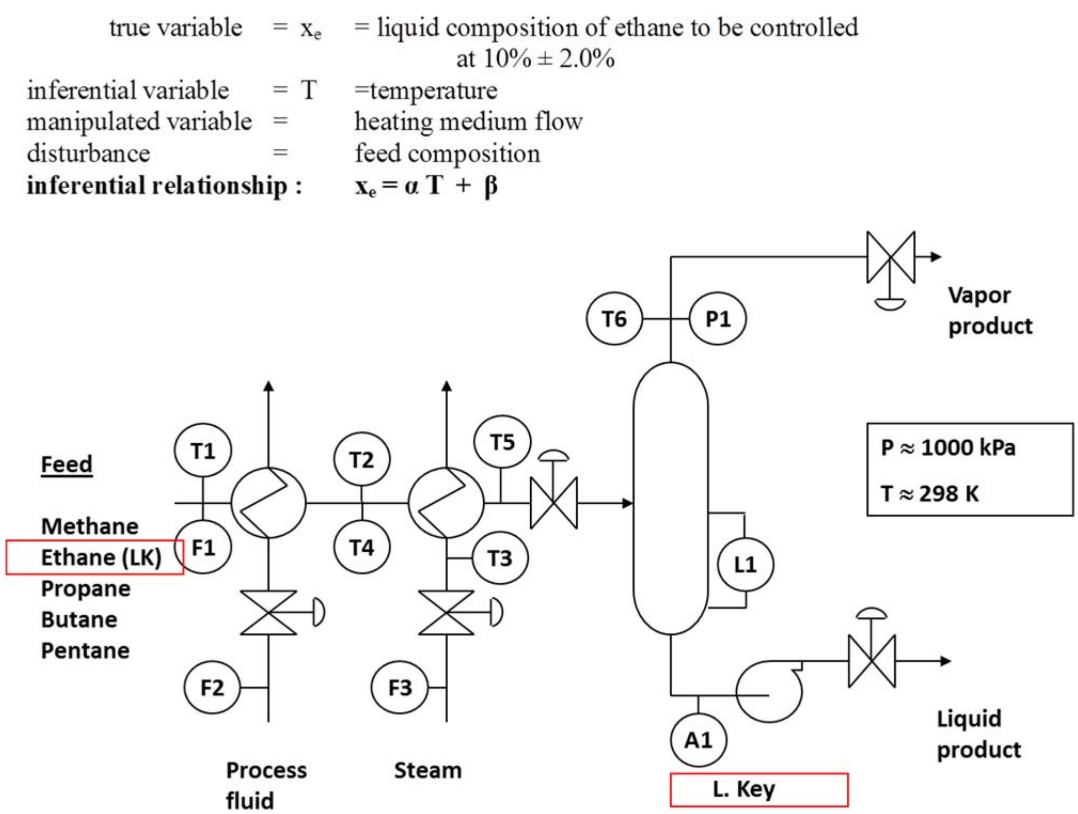


Figure 6.19 Typical flash separation process. (Marlin, 2000)

We investigate the best manner for achieving the final goal. One possibility would be to install an on-stream analyzer, likely a gas chromatograph that extracts samples from the bottoms stream and determines the percentage ethane in the sample. Perhaps, the economics do not justify the on-stream analyzer. Therefore, we will seek an inferential variable. From our knowledge of the flash process, we know that three factors influence the compositions: temperature, pressure and feed composition. The feed composition is not measured (and would be as expensive as the product measurement), so it is eliminated from consideration. The pressure has a strong effect on the compositions; however, changing the pressure of the vessel requires a more costly vessel to accommodate a wide range of pressures; therefore, we do not select pressure. Now, we proceed to determine whether the flash temperature is an appropriate inferential variable for this process equipment, operating conditions, and control goals.

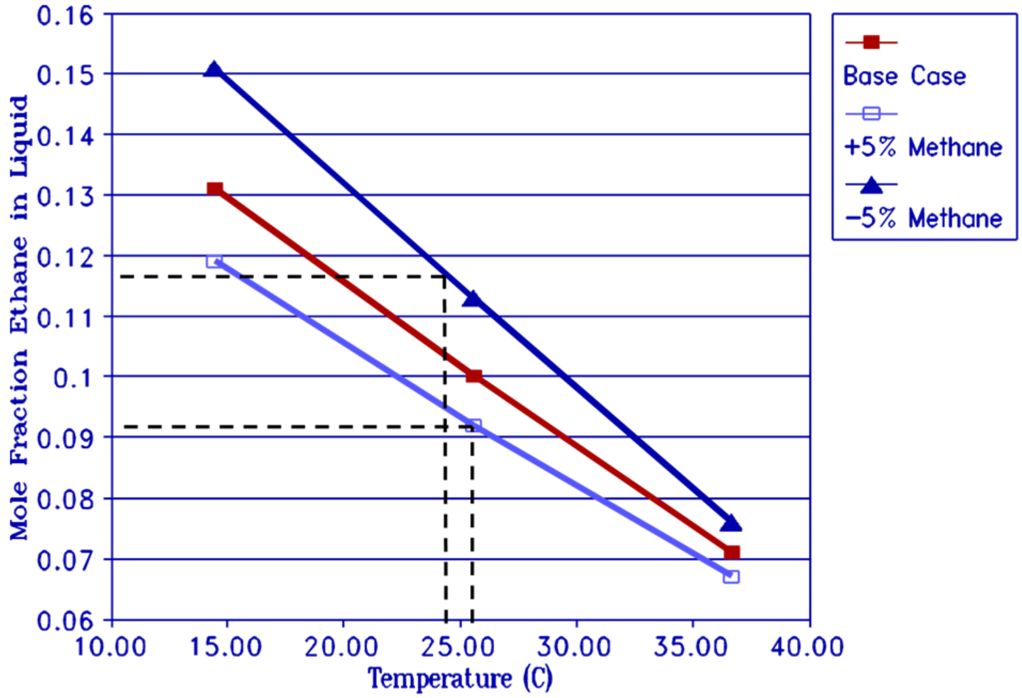
First, we evaluate the relationship between the candidate inferential variable and the true controlled variable, which is shown as the base case in Figure 6.20 at the base case pressure and feed composition. We see that the relationship is strong, especially in comparison to the measurement error in temperature.

Second, we determine whether the disturbances influence the composition-temperature relationship. The results in Figure 6.20 display the effects of temperature measurement error and the feed composition disturbance. As we would expect, the disturbances introduce errors into the assumed single-variable relationship. The key question is, "Is the inferential relationship, while not exact, good enough to achieve the performance goal?" The analysis to answer the key question is shown graphically in Figure 6.20, where the effects on the estimated composition of the maximum sensor error and composition disturbances are evaluated. We see that the actual composition will remain within the stated goal (9 to 12 percent) when maintaining the measured temperature at 25 °C.

Third, we evaluate the dynamics of the potential temperature control loop. The temperature can be measured with minimal delay, and the steam to the preheater can be adjusted rapidly, so that the closed-loop dynamics would be on the order of minutes. Therefore, the loop dynamics are fast enough for the application.

Based on this analysis, we conclude that the T6 inferential variable can be used in this process application. Naturally, other disturbances or a tighter goal for composition deviations might change this conclusion.

An inferential variable can be employed to improve control performance even when the true controlled variable can be measured, but with significant delay. As shown in Figure 6.21, an analyzer has been installed downstream to measure the butane concentration in the propane product. At essentially no cost, the same analyzer could measure the ethane concentration for feedback to the flash process! Because of the long delay from the flash drum to the analyzer, the inferential temperature control would be retained, with its set point being adjusted by the analyzer feedback in a cascade design.



Changes in methane are accompanied by changes in butane of equal magnitude and opposite sign. Pressure is constant at the base case value.

Figure 6.20. Analysis of the inferential relationship for the flash process at 1000kPa. From Marlin (2000)

The general criteria for successful inferential control are summarized in Table 6.4. The application of these criteria requires process insight from the engineer to develop reasonable candidate inferential measurements.

The use of an inferential variable must be tailored to the specific process application. Applications of similar concepts on unit operations must be evaluated because of differing control goals, disturbances, inferential sensor accuracy, materials and operating conditions.

Much more can be learned about the important topic of inferential control. Further process examples of single measured inferential variables and building calculated inferential variables (using multiple measurements) are available in Marlin (2000). An alternative method for designing inferential variables utilizes the enormous amount of information available in plant operating data. This data can be used for determining key correlation relationships among process variables that provides a basis for building inferential control variables. A good introduction to using plant data for inferential design is given in Kresta et. al. (1994).

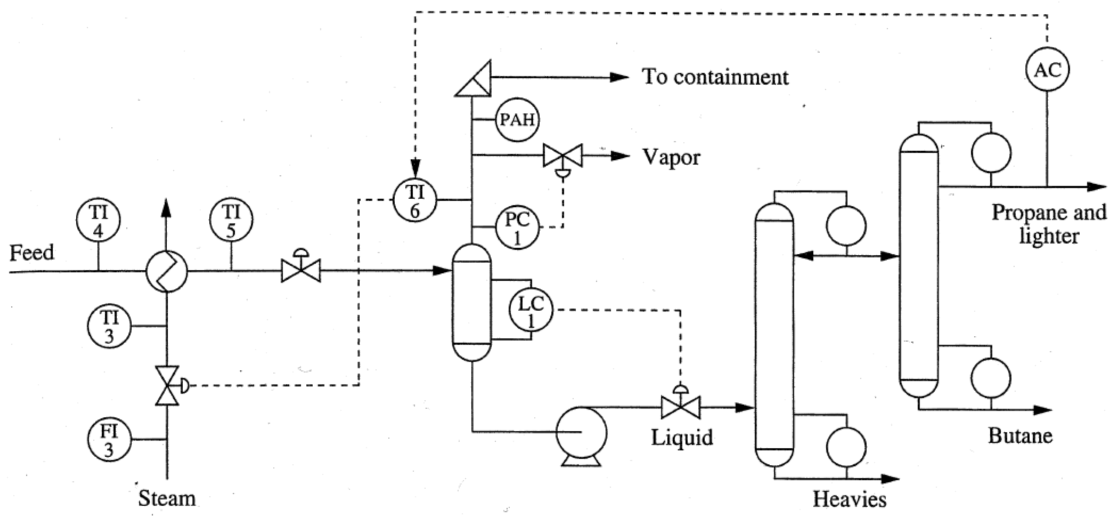


Figure 6.21 Inferential control combined with analyzer feedback with substantial delay. Marlin(2000)

Table 6.4 Criteria for designing inferential control systems

Necessary situation
• Measurement of the true controlled variable is not available in a timely manner - Not measured: on-stream sensor not possible or unreliable - Not measured: on-stream sensor too costly - Unfavorable feedback dynamics: sensor has poor dynamics, e.g., long dead time or analysis time, or is located far downstream • Measured inferential variable(s) is available
Inferential variable features
• The inferential variable must have a good relationship to the true controlled variable for changes in the manipulated variable • Relationship above is insensitive to changes in operating conditions, i.e., unmeasured disturbances, over their expected ranges • Favorable (fast) dynamics for use in feedback control
Correction of inferential variable
• By primary controller in automated cascade design • By plant operator manually based on periodic information • When inferential variable is corrected frequently, the sensor for the inferential variable must provide good reproducibility, not necessarily high accuracy

6.3 The influence of process design on control

The process being controlled has a profound effect on our ability to apply control and on the quality of the control performance. For example, we could choose to drive the bus or the ride bicycle depicted in Figure 6.22. If we need to implement a command for a direction change,

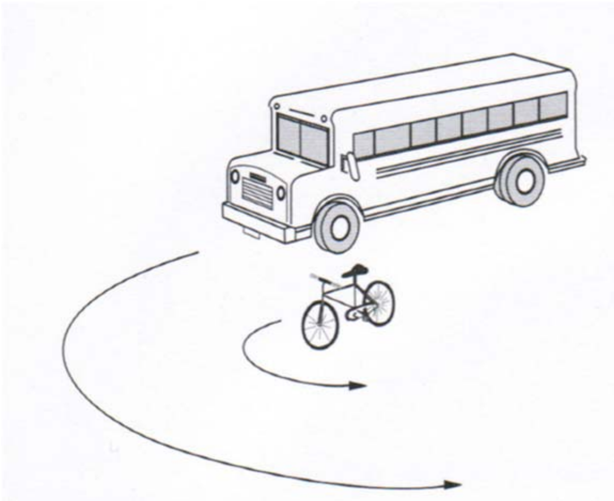


Figure 6.22 Control of different vehicles. (from Marlin (2000))

specifically to execute a 180-degree turn, which vehicle would be superior? Naturally, the bicycle could turn more rapidly and in a small diameter, so it would be superior. Now, let's consider another situation; consider both vehicles travelling at the same speed and encountering a large bump in the road. In this new situation, the bus is superior because it is less sensitive to the disturbance. Similarly, the design of chemical processes can have profound effects on dynamic performance.

Chemical engineers design processes that are safe and reliable, satisfy production requirements (product qualities, production rates, etc.) over a range of conditions, have good steady-state efficiency, and have dynamic behavior that favors good dynamic control performance.

In this section, a number of process characteristics will be introduced that affect control performance. Each characteristic can have favorable or unfavorable effects of control performance depending on the details of the process design. This does not imply that all characteristics can be designed for favorable performance because the process chemistry and physics set requirements that override dynamic behavior is some cases. For example, a reactor volume is required to achieve a specific conversion and yield, even if a smaller volume would be easier to control. However, much opportunity exists for the design engineer to improve (or degrade) the operability and dynamic behavior of the process.

This section will concentrate on characteristics that affect a single controlled variable. The characteristics are summarized in Table 6.5. In general, the conclusions presented here will extend to multivariable processes as well. Important aspects of multivariable control are addressed in Section 6.4.

Table 6.5 Process characteristics that influence control performance

Characteristic	Favorable	Unfavorable
1. Process equipment in the feedback loop		
a. Dead time	Short	Long
b. Inverse response	None	Large
c. Time constant(s)	Small, few	Large, many
d. Limitation in the manipulation rate of change	None	Slow compared with needed dynamic response
e. Limitation to manipulation range	None	Small, limits achievable steady-states
f. Process non-linearity	Minimal	Significant, affects damping and stability
g. Sensitivity to manipulation	**Moderate** (~1% CV/1% manipulation)	**Too low** (small CV range) **Too large** (very small manipulation needed)
h. Process stability without control	Stable	Unstable
2. Process elements in (only) the disturbance path		
a. Disturbance time constant(s)	Many, large	Few, small
b. Disturbance frequency	**Small or large period** compared with feedback dynamics	**Near critical frequency** of the feedback loop
c. Disturbance magnitude	Small	Large
3. Instrumentation elements in the feedback loop		
a. Sensor and final element dynamics	Fast compared with feedback process	Slow compared with feedback process
b. Measurement noise	Small magnitude compared with allowable variation	Large magnitude compared with allowable variation
c. Non-ideal final element behavior	Element follows controller output closely	Large deviations between element and controller output
d. Controller execution period	Short compared with feedback dynamics	Long compared with feedback dynamics
4. Process structure		
a. Feedback from integrated processes	Negative feedback	Positive feedback
b. Interaction among control loops* (Control structure also a factor)	Small	Strong
5. Control performance goals		
a. Quality specifications	Large short-term variation allowed	Small short-term variation allowed
b. Penalty for constraint violation	Small	Large
c. Production rate specifications	Specification on average, short-term fluctuations allowed	Must meet demands immediately
d. Profitability sensitivity for different feasible operations	Small	Large
e. Safety	Safety is paramount. See Chapter 7 for safety hierarchy.	

* The entries for interaction are simplified. Some designs with strong interaction can yield good control performance. See Section 6.4

The structure of Table 6.5 indicates that the location of some characteristics, in the feedback or disturbance path, is critical to the characteristic's effect on feedback control performance. The distinction is shown schematically in Figure 6.23 using both a block diagram and a simple process drawing. The block diagram shows the distinction between the two paths, the feedback path is every element in the loop, including the temperature sensor, the control valve v3, tank 2 liquid volume, and the coiled heat exchanger. The disturbance path is external to the feedback, e.g., tank 1 liquid volume, and does not affect stability.

6.3.1 Process Equipment in the Feedback Loop

The feedback loop performance is improved by fast and strong effects of the manipulated variable on the controlled variable. Let's look the characteristics in the feedback loop.

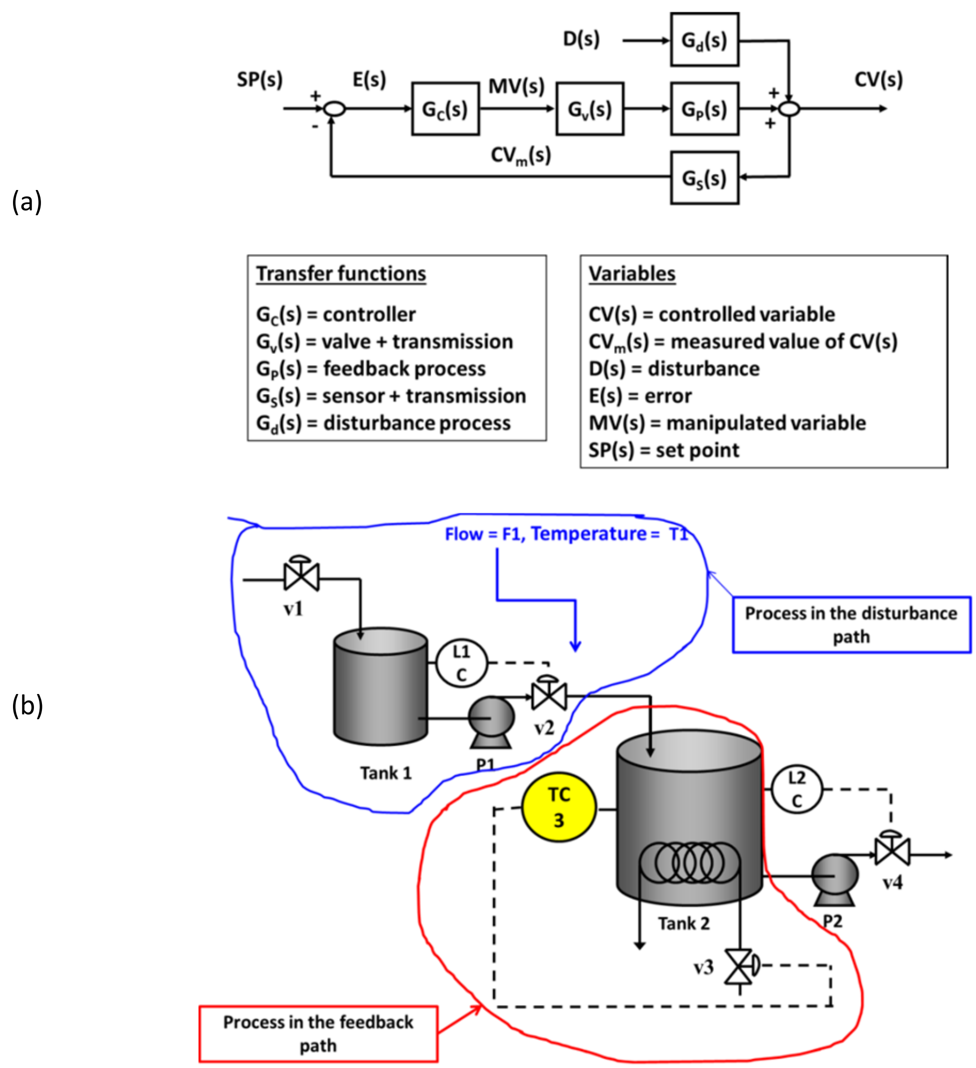


Figure 6.23 Single-loop control system for TC-3. a. Block diagram, b. Process schematic

6.3.1a Dead time – During the dead time, the effect of a change in the manipulated variable does not influence the controlled variable, as shown in Figure 6.4a. This delay in information degrades feedback control performance, as shown clearly in Figure 6.4b. The red box shows an area of controlled variable deviation from set point that *no feedback control algorithm can reduce*; the cause is the process dead time. The engineer should make every effort to reduce or remove dead time from the feedback loop.

Feedback control performance is improved by process designs with short dead times in the feedback path.

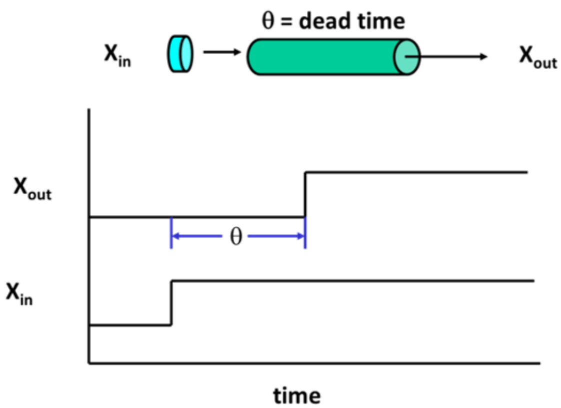


Figure 6.24a Response of a dead time to a step input

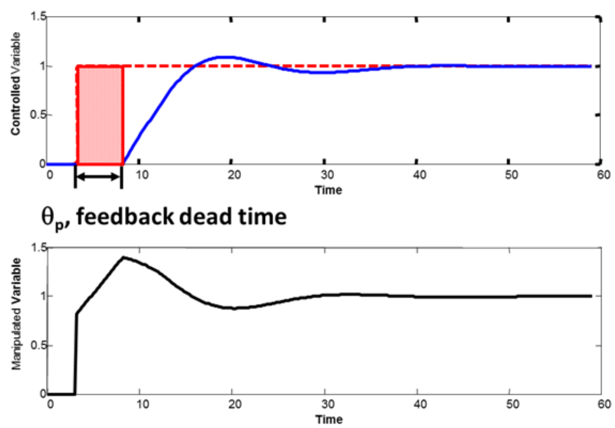


Figure 6.24b. Effect of dead time on feedback control performance

6.3.1b Inverse response – An inverse response occurs when the manipulated variable has two effects on the controlled variable. One effect is faster and has a smaller magnitude positive (negative) gain, while the other path is slower and has a larger magnitude negative (positive) gain. A typical inverse response is shown in Figure 6.25a. This "wrong-way" initial response delays return to desired operation, and therefore, degrades feedback control performance. Feedback performance is shown in Figure 6.25b. For further details on modeling inverse response behavior, see Marlin (2000) Chapter 5 and Appendix I.

Feedback control performance is improved by process designs without inverse response in the feedback path.

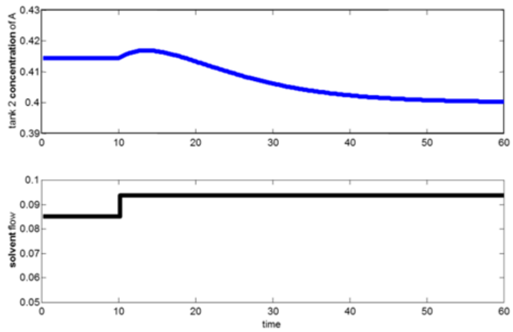


Figure 6.25a Typical open-loop inverse response

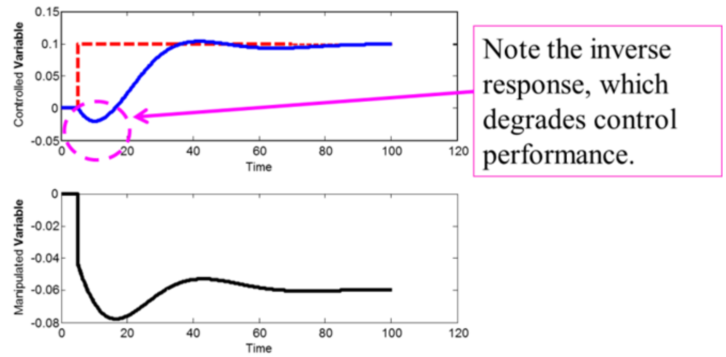


Figure 6.25b Typical closed-loop inverse response for a set point change

6.3.1c Time constants – One can think of each time constant being the result of a dynamic balance, e.g., material or energy balance, for the process. When modeled using fundamental balances, typical process systems consist of many balances and thus time constants. For example, a distillation tray has one energy balance and as many material balances as components; therefore, a distillation tower model has many balances. When we model the same process empirically, we usually find that one or a few time constants can represent the process adequately for the purpose to selecting a control strategy and tuning the controller.

Naturally, these time constants introduce delays in the feedback loop. In additional, numerous series time constants can introduce a behavior very similar to dead time, which we have seen is particularly deleterious for feedback. This effect is shown in Figure 6.26. Therefore, design engineers should reduce time constants in the feedback path, when possible.

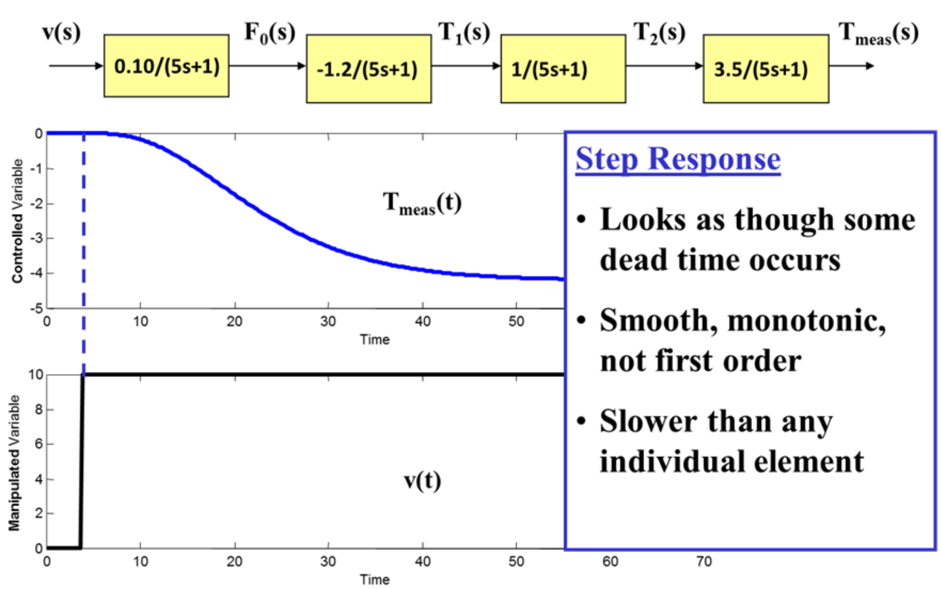


Figure 6.26 Step response for time constants in series.

Feedback control performance is improved by process designs with few and small-valued time constants in the feedback path.

6.3.1d Manipulated variable rate of change – Process equipment is designed for extreme conditions, e.g., high temperature and pressures, and appears very sturdy. One might wonder why we need to observe limitations to the maximum rate of change of manipulated variables. Let's look at an example of a fired heater In Figure 6.27, in which the outlet temperature of the process stream is controlled by adjusting the fuel flow to a burner. The process is typically operating near its material limitations, and exceeding the maximum temperatures can reduce the operating life of brickwork and piping. In addition, changes to the fuel firing can cause thermal expansion and contractions that reduce equipment life. Therefore, the process controller must balance the needs of maintaining the controlled temperature near its set point with the need to limit manipulation velocity and extend equipment life. This balance is achieved through appropriate controller tuning.

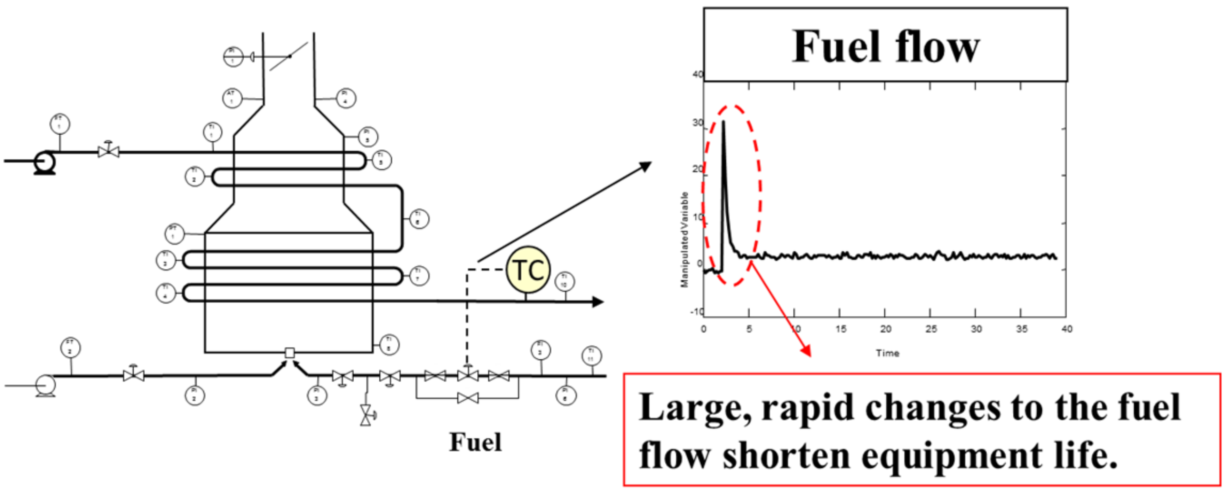


Figure 6.27 Aggressive manipulations can reduce equipment service factor.

Feedback control performance is limited by the allowable rate of change of the manipulated variable. Design engineers must recognize where these limitation are required and in these situations, not expect to achieve excellent controlled variable performance through very aggressive manipulations.

6.3.1e Manipulated variable range – Naturally, a limitation to the range of manipulation reduces the range of achievable set points and compensation for disturbances. This limitation affects the steady-state operating window, but it can be overcome if the design engineer provides additional manipulated variable(s) and a control system for this "non-square" system.

The range can also affect the dynamic response of a control system. Let's consider a scenario in which you have rushed home and are preparing soup for dinner. You are in a rush because you want to get working on your process control assignment. (Apparently, this scenario is hypothetical.) How would you adjust the heat to the burner? If you turned the burner adjustment to the value that just brought the soup to a slight boil, which is the desired end point, you would have to wait a long time. You could follow an alternative strategy in which you turned the burner to maximum heat and when the soup began to boil, reduced the burner adjustment accordingly. This second strategy has a faster response, which is possible because the burner has "spare capacity". This concept can be applied in process control equipment design. The feedback controller tuning can be selected to provide some manipulated variable overshoot during the transient response. Naturally, the aggressiveness required for rapid controlled variable response must be balanced with the potential equipment damage due to overly aggressive manipulated variable behavior. The control engineer must understand process and equipment operating goals!

Feedback control is limited by the range of the manipulated variable. For steady-state behavior, returning the controlled variable to its set point is possible when the range of the manipulated variable is sufficiently large and is not possible outside of the operating window. For dynamic behavior, the feedback response can be rapid if the equipment is capable of manipulated variable overshoot

6.3.1f Process Non-linearity – We recognize that the controller algorithm and tuning values are constant in most control designs. The engineer adjusts tuning once and the control system functions for a long time. This approach can give acceptable dynamic performance when the feedback behavior is also unchanging. Since essentially all processes are non-linear to some extent, we have to evaluate when non-linearity – causing changing feedback process behavior – is insignificant and when significant. This analysis is specific to each control loop. However, for a rough guideline a feedback controller should function reasonably well (but not optimally) for a change of ± 25% in all parameters, e.g., dead time, time constant and gain. This is a minimum change that a process can be expected to experience, and much larger changes can occur in some processes. We have seen an example of a closed-loop behavior for a single controller with ± 35% that can be achieved with PI control in Figure 6.8.

Some chemical processes are very non-linear and are challenging to control. An example is the process for mixing a strong acid and base shown in Figure 6.28. When the pH set point is 7.0, the process is strongly non-linear, with the gain changing by orders of magnitude! This is a challenging problem. When tight control is required, a typical design provides two mixing tanks and sizes them to attenuate high frequency fluctuations (Hoyle, 1976; Moore, 1978). When larger, short-term fluctuations are acceptable, one tank can be used.

An approach to compensate for process gain non-linearity is to include a correcting non-linearity in the feedback loop. If this is designed well, the product of the two non-linearities can be approximately linear input-output behavior. The PID controller gain can be adjusted to account for process non-linearities; for an example applied to pH control, see Liptak (2003).

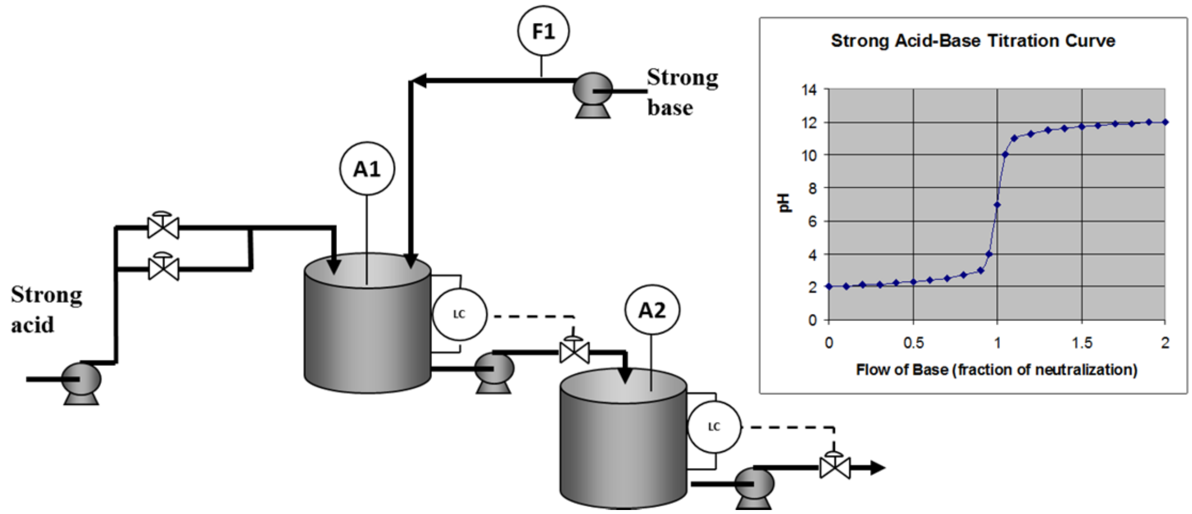


Figure 6.28. pH neutralization mixing process.

Control performance over a range of operating conditions, including disturbances, is improved by a linear feedback loop. Non-linearities can degrade performance, and for significant process non-linearities, the design engineer should evaluate including compensating non-linearities in the control equipment or calculation.

6.3.1g Sensitivity to manipulation – The gain relationship between the manipulated and controlled variables is clearly important. To have a general discussion, we will consider the variables expressed as percentage of range. For controlled variables, the range will be the sensor range, which should span the values expected under "normal variation", excluding startup and extreme disturbances due to, for example, a pump failure. The manipulated variable is expressed as 0-100% valve opening (closing), or if the output is to a secondary in a cascade, as percent of the secondary sensor range. With scaled variables, the process gain is expressed in units of %CV/%MV.

A desired value for the process gain is around 1 %CV/%MV. With this value, the control variable can be directed to the entire range of the sensor span by adjusting the manipulated variable through its entire range. If the process gain value is too small, the manipulated variable has a weak effect and cannot compensate larger disturbances. If the process gain value is too large, the lack of precision in adjusting a valve will lead to poor control. (It is not possible to adjust the valve from 65.2% to 65.4% precisely.) The situations with inappropriate process gains are shown in Figure 6.29.

For good control performance, the process gain should be approximately 1 %CV/%MV. If the value is too low, the design engineer should provide an additional manipulated variable with wider range. If the value is too high, the design engineer should include a manipulated variable with a smaller gain.

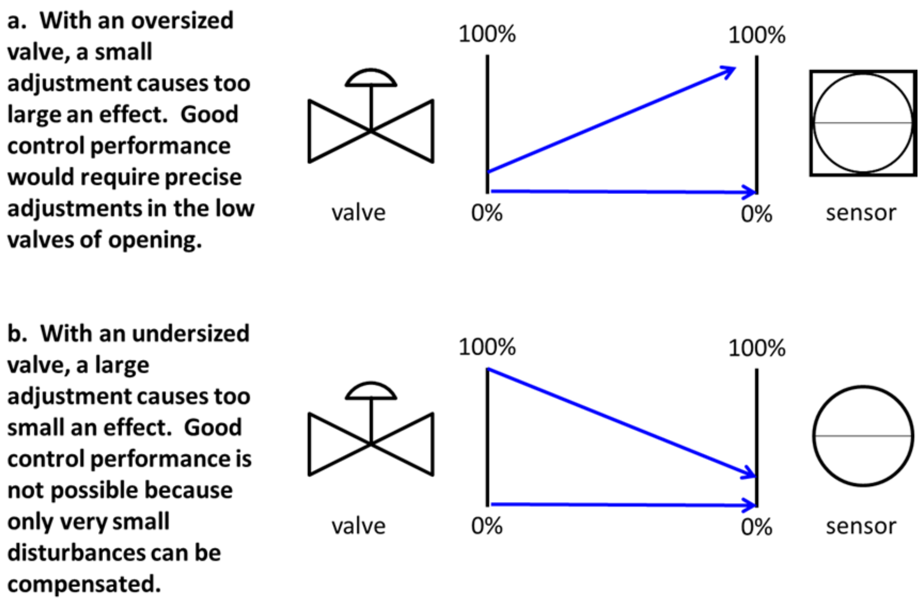


Figure 6.29 Schematic of effect of process gain ($K_P = \Delta CV/\Delta MV$).
a. Gain is too large, b. Gain is too small

6.3.1h Process stability without control – One final effect of the process on dynamic performance is the stability of the process without control. Most processes are stable and therefore, will approach a steady state when all input variables, i.e. disturbances and manipulated final elements, are constant. However, some important processes are not stable.

A stable process has output variable(s) that are bounded in magnitude when all input variables (disturbances and manipulated variables) have bounded magnitudes.

We see that the definition of stability considers whether a process variable is unbounded and tends toward infinity. From an engineering viewpoint, no variable gets near infinity because something very bad (equipment damage, explosion, and so forth) will occur in the plant before infinity. However, the concept of stability is useful because an unstable variable will not tend to a steady state, and operating personnel would have to spend an inordinate amount of time monitoring and adjusting the process for each unstable variable. Therefore, we anticipate that unstable variables must be controlled.

Inventories - One important category of unstable process variables is inventories, both liquid and solid, that serve to store materials without changing the material properties. A simple material balance model for the inventory shown in Figure 6.30a is given in the following.

$$\frac{dm}{dt} = A\rho\frac{dL}{dt} = \rho F_{in} - \rho F_{out} \tag{6.5}$$

with

A	=	cross sectional area (assumed constant)
F	=	volumetric flow rate
L	=	level
m	=	mass of material in the vessel
ρ	=	density

When the flows in and out do not depend on the inventory, the system is termed “non-self-regulatory”. Non-self-regulatory inventories have all inputs and outputs independent of the inventory itself; therefore, as the inventory changes (increases or decreases) no change occurs to an input or output that might stabilize the inventory. Without intervention by people or controllers, a non-self-regulatory system is unstable. As an example, consider a situation in which the flows in and out are initially equal, and the outlet flow increases. Then, the derivative of level would be a negative constant. The dynamic response for a step change in the flow out from an initial steady state is given in Figure 6.30b. A mathematician would say that the inventory would decrease “without limit”; in contrast, an engineer would say that the inventory would increase until the occurrence of an incident, such as a zero level. The unstable behavior is clearly demonstrated.

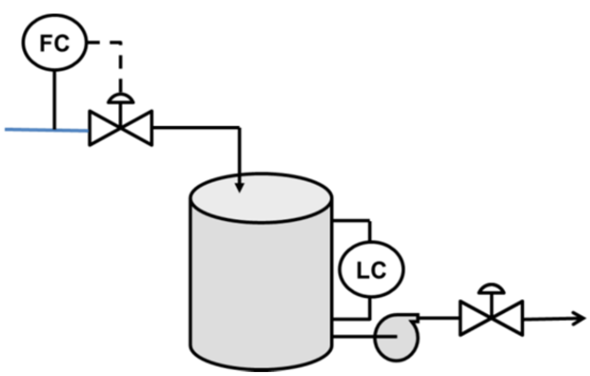


Figure 6.30a. Liquid inventory with flow out pumped.

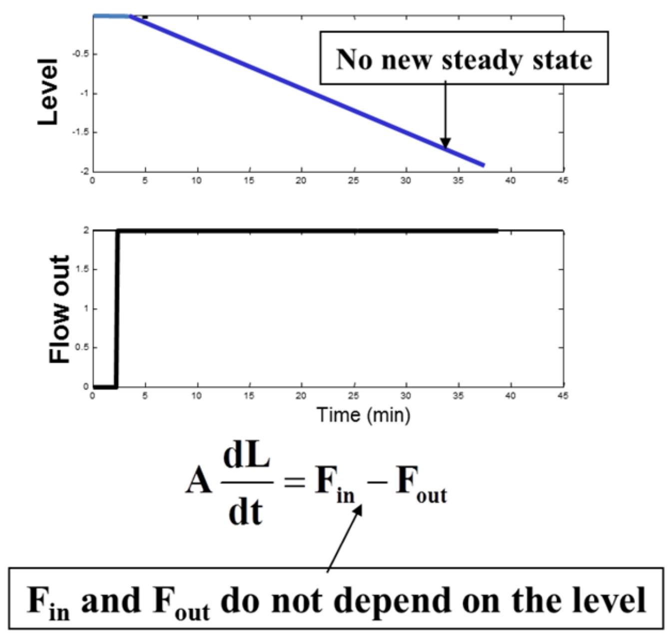


Figure 6.30b. Step response for a non-self-regulating level, clearly showing instability.

Fortunately, control of inventories is generally quite easy, because essentially no dead time exists between the level and potential manipulated variables, the flows in and out. Therefore, a feedback controller can maintain the level near its set point or can allow it to vary about the set point to modulate the manipulated flow variations. Details on level modeling, control and tuning for various process objectives are available in Marlin (2000).

Should unstable levels be controlled? Let's consider some situations.

- Large tanks with only one continuous flow rate. Raw material feed tanks have periodic inflows during deliveries but can have continuous outflows to the process. Finished product tanks can have continuous inflows from the process but only periodic outflows for shipment to customers. These tank levels cannot be controlled. The design engineer must provide adequate storage volumes to allow continuous process operation between deliveries and shipments, including consideration for disruptions in transportation due to weather and other factors.
- Large tanks with continuous flows in and out. These tanks can store material between a sequence of process units. This material can be termed "intermediate products" or "work in progress". The size of the tanks can be large to provide reliability, e.g., time for maintenance, as discussed in Chapter 4. Usually, the daily production rates of the individual units are set independently, although they must balance over a long period of time. Therefore, these tank levels are not usually controlled. When designed with adequate volumes, operating personnel have sufficient time to monitor and manage the flows to maintain the levels in acceptable ranges.
- Chemical processes have many smaller tanks and drums, such as reflux drums, vapor-liquid separators, kettle reboilers, and so forth. The ratio of inventory to flow rate (V/F) is usually in the range or 5 to 15 minutes; therefore, disturbances can rapidly lead to overflow or emptying of the vessel. All of these liquid inventories must be controlled.
- Many chemical reactions occur in a liquid inventory. These reactors are different from storage facilities because the reactor volumes influence the process operation (Fogler, 1986). These reactor levels must be controlled because they are unstable and influence the product quality and process profitability.

In summary, inventories with very large volumes are not controlled automatically; people provide the feedback action. For smaller inventories, people cannot reliably monitor the process, so that automatic feedback control is essential. In addition, when the volume of the inventory affects the process performance (or safety), with chemical reactors being the most common example, feedback inventory control is essential.

Unstable processes - A second category of unstable processes involves processes that have self-regulation. Most self-regulatory processes are stable; however, some self-regulatory processes can be unstable for specific design parameters. An example is a CSTR with an exothermic reaction and cooling, which can experience multiple steady states and unstable steady states; for a clear exposition of this behavior, see Fogler (1986). The reader might be thinking, "This doesn't happen frequently; so, is it important to evaluate the design for stability?" The answer comes from industrial experience related by Bush (1969). The ICI Chemical Company was developing a process for chlorinated hydrocarbons using a pilot plant reactor; unfortunately, it operated in continuous cycling operation, so it never achieved steady state (as desired). A thorough dynamic analysis confirmed that the reactor had a wide range of conditions where the operation was unstable without control. We can learn two lessons from this industrial experience. One, unstable process operation, while not common, does occur. Second, the engineer is wise to evaluate a potentially unstable system before completing the design.

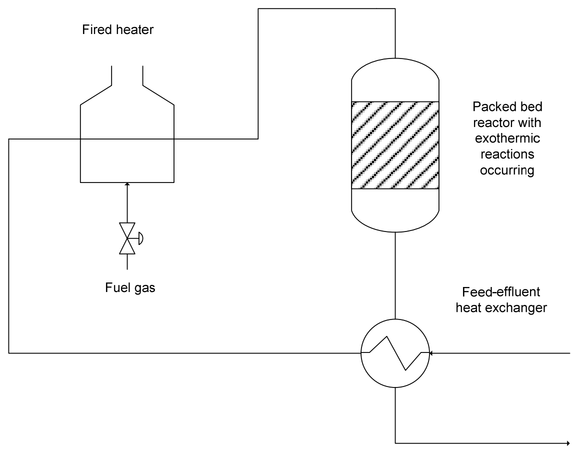


Figure 6.31a Reactor with feed-effluent heat exchanger. If each process is stable, the system can be unstable.

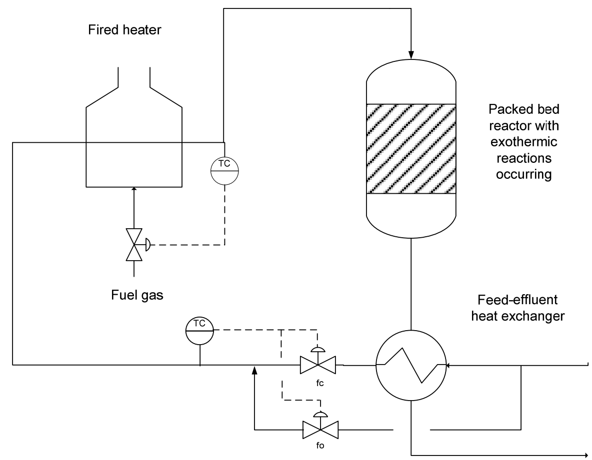


Figure 6.31b Reactor with feed-effluent heat exchanger and one possible stabilizing feedback control design.

Even when each individual process is stable, the process integration of material and energy can cause the entire unit to be unstable. A typical design structure involves feed-effluent heat exchange that increases the energy efficiency of a design by exchanging heat to reduce utility costs for heating (or cooling). An application of this design structure is shown in Figure 6.31a for a packed bed reactor that has exothermic chemical reactions. Note that an increase in the reactor outlet temperature causes an increase in the reactor inlet temperature. This disturbance can continue "around the process loop", increasing in magnitude each time. The result is instability. For this design, when the individual processes (heat exchanger, fired heater, and reactor) are stable, the positive feedback due to the feed-effluent exchanger can result in the process system being unstable!

Such a system is analyzed by Silverstein and Shinnar (1982). Process controls shown in Figure 6.31b stabilize the system by tightly controlling the reactor inlet temperature, preventing a reactor outlet deviation from affecting the reactor inlet. Silverstein and Shinnar also provide guidance on the information and models useful in analyzing reactor dynamics at the design stage.

Depending on the strength of the positive feedback in the process system, other design modifications might be required to moderate the effects of disturbances. Some other possibilities are summarized in the following.

- Control the reactor outlet temperature by adjusting the inlet temperature via a cascade
- Reduce the amount of heat exchanged in the feed-effluent heat exchanger and increase the duty of the fired heater
- Split the reactor into multiple beds with inter-bed heat exchangers for cooling
- Split the reactor into multiple beds and inject cold feed into the inlet of each bed to control temperature. The cold feed would by-pass all preheat equipment.
- Add inert packing in the reactor to reduce the rate of change of bed temperature

Note that the use of multiple reactor beds also corrects uneven flow distribution, reducing the chance for catalyst bypassing, low fluid flow near dense packing, and local "hot spots".

Clearly, process structures with heat integration improve efficiency and introduce challenges to the dynamic operation. The creative design engineer can realize the energy savings without sacrificing safety or stable operation with consistent product quality.

Good control performance requires the control of unstable variables, with the exception of inventories that have a very large V/F (volume/flow) ratio enabling effective manual operation.

6.3.2 Process elements in (only) the disturbance path

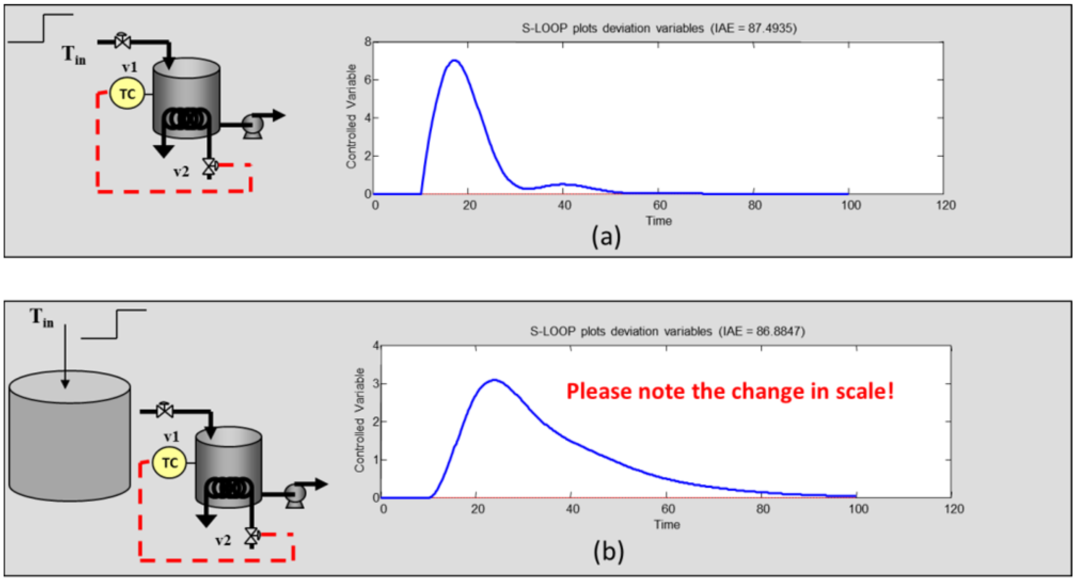
It makes sense that the disturbance characteristics will influence the variability of the controlled variable. Here, we will consider key factors and their influences. Before we begin, we should emphasize an overriding perspective in control.

The first approach should be to eliminate or reduce disturbances. The design steps discussed here are performed after reasonable effort has been expended to prevent disturbances.

6.3.2a Disturbance time constant(s) – Disturbance time constants are between the disturbance origin (flow rate, composition, temperature, and so forth) and the control loop. These time constants will "slow" the disturbance effect, which will be beneficial by reducing the deviations experienced by the controlled variable.

Example 6.8 Disturbance time constant - Let's look at an example of the effect of a disturbance time constant on control performance. In Figure 6.32a, the step disturbance affects the feedback loop directly, and the maximum deviation of the controlled variable from its set point is greater than 7. For the same feedback process, disturbance and controller in Figure 16.32b, the step disturbance passes through a tank (first order system) before affecting the feedback loop, and the maximum deviation is much lower, about 3.2. Clearly, the performance is better with the extra tank.

Control performance is improved by large disturbance time constants (that do not appear in the feedback loop). The improvement is in reducing the maximum deviation from set point, or equivalently smaller variance.



Case (a) $\quad G_P(s) = \dfrac{1.0e^{-5s}}{1+5s} \quad G_{Tank}(s) = 1.0 \quad K_C = 0.8,\ T_I = 7.0$

Case (b) $\quad G_P(s) = \dfrac{1.0e^{-5s}}{1+5s} \quad G_{Tank}(s) = \dfrac{1.0}{1+15s} \quad K_C = 0.8,\ T_I = 7.0$

Figure 6.32 Feedback control performance for (a) a disturbance directly entering the feedback loop and (b) a disturbance through a time constant before entering the feedback loop.

6.3.2b Disturbance frequency – A periodic disturbance can usually be characterized by a frequency. The question is, "Does this frequency affect control performance?" The answer is a resounding "Yes!" As explained in Appendix 6.A, disturbances near the critical frequency are not controlled well by a feedback controller. This is because the controller is not fast enough to compensate for the positive half wave before the negative half wave occurs. The control system is "chasing its tail", and the performance can be worse than if the controller were switched to manual.

On the other hand, disturbances with much higher frequencies contribute little variation because the process (time constants) attenuates the disturbances. In addition, disturbances with a lower frequency are easily attenuated by the feedback controller. Therefore, disturbances near the critical frequency lead to poor control performance. This situation is shown in Figure 6.33, and further discussion of this topic is given in Appendix 6.A.

Process modifications are needed for disturbance frequencies near the critical frequency. One modification would be to add a mixing tank between the disturbance source and the feedback loop. Another modification introduces a fast control loop to compensate for the disturbance before affecting the loop; an example would be a heat exchanger to regulate the temperature of an input stream that affects the feedback loop.

We might wonder if disturbances near the critical frequency of typical control loops occur at all. Sadly, they do, and the cause is often other control systems that are poorly implemented. Many process units have similar dynamics, and when integrated units have poorly tuned control loops, the oscillations from one process are near the critical frequency in another unit. Therefore, one poorly tuned loop can cause difficulties in many other interacting control loops. This is the reason that some PID tuning rules that yield oscillatory responses, such as ¼-decay ratio, are not used in practice.

Control performance is good for disturbances with very low (easily controlled) or very high (attenuated by process time constants) disturbances. The design engineer should consider process or control modifications for disturbances near the feedback loop critical frequency.

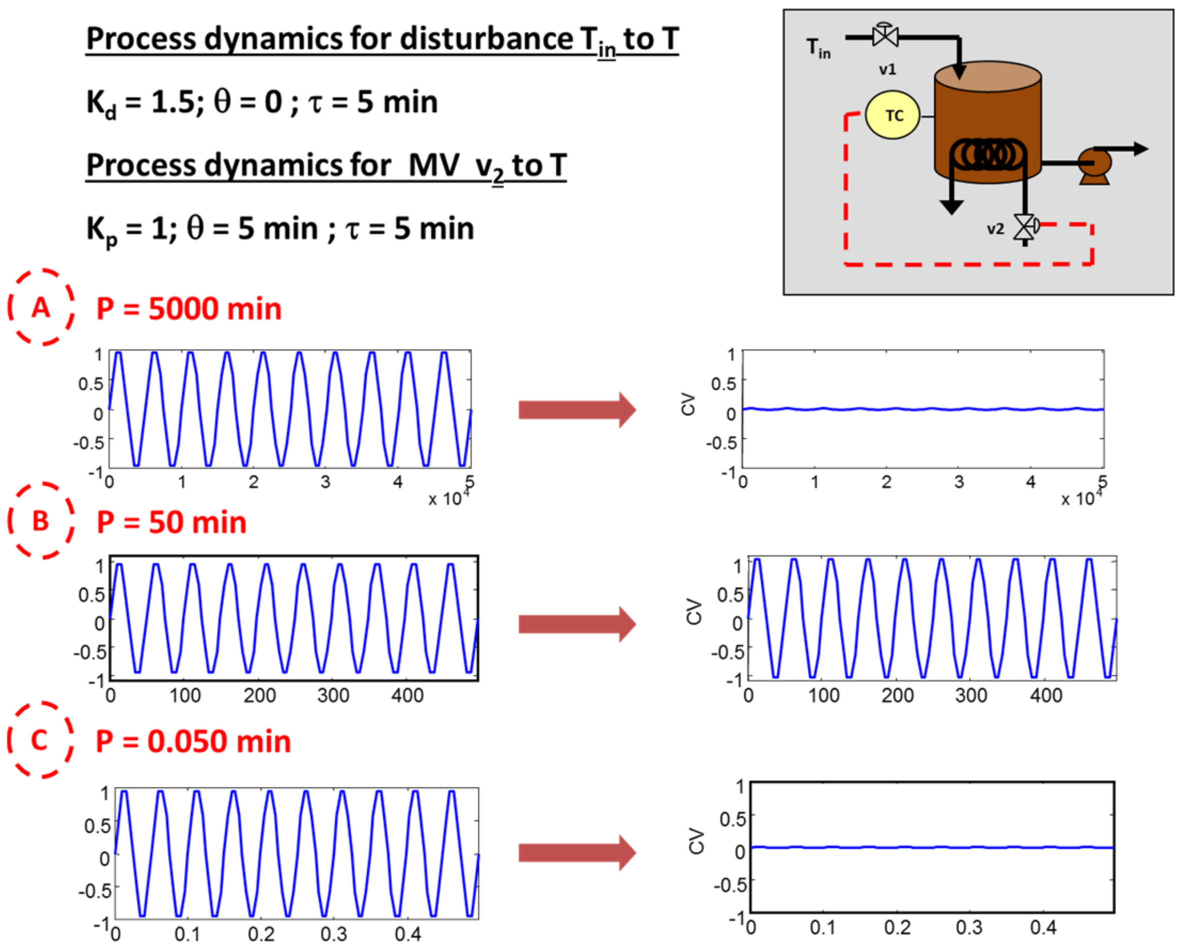


Figure 6.33 The effect of a sine disturbance on the process with control at three frequencies.
A is very low frequency where feedback is effective
B is near the critical frequency where control is ineffective
C is high frequency where the process attenuates disturbances

6.3.2c Disturbance magnitude – Naturally, large magnitude disturbances are worse than smaller magnitude disturbances. For a truly linear system, the controlled variable deviation from set point is proportional to the disturbance magnitude. While the strict proportionality is not valid for non-linear systems, the general conclusion remains valid. The process control system (controller and process equipment) might not be able to compensate for very large disturbances because the manipulated variable could be adjusted to its upper or lower limit, where no compensation will be possible. Engineers must account for very large disturbances by including additional systems with greater capacity for compensation. These systems are addressed in Chapter 5 on Safety and here in Section 6.2.4.

Preventing disturbances of large magnitude requires actions involving process elements outside of the loop being considered. For example, fuel gas systems often mix purchased natural gas with plant by-products. Occasionally, a unit upset will result in a large flow of by-products into the fuel gas, introducing major disturbances in the fuel gas pressure and molecular weight (and heat of combustion). Control systems can be implemented that compensate for these disturbances and maintain the desired heat release in burners in boilers and fired heaters. From this brief example, we note that the engineer must know typical disturbance sources to properly design control systems!

Control performance is good for small disturbance magnitudes.

6.3.3 Instrumentation elements in the feedback loop

Sensors, final elements, signal transmission and the controller calculation all play vital roles in successful process control. In the introductory process control course, we often assume that the instrumentation functions perfectly, so that we can concentrate on the process elements. Excellent instrumentation performance can be approached when engineers select, install, and maintain instrumentation that is well matched to the plant requirements. Here, we will introduce some common instrumentation factors that affect control performance.

6.3.3a Sensor and final element dynamics – The sensor and final element are in the feedback control loop, so that any delays introduced by the instrumentation will degrade control performance. Since process equipment has rather slow dynamics, the instrumentation can often be assumed instantaneous. Typically, signals from F, T, P and L sensors respond within a few seconds of a change in the process variable, and control valve stem positions respond in a few seconds from a change in the signal from the control room. This is much faster than the dynamics of heat exchangers, distillation towers and most industrial chemical reactors. (In contrast to process equipment, think about controlling a disk drive or a machine tool cutting metal.)

Sensors can introduce significant delay when performing a complex chemical analysis on material sampled from the process. The sampling and analysis can be entirely automated and performed periodically without manual intervention. Two aspects of this procedure can slow feedback control. The first is the sampling, involving extracting the material and transporting it a short distance (tens of meters) to the analyzer. Since the analyzer needs only a small amount of material, the flow from the process to the analyzer could be small, which would result in a very long transportation dead time, and we know that dead times are bad! An easy solution involves providing a "fast loop" with a high flow rate that brings the material close to the analyzer, which can then extract a small amount. The fast loop returns the material to the process, so that no product is lost. A schematic of a fast loop is given in Figure 6.34. The second aspect of analyzer for control is the execution period that is addressed in a subsequent section.

Typical final element response can be too slow for some machinery control applications. An example is recycle control around a centrifugal compressor. The recycle is needed to ensure a minimum flow rate through the compressor when the feed rate is too low, because a low flow through the compressor causes surge and damages the compressor. Surge occurs very quickly, so that the recycle response must be rapid (Staroselsky and Laudin, 1979; Smith and Kurz, 2005, Engencyclopedia, 2012). An example is shown in Figure 6.35, with the set point for the FC defining the minimum flow rate. The control valves for this application must be selected to have fast response, which can involve volume boosters to increase the air supply flow rate to the pneumatic actuators.

Control performance is good when the instrumentation dynamics are negligible compared with the process dynamics in the loop.

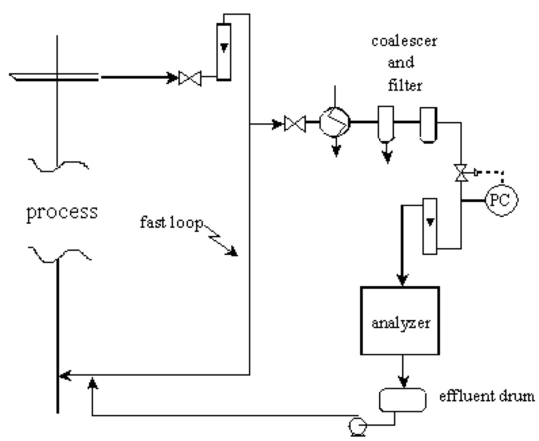


Figure 6.34 A typical analyzer sample system with a fast loop to reduce dead time.

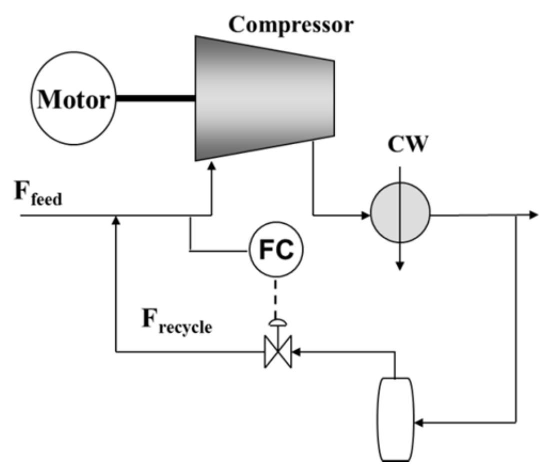


Figure 6.35 Compressor anti-surge recycle control where fast instrumentation is essential.

6.3.3b Measurement noise – We use the term "noise" for contributions to a measurement or signal that is not repeatable and generally does not represent the behavior of the variable being measured. In this situation, electrical interference and mechanical vibrations that artificially modify a signal certainly qualify as noise. Generally, we also categorize high frequency process variation, like liquid level oscillations, as noise; they represent the real process but are high frequency and should not be considered for control. The controller acts on the measurement, and any part of the measurement that does not represent true process changes will lead to incorrect controller actions. Therefore, we would like to separate the "noise" from the true "signal". The effects of electrical noise can be reduced by proper cabling. We use filters to remove additional high frequency components from the signal, as shown in Figure 6.36a.

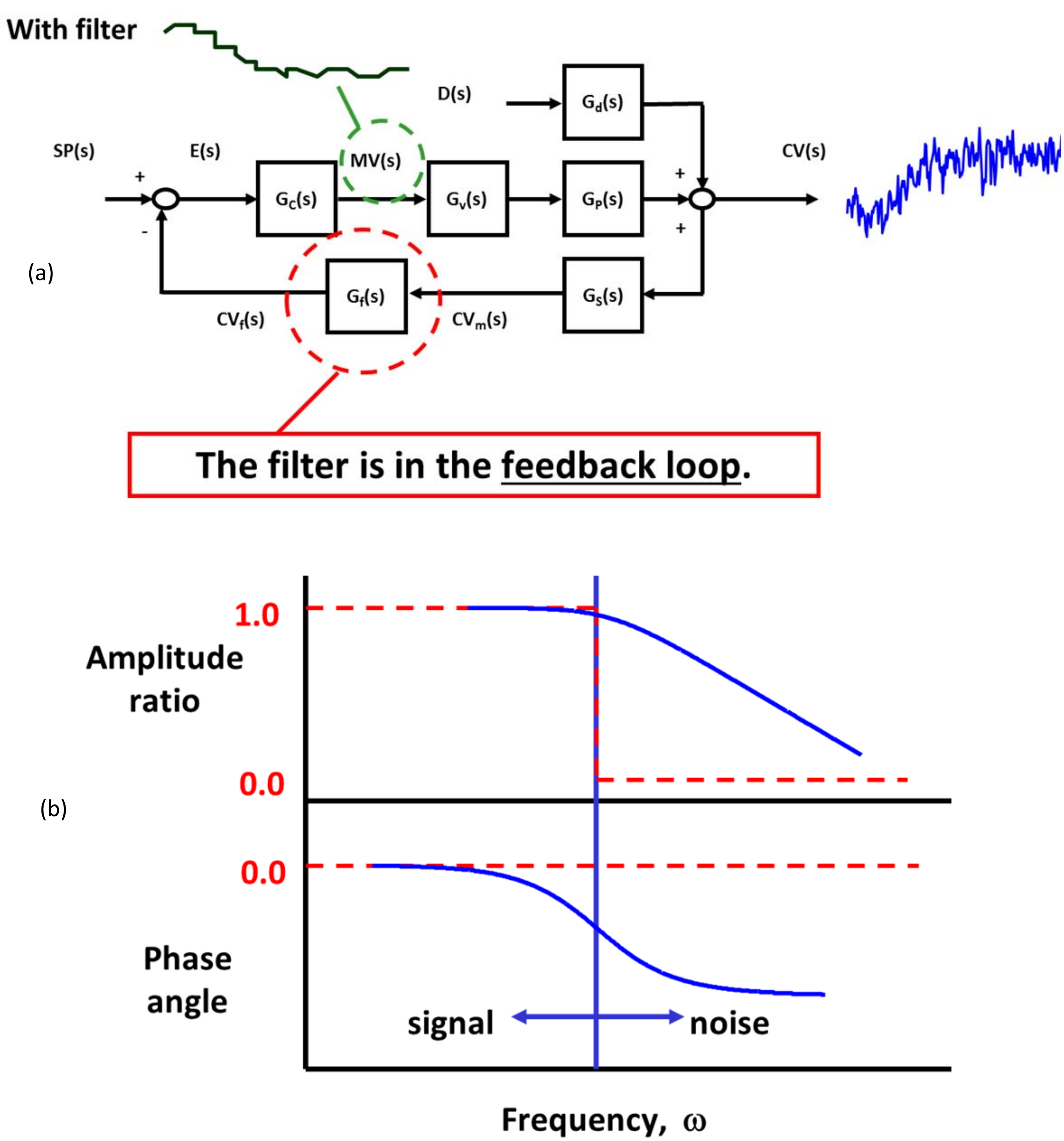


Figure 6.36. Filtering measurement noise.
a. Block diagram showing the filter location and decrease in manipulated variable variation
b. Bode plot for $CV_f(s)/CV_m(s)$; the distinct signal-noise boundary is not actually known (red dashed line is "ideal" filter; blue solid line is first order filter)

We often think about the frequency of control signals according to bands. Frequencies much higher than the process dynamics are considered noise. However, overlap usually exists so that there is no clear boundary separating noise from signal. The "ideal" filter performance is shown in Figure 6.36b as a red, dashed line; it does not affect the "true signal", reduces the noise amplitude to zero, and introduces no delay. This ideal filter is <u>not possible</u>. The most common filter for process control signals is a first order filter, which is similar to a mixing tank. The performance of a first order filter is compared with the ideal filter in Figure 6.36b. The filter time constant is chosen to reduce the part of the signal thought to be noise.

A word of caution is appropriate here. The filter appears in the feedback loop, so it delays control. A common error is to "over-filter" a signal, i.e., apply a filter with too large a time constant. With (too) large a filter time constant, trend plots of the filtered signal look smooth, with little variation. However, the real process variable could have more variation because a large filter slows feedback, requiring looser PID tuning.

Control performance is improved when measurement noise is small. Engineers can use filtering, as long as it does not significantly influence the feedback loop dynamics.

6.3.3c Non-ideal final element behavior – The most common final element in chemical processes is a pneumatically actuated control valve. These valves are well manufactured, but they are not precision instruments. The typical valve has some difference between static and dynamic friction, so that a minimum force is required to start motion, and it has some "gap" when changing direction (e.g., opening to closing) (Choudhurya et. al., 2005; Wallen, 1997). The result is non-ideal behavior. As shown in Figure 6.37, oscillations and poor performance occur in the control loop when the final element non-ideality is large. An important step in the design stage to prevent this behavior is to ensure that control valves are sized properly. The guideline "always oversize equipment to prevent mistakes" will lead to a valve that is operating in a narrow range, e.g. 10-30%, which will exacerbate the effects of valve friction.

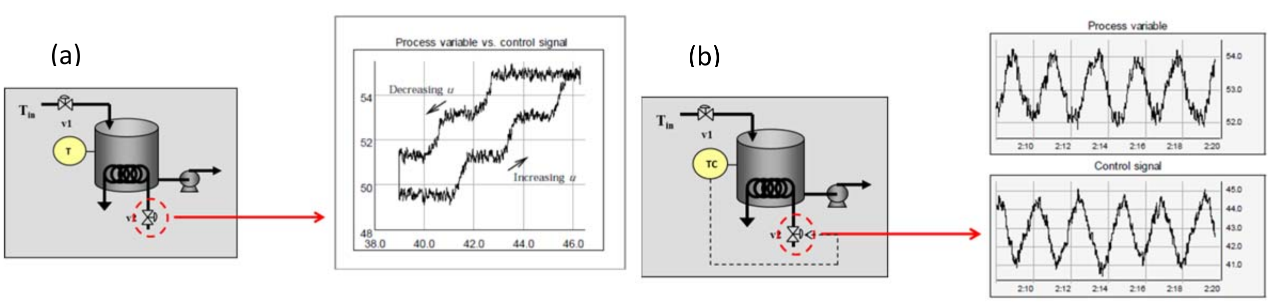


Figure 6.37 Effects of non-ideal valve performance (Wallen, 1997)
a. Behavior during an experiment without control (u = control signal)
b. Behavior of closed-loop feedback control with poorly performing valve

Several solutions to inadequate valve performance are given in the following.

- A valve with smaller maximum flow rate can be used. This will only be appropriate if the initial valve were oversized, requiring adjustment in much less than a normal range of 50-100% of opening.
- A valve positioner can be added to the valve. A valve positioner is essentially a fast secondary in a cascade structure. It measures the actual valve stem position and controls it to be close to the command signal from the controller. A schematic of a valve positioner is given in Figure 6.38a.
- The valve can be repaired or replaced.
- If small adjustments are required for a large base case flow rate, two parallel valves can be included in the design. Normal, small adjustments from the controller are made to the smaller valve. The larger valve is only adjusted infrequently to ensure that the smaller valve is able to affect the flow, i.e., that it is neither fully opened nor closed. A schematic of this control is shown in Figure 6.38b

Good control performance relies on final elements responding precisely to controller commands. Valve performance can be monitored so that poorly performing valves can be repaired.

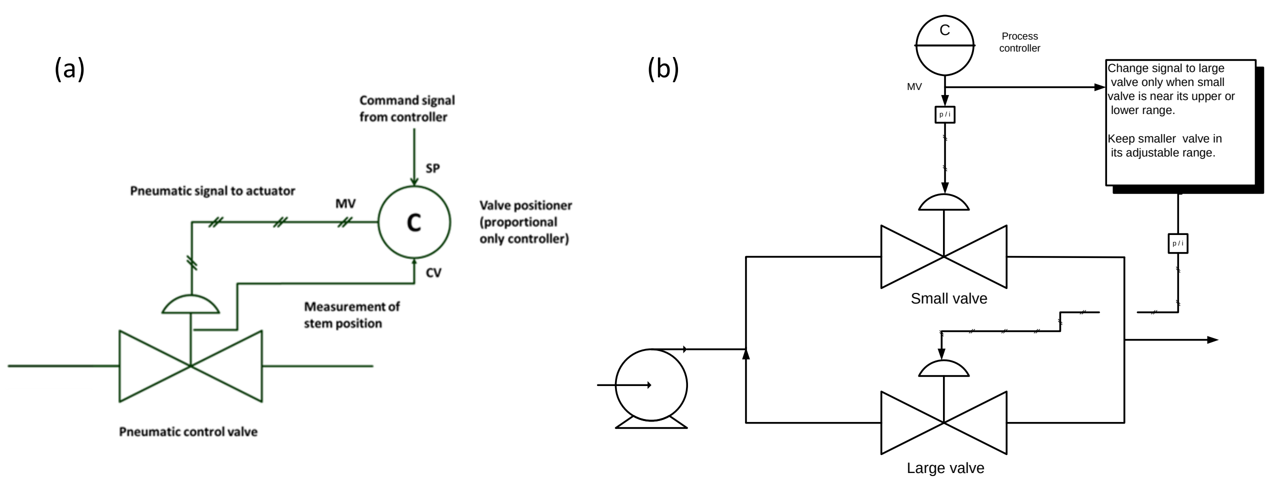


Figure 6.38. Improving the performance of control valves
a. Schematic of valve positioner.
b. Strategy using small valve for high precision with small adjustments.

6.3.3d Controller execution period – Digital controller computations are performed at a fixed period. Naturally, we would like this period to be short compared with the feedback process dynamics. Usually, this is not difficult to achieve with modern digital computation. The most frequent factor that lengthens controller execution is on-stream analyzer time, where a substantial amount of time is required for analysis. In general, control performance is not affected significantly if the controller execution time is a few percent of the sum of process dead times and time constants in the feedback loop; a goal is to keep this percentage below five percent. A comparison of continuous and digital control with an excessively long execution period is given in Figure 6.39; note that the IAE has nearly doubled for less frequent feedback control.

When a long execution period cannot be avoided, the design engineer should seek an inferential variable for the analyzer measurement. The fast inferential variable can be used to compensate for most disturbances, while the much slower analyzer feedback can reset the inferential set point in a cascade structure.

Good feedback control performance requires rapid digital execution when compared with the process dynamics.

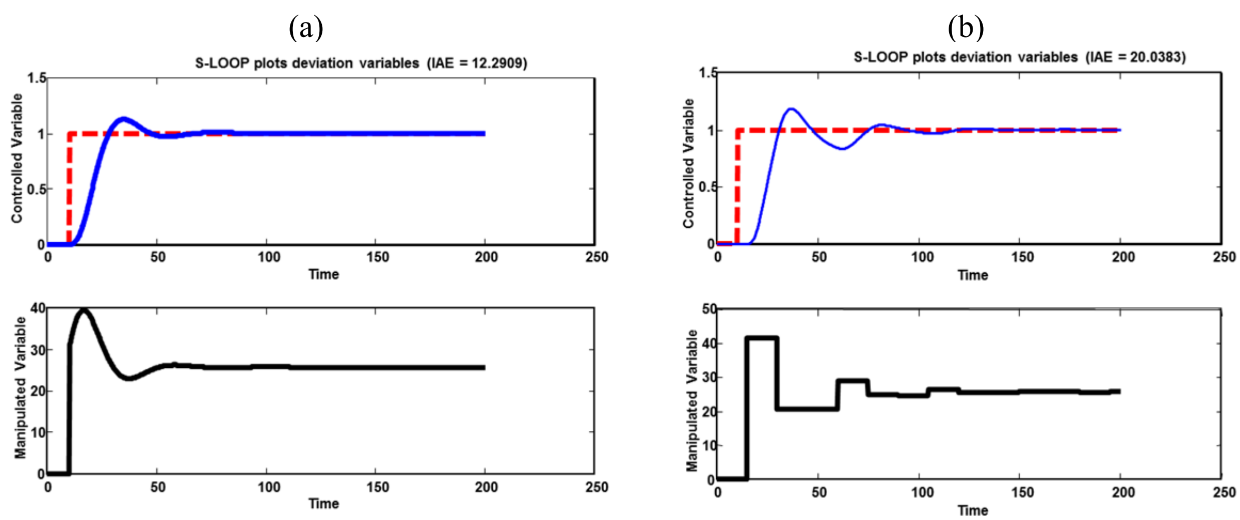


Figure 6.39 Feedback control performance (a) continuous controller and (b) digital controller with execution period equal to the sum of the time constants in the feedback path.

6.3.4 Process Structure

Engineers like to concentrate on individual processes and design them to function well. This is a good start, but we must remember that many units share material and energy, so that there is a recycle or feedback *due to the process structure*. These recycles are included in the plant to improve the profitability by recovering energy and materials that would otherwise be lost for productive purposes. It is the job of the design engineer to ensure that efficient performance can be maintained during dynamic operation.

6.3.4a Feedback from integrated processes – The streams exiting a process proceed to other units in the plant. In some designs, these streams return to exchange energy of material with the original process. This feedback is nearly always positive feedback, i.e., the effect is to drive the process away from the desired operation. Special process and control structures are required to moderate the unfavorable dynamic effects of integration. Since this topic is addressed in detail in Section 6.5.4, no further discussion is given to this important topic here.

Generally, recycle systems are more difficult to control, with higher sensitivity to disturbances and slower dynamics.

6.3.4b Interaction among control loops – Most individual processes require many single-loop controllers to achieve desired dynamic performance. Because each valve adjustment affects more than one controlled variable, interaction among loops occurs. This interaction has a profound effect on control design and achievable performance. Since this topic requires considerable development, interaction is covered fully in Section 6.4.

Feedback control design is simpler when little interaction occurs between loops. When strong interaction exists, determining the best loop pairing requires analysis by the engineer, which is addressed later in this chapter.

6.3.5 Control performance goals

Naturally, the performance goals have a substantial impact on the challenges to achieve the goals. Very restrictive goals are more challenging and usually require more complex and expensive designs.

6.3.5a Product quality specifications – Material quality specifications can allow considerable variation or they can be very restrictive. Naturally, the more restrictive the quality specifications, the more challenging it becomes to achieve the control objectives. Restrictive specifications follow from customer requirements. For example,

- Monomer purities must be very high to produce polymers with consistent molecular weight distributions
- Food produces must prevent foreign materials
- Pharmaceuticals must adhere to specifications agreed by licensing agencies, such as the FDA

A key aspect of the quality specification is whether it applies to the time-average property or to each individual sample of the product. Let's consider the gasoline blending process shown in Figure 6.40. The product must satisfy many specifications, like octane, reed vapor pressure, volatility, and many compositions, e.g., percent alcohol. These specifications are imposed on material at the end of the batch in the product tank, so that the instantaneous properties at the blending point need not satisfy the specification. This situation makes the control easier because a short-term deviation (such as a low octane) can be compensated by a subsequent deviation (of higher octane) during the batch blend. The key process factor here is the mixing of the product before specifications are imposed.

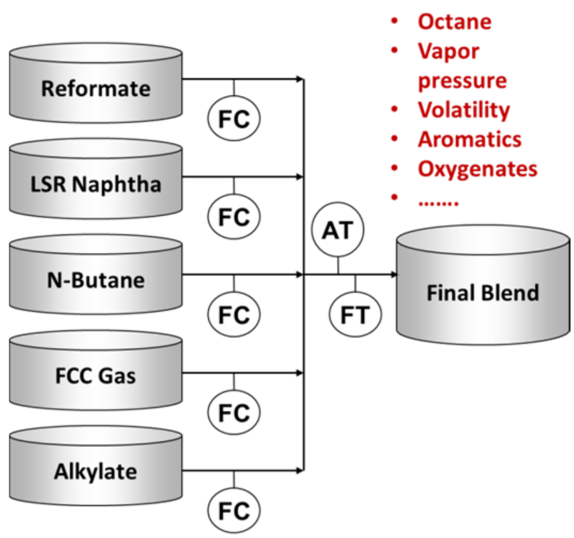


Figure 6.40 Quality control where material is mixed before qualities are determined.

In contrast, a pharmaceutical manufacturer making pills must meet specifications for every pill. What if the process produced four pills containing no active ingredient and the fifth containing five times the desired ingredient. It would not be acceptable, and it might be deadly! A similar situation occurs for a manufacturer of sheet materials, such as paper, steel and plastic, as shown in Figure 6.41 for sheet steel. The entire sheet must be uniform; the quality would be unacceptable if part of the sheet were half as thick as desired, while another part of the sheet were 50% too thick. The key process factor here is the lack of mixing or other compensation for off-specification material.

In general, material mixing before imposing quality specifications eases achieving these goals. Instantaneous and spatial specifications are more challenging to achieve.

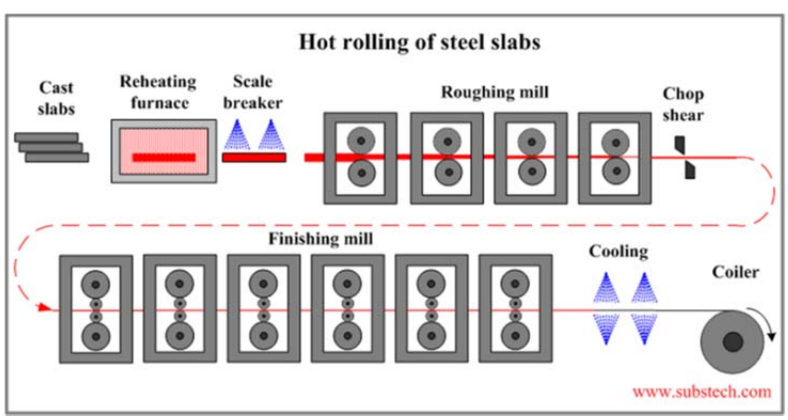


Figure 6.41 Quality control where every location must be manufactured to specification. (Substech, 2012)

6.3.5b Penalty for constraint violations – Essentially, every process involves many limitations to operating conditions. The boundaries of the operating window are defined by limits on

- Stream properties, for example, product qualities.
- Operating conditions affecting equipment performance, for example, liquid and vapor flows on distillation trays.
- Capacities of equipment, for example, maximum flow rate for a pipe, pump and control valve combination.
- Safe operation, for example, the rate of reactant additional in a batch reactor.
- Conditions leading to equipment damage, for example, maximum pipe temperatures in a fired heater.

The penalty for constraint violation can be minimal or it can be very high. A high penalty could be due to very undesirable consequences for the violation such as (i) loss of production due to spoiling a batch of product, (ii) damaging equipment, or (iii) activating a Safety Instrumented System (SIS) that shuts down the process.

Example 6.9 Approach to constraint - When the penalties are high, plant personnel tend to operate equipment far from the limiting conditions. In many situations, operating near such constraints generates high economic returns. As a result, a conflict exists concerning how much short-term risk should be accepted to realize high profit (if a violation does not occur) versus the penalty for the violation. Operating near limitations requires excellent feedback control performance, i.e., low variance, that is usually too demanding for personnel to achieve manually.

Let’s consider an example of a pyrolysis fired heater that has a chemical reaction occurring in the piping; a process schematic is given in Figure 6.42. There can be a high incentive to operate at a high temperature that gives a high conversion, but a maximum temperature must not be violated to protect the equipment. This situation is shown in Figure 6.43. Therefore, we anticipate that a large economic benefit exists to operate at a high temperature without violating the temperature constraint.

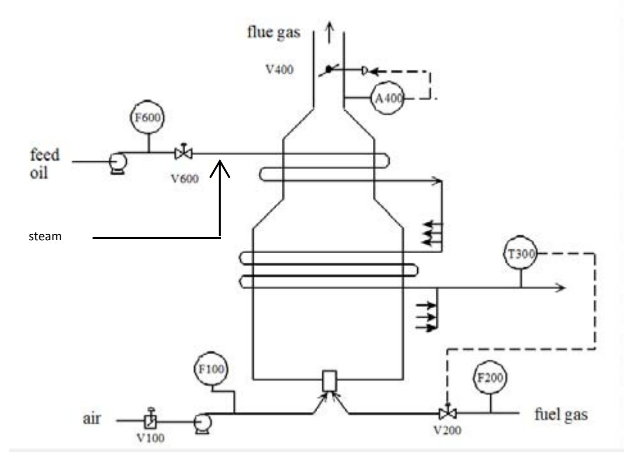


Figure 6.42. Pyrolysis furnace process schematic.

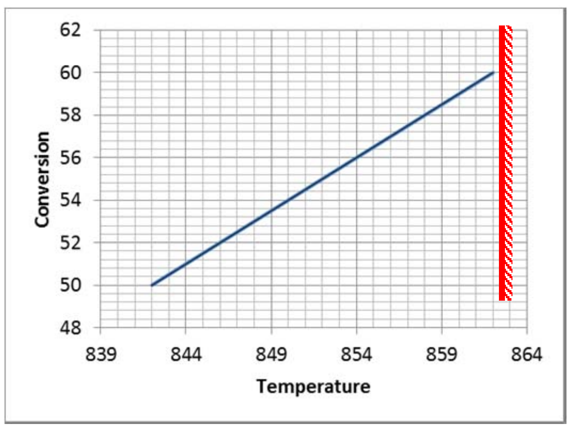


Figure 6.43. Pyrolysis reactor, relationship between outlet temperature and conversion

We will consider three different plant performances achieved by feedback control; the distributions of temperature data for all cases are shown in Figure 6.44 in histograms. In the base case (a), the temperature control is not good; the temperature experiences a large variation, and the set point must be located far from the limitation to ensure that (essentially) all operation will be below the maximum temperature. In Case (b), the control has been improved, and the variation is much lower. However, the set point has not been modified, so that no advantage has been realized from the improved control performance. In Case (c), the temperature control performance is improved and the set point has been increased to take advantage of the reduced variation. Note that Case (c) achieves higher average conversion without increasing the risk of violating the maximum limitation. Further discussion on process control benefits and the use of historical data in histograms is given in Appendix 6.B, and more detailed case studies are available in Marlin et.al. (1987).

Where process limitations with high penalties exist, the tendency is to operate far from the limitations. Often, good control performance can reduce the variability and enable more profitable operation nearer to the limitation.

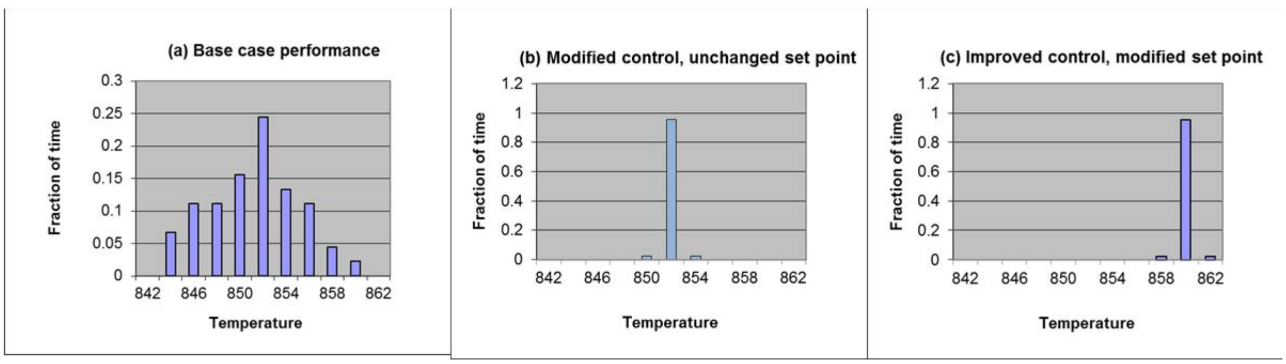


Figure 6.44 Histogram of reactor temperature variation.
Case a. Base case with poor initial control
Case b. Improved control with same set point as the base case, Case (a)
Case c. Improved control with modified set point closer to the temperature limit

6.3.5c Production rate specifications – We must ultimately manufacture the amount that we need to satisfy sales, or if the process is meeting another need, such as a waste treatment plant, the process must handle all effluent it receives. Production control involves issues such as the following.

- Periodic delivery of raw materials and dispatch of products
- Need to immediately process inputs or provide products on demand
- Storage of raw materials, intermediate products (work in progress) and final products
- Balancing the production in integrated plant sections
- Observing product rate limitations throughout the plant (bottlenecks)
- Scheduling maintenance while providing products when required

Defining the product rate control structure is one of the first decisions is plant-wide control. This topic is addressed in Sections 6.4 and 6.5; therefore, it will not be discussion further here.

Production rate specification, specifically how rapidly the process must respond to demand changes, has a strong impact on control design. This topic is addressed later in the chapter.

6.3.5d Profit sensitivity – The feasible operating window might be very large for a process, allowing a range of operating conditions to safely manufacture products satisfying quantity and quality specifications. Usually, the profitability of the plant operation varies within the operating window, so that the control system should maintain operation near the maximum profitability within the window. If the location of the highest profit does not change (or changes infrequently), the engineer can perform a study once to determine the best conditions and define these in the operating procedures. In many plants the highest-profit operating conditions change with disturbances like raw material composition, production rate, product grade, equipment performance (e.g., reactor coking), equipment maintenance schedule (e.g., reactor decoking), and environmental conditions (e.g., refrigeration system capacity).

There are many approaches to tracking a changing optimal operating condition. These are presented briefly in the following.

- **Conditions enforced by process control** – In some instances, a study determines that the highest profit can be at least partially achieved using relatively simple strategies. In the pyrolysis fired heater reactor in Figure 6.42 discussed above, studies show that the steam flow should be maintained in a ratio to the hydrocarbon feed rate, which can be achieved with a simple flow ratio control design.
- **Standard operating procedures** – In other situations, the best operating conditions depend upon a few key variables, such as the type of raw materials. The engineer can perform studies to determine the best conditions and define rules for plant operations. Considering again the pyrolysis reactor, the proper steam to hydrocarbon feed ratio depends on the feed type (ethane, propane, naphtha, gas oil, etc.).
- **Direct search** – In direct search methods, the operating personnel introduce (small) changes to key operating variables and determine the profit at each experimental point from plant measurements. The experiments indicate a direction in which the profit increases. The process conditions are changed to a new point with higher profit. Then, the procedure is repeated until experiments do not yield discernible profit improvement. These methods are also termed "response surface" approaches.
- **Real-time optimization** – In challenging situations, the profit changes frequently and by a large amount and the best conditions cannot be determined by rules or simple strategies. In these situations, an advanced control strategy can characterize the plant by a model that is corrected or "updated" using current plant data, and the model can be optimized to determine the optimal operating conditions. Naturally, this is the most complex approach, and it is justified where the economic benefits are substantial (Marlin and Hrymak, 1996; Pedersen, 1995; Vermeer, 1996).

Process automation at a higher level in the control hierarchy can address profitability directly. These profit controllers adjust the process slowly to maintain safety, product quality, and production rates.

6.3.5e Safety- We conclude this discussion of factors affecting control design with the most important factor, safety. The Basic Process Control System (BPCS) provides the first of many barriers between a potential disturbance and an accident. The engineer must consider potential disturbances and provide BPCS designs to compensate for the disturbances and maintain conditions within safe limits. Naturally, the control systems cannot compensate for the largest disturbances, which is the reason for many additional barriers. However, the control system should keep the process in acceptable conditions nearly all of the time, so that activation of higher barriers (alarms, Safety Instrumented Systems (SIS), pressure relief valves, and containment) should be very infrequent. Process safety and the safety hierarchy are the topic of Chapter 5, where much more on safety is presented in an compelling and informative manner.

Good control performance requires tight control of all safety-related process variables by the Basic Process Control System (BPCS). The process design should provide a manipulated variable with a fast and strong effect on each safety-related variable.

6.4 Multivariable Process Systems

Most processes require the control of several variables for successful dynamic operation. The engineer needs to understand some basic concepts in the behavior of multivariable systems to design their controls. In this section, these concepts are introduced and are applied in example multiloop control designs.

6.4.1 Which variables can be controlled?

We must ensure that the process design allows all key variables to be controlled. Let's start with a single controlled variable and determine what is required for feedback control of the variable. As we saw in Section 6.1, a feedback loop requires a sensor, a control calculation, and a final element, usually a valve. Naturally, the sensor is selected to measure a key variable requiring control. What is a fundamental requirement for the valve? The valve must influence the control variable, i.e., there must be a causal relationship between the valve and the sensor. Therefore,

For a single-variable process, feedback control is possible when $K_P \neq 0$.

The existence of a causal relationship (or its absence) is relatively easy to determine for a single controlled variable using fundamental principles and/or empirical data, but the analysis becomes considerably more complex for multiple controlled variables. Next, we consider how many variables can be controlled in a process. Since we require an independent valve adjustment for each controlled variable, we conclude the following.

Degrees of Freedom: In a multivariable system, the maximum number of controlled variables is equal to the number of adjustable final elements, i.e., valves.

We must recognize that this statement does not guarantee that every valve can be used to independently control a *meaningful* process variable. Without evaluating the process and valve location, the only thing that can be stated with certainty is that all remote valve positions can be adjusted. Therefore, we proceed to the critical issue of controllability.

In automatic control literature, the term controllability has many definitions, each of which can be applied to a specific set of relevant control issues. We will use a limited definition here that will enable us to analyze most continuous process control designs.

Controllability: A multivariable process is controllable if the controlled variables can be maintained at their set points by adjusting the selected manipulated variables as disturbances enter the system.

Example 6.10. Process Controllability: Chemical plants have boilers to generate steam for power and heat transfer. The simplified process in Figure 6.45 shows a boiler vessel where boiling water and steam separate. We would like to control the temperature and the pressure in the vessel, the design shows two valves adjusting the fuel combusted and the steam leaving the vessel. Is this an appropriate design that enables the two variables (T and P) to be controlled?

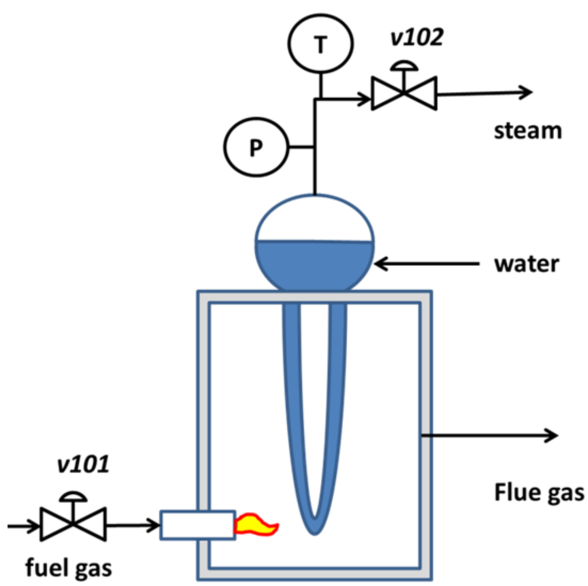


Figure 6.45 Schematic of fired boiler

Let's first determine whether each valve influences at least one controlled variable, i.e., do individual input-output causal relationships exist?

Opening valve	***steam temperature***	***vessel pressure***
v101	***increase***	***increase***
v102	***decrease***	***decrease***

Opening the fuel valve will result is a hotter flame and greater heat transfer to the vessel contents, which will generate more steam, increase the pressure and increase the temperature. Opening the steam exit valve will decrease the pressure in the vessel and decrease the temperature.

We might conclude that sufficient causal relationships exist for controlling the temperature with the fuel valve (v101) and the pressure with the steam valve (v102). However, this would not be correct!

We have to use our chemical engineering knowledge of the process. The water in the vessel is boiling. We know that for a single component at its boiling point, the temperature and pressure are related. (Consult steam tables for saturated steam.) Therefore, the temperature and pressure cannot be controlled to independent values. Since pressure is important for safety, we would control pressure by adjusting the fuel flow. In most designs, the steam control valve (v102) would be removed and saturated steam temperature would not (cannot) be controlled. To control the steam temperature, an additional heat exchanger, termed a steam superheater, must be added to the process design (Ganapathy, 2001; ISA, 2007).

We conclude that the system is not controllable as originally designed.

From Example 6.10, we conclude that having the proper number of *individual* causal relationships is not sufficient for controllability of a multivariable system. We must determine whether the multiple input-multiple output system has the sufficient number of *independent* relationships. We can evaluate the controllability (as defined above) by evaluating the steady-state relationship between manipulated and controlled variables. The system is controllable if a solution exists for this set of linear equations for arbitrary values for the disturbances. (A two-variable system is shown and extension to "n" variables is straightforward).

$$\begin{bmatrix} CV_1 \\ CV_2 \end{bmatrix} = \begin{bmatrix} 0 \\ 0 \end{bmatrix} = \begin{bmatrix} K_{11} & K_{12} \\ K_{21} & K_{22} \end{bmatrix} \begin{bmatrix} MV_1 \\ MV_2 \end{bmatrix} + \begin{bmatrix} K_{d1} \\ K_{d2} \end{bmatrix} D \qquad 6.6$$

where the K_{ij}'s are steady-state gains.

A solution exists for a square system of linear equations when the inverse exists for the steady-state gain matrix. Therefore, a process system modeled by the equation is controllable (in the steady state) if the determinant of the gain matrix is non-zero.

Example 6.11 Controllability: The two streams in Figure A6.46 are being blended to make a desired quantity with a desired concentration of component "A". One stream is pure A, while the other has no A, and there is no volume change on mixing. Can the desired total flow and concentration be achieved by adjusting both component flow rates?

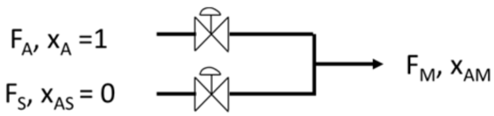


Figure 6.46 Blending process

We begin by deriving the following total and component material balances on the mixing point.

$$F_A + F_S = F_M$$
$$F_A \underbrace{x_A}_{1} + F_S \underbrace{x_{AS}}_{0} = F_M x_{AM}$$

The linearized model is directly determined to be the following.

$$F'_M = F'_A + F'_S$$
$$x'_{AM} = \left[\frac{F_S}{(F_s + F_A)^2}\right]_{ss} F'_A + \left[\frac{-F_A}{(F_s + F_A)^2}\right]_{ss} F'_S$$

The determinant of the gain matrix is given in the following.

$$Det(K) = \frac{-F_A}{(F_A + F_S)^2} - \frac{F_S}{(F_A + F_S)^2} = \frac{-1}{(F_A + F_S)} \neq 0$$

Therefore, we conclude that the system is controllable.

What if the number of inputs and outputs are not equal?

- If the number of (independent) controlled variables is greater than the number of manipulated variables, the system is not controllable.
- If the number of controlled variables is less than the number of manipulated variables, the rank of the gain matrix must be equal to the number of controlled variables. This means that at least one selection of manipulated variables can be used to control all outputs.

The conclusion about controllability depends on the process and not on a specific control algorithm. If the system is not controllable, the conclusion is valid for any possible feedback control. In addition, the conclusion depends on the "type" of controllability desired. The definition used here for controllability is applicable to stable, continuous processes that are operated at steady state. Batch processes are inherently dynamic, so that a different approach is required to determine controllability. Since controllability determines only whether control is *possible*; it does not evaluate the control performance. Therefore, a controllable system could provide excellent or unacceptable dynamic performance; consequently, further analysis is required. However, a system that is not controllable certainly is not acceptable. For a broader analysis of controllability, see Skogstad and Postlethwaite (2005).

6.4.2 Operating Window

The controllability criterion ensures that feedback control is possible, but it is exactly valid for only small (differential) changes. Therefore, we also need to determine whether the manipulated variables are "powerful" enough to compensate for expected disturbances. Therefore, the next stage of process analysis addresses the operating window.

The operating window defines the range of variables (set points and disturbances) over which the process can operate.

This issue is the topic of an operability chapter, so it is not covered here. However, one example is given to reinforce the relationship between the equipment design (that provides capacity) and process control (that can utilize the available capacity).

Example 6.12 Operating window: We wish to determine the operating window for the total flow and composition for the blending process in Figure 6.46. The maximum flow rates are 60 m^3/h for F_S and 30 m^3/h for F_A.

The solution is plotted in Figure 6.47, which for this simple system can be presented in two dimensions. The results represent the achievable set points for the base-case stream compositions. We see that a total flow of 60 m^3/h at a composition of 80% A is not possible.

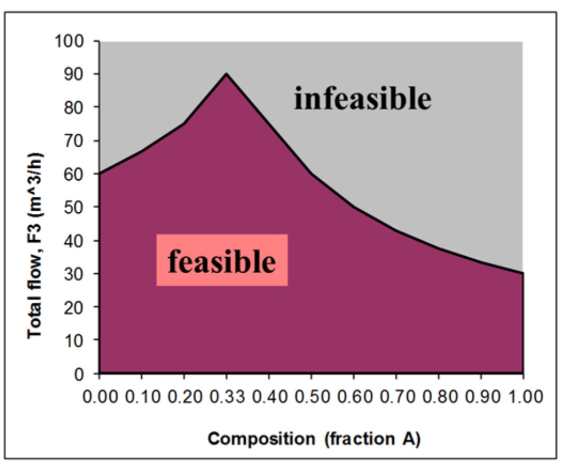


Figure 6.47 Operating window for the two-component blending process.

6.4.3 Multi-loop systems: Interaction

Let's consider processes that are controllable and have an acceptable range for compensation, i.e., have a sufficiently large operating window. We now venture into the large and complex topic of designing multi-loop control systems; by multiloop, we mean a control design consisting of multiple single-loop controllers using PID algorithms. This can be the topic of an entire book, so the discussions will be limited to understanding some basic concepts and key guidelines. Fortunately, a good understanding of process dynamics and the basics from control engineering can often provide insight for good designs, but considerable expertise can be required for complex processes.

First, we consider why multi-loop control is different from single-loop control, or stated differently, why can't each individual loop be designed without consideration for the other loops? The main reason is interaction.

Interaction exists when one manipulated variable can influence more than one controlled variable.

The importance of interaction is shown schematically in Figure 6.48. Manipulated variable MV_1 influences controlled variable CV_1, but MV_1 also influences other parts of the process, which have other controllers. We see that the total effect of MV_1 is the sum of

- the "direct effect" CV_A, which is independent of potential effects from other control loops (in manual) and
- the "interaction effect" CV_B, which is the result of (only) other control loops responding to the change in MV_1.

It is important to note that the measured variable CV_1 is the sum of CV_A and CV_B and that CV_A and CV_B do not exist as separate process variables – they cannot be independently measured. To determine the importance of interaction on the original control loop, we could evaluate the steady-state gain between MV_1 and CV_1 ($\Delta CV_1/\Delta MV_1$) for two scenarios, one with all other controllers in manual and another with all other controllers in automatic. This ratio is given in the following, with all changes based on the same change in MV_1 (ΔMV_1).

$$\text{Measure of interaction} = [\Delta CV_A/\Delta MV_1]/[(\Delta CV_A + \Delta CV_B)/\Delta MV_1] = \Delta CV_A/(\Delta CV_A + \Delta CV_B)$$

If this ratio is 1.0, interaction does not affect the steady state of MV_1-CV_1; if it differs from 1.0, interaction affects the steady state of MV_1-CV_1.

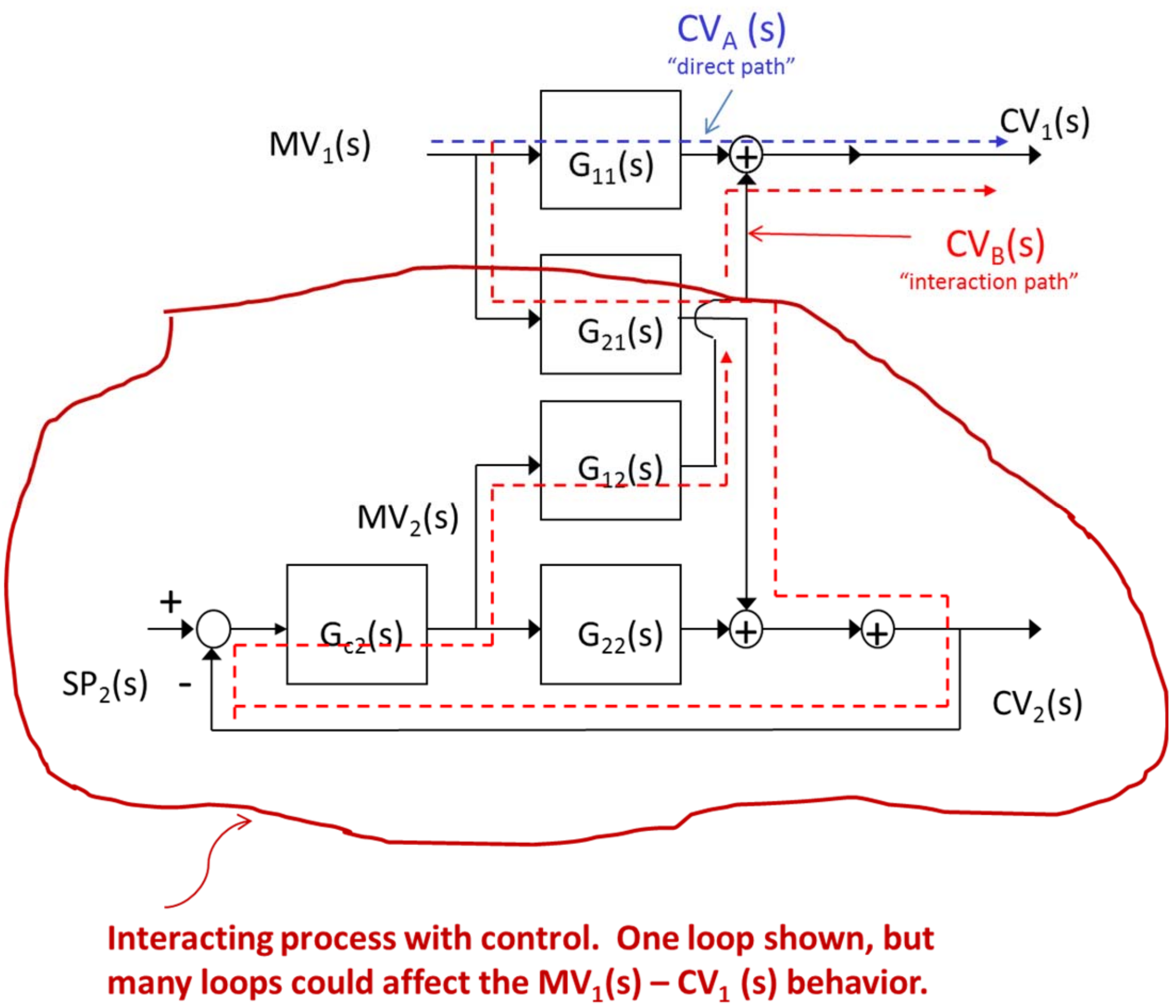


Figure 6.48 Interaction effects on $MV_1(s)$-$CV_1(s)$ due to associated control loops

In fact, this ratio was invented by Bristol (1966), who termed it the "relative gain" and represented it by the symbol λ_{ij}, with i = controlled variable and j = manipulated variable.

$$\lambda_{ij} = \frac{\left(\dfrac{\partial CV_i}{\partial MV_j}\right)_{\text{other loops open}}}{\left(\dfrac{\partial CV_i}{\partial MV_j}\right)_{\text{other loops closed}}} \tag{6.7}$$

Details on calculating the relative gain values for all possible control loop pairings are given in Sidebar I on Calculating the Relative Gain.

Sidebar I: Calculating the Relative Gain*

The relative gain is defined in the following equation.

$$\lambda_{ij} = \frac{\left(\dfrac{\Delta CV_i}{\Delta MV_j}\right)_{MV_k = \text{constant}}}{\left(\dfrac{\Delta CV_i}{\Delta MV_j}\right)_{CV_k = \text{constant}}} = \frac{\left(\dfrac{\Delta CV_i}{\Delta MV_j}\right)_{\text{other loops open}}}{\left(\dfrac{\Delta CV_i}{\Delta MV_j}\right)_{\text{other loops closed}}}$$

The relative gain array, which is a matrix, can be evaluated directly from the open-loop process gain matrix, K. (We use the Hadamard product, which is element-by-element multiplication.)

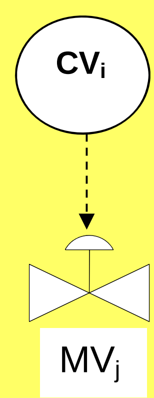
$$\Lambda = K \otimes \left(K^{-1}\right)^{\mathrm{T}} \qquad\qquad \lambda_{ij} = \left(k_{ij}\right)\left(kI_{ji}\right)$$

Thus, knowledge of the steady-state process gains enables us to evaluate the relative gains.

$$K = \begin{bmatrix} k_{11} & k_{12} & k_{13} & & k_{1n} \\ k_{21} & k_{22} & k_{23} & & \\ .. & & .. & & \\ .. & & & .. & \\ k_{n1} & .. & .. & .. & k_{nn} \end{bmatrix} \Longrightarrow \Lambda = \begin{bmatrix} \lambda_{11} & \lambda_{12} & \lambda_{13} & & \lambda_{1n} \\ \lambda_{21} & \lambda_{22} & \lambda_{23} & & \\ .. & & .. & & \\ .. & & & .. & \\ \lambda_{n1} & .. & .. & .. & \lambda_{nn} \end{bmatrix}$$

The relative gain λ_{ij} gives insight into behaviour for multiloop systems when CV_i is paired with MV_j. For example, the following proposed design controls CV1 By adjusting MV2, controls CV2 by adjusting MV1, and so forth.

$$\Lambda = \left[\begin{array}{c|cccc} & MV_1 & MV_2 & & MV_n \\ \hline CV_1 & \lambda_{11} & \lambda_{12} & & \\ CV_2 & \lambda_{21} & \lambda_{22} & & \\ .. & & & .. & \\ CV_n & \lambda_{n1} & .. & .. & \lambda_{nn} \end{array}\right]$$



We can determine some important properties about the control design from the values of the relative gains; see Table 6.6.

*** The relative gain calculation and interpretation presented here is limited to stable processes**

Possible values for the relative gain and how each affects control design are summarized in Table 6.6. We observe that interaction can have a profound effect on the behavior of multiloop systems.

While the relative gain only addresses the steady-state behavior of the system in Figure 6.48, it provides considerable insight into the effects of interaction. Even with only steady-state information, the contents of Table 6.6 show that many potential designs can be *eliminated from further consideration*, e.g., those with one or more negative relative gain values. Therefore, we will use relative gains as one tool for deciding the proper control loop pairing.

We conclude that the relative gain provides useful insights and quantitative analysis. Some potential designs can be eliminated, but loop-pairing designs <u>cannot be based solely on relative gain values</u>.

Many additional performance issues are important in designing the feedback controller loop pairing.

- Pair manipulated and controlled variables with fast feedback dynamics, emphasizing the reduction in loop dead times.
- Pair manipulated and controlled variables to give large process gains (K_{ij}). This provides the controller with a large operating window and tends to reduce interactions. Here, "large" does not necessarily mean a large absolute value; remember that the gain has units. When evaluating the process gain, it is best to express a dimensionless gain, as given in the following.

$$K_{ij}^* = K_{ij} \frac{range\ of\ manipulated\ variable}{range\ of\ controlled\ variable}$$

  The range of the manipulated variable would be (i) typically 0-100% when adjusting a valve directly or (ii) the *secondary* sensor range when adjusting a set point in a cascade. The range of the controlled variable is the expected range of the measured variable would experience, which would normally be the sensor range. When expressed as a dimensionless value, the process gain (K_{ij}^*) should be near 1.0.

- Associated with the above point, a design should favor loop pairings that have relative gains around 1.0. Loops with positive relative gains but with a significantly different from 1.0 (e.g., 5.0 or 0.20) could require controller tuning to be adjusted depending on the manual-automatic statuses of interacting loops. While such retuning can be implemented automatically, the resulting control design is substantially more complex.

- Designs should avoid "nested loops", where one control loop will function properly only when another interacting control loop is in automatic. An example of a design with nested loops is shown in Figure 6.49a. The flow controller depends on the proper functioning of the level controller. An acceptable alternative design is shown in Figure 6.49b.

Table 6.6. Possible relative gain values and implications for control design.*

Case	ΔCV_A	ΔCV_B	$\Delta(CV_A+CV_B)$	Relative gain, λ	Control Design
I	0	$\neq 0$	$\neq 0$	0	Feedback control is not possible when other loops are in manual. **In nearly all situations, this loop pairing would be eliminated from consideration.** (However, a special case where this design is used is presented later in this chapter.)
II	$\neq 0$	0	ΔCV_A	1.0	Feedback control is possible, and the steady-state effect of interaction is zero. Interaction does not affect the steady-state relationship between CV_i and MV_j. This loop remains a candidate for implementation.
III	+	-	+ $\lvert\Delta CV_A\rvert > \lvert\Delta CV_B\rvert$	$\lambda > 1.0$	Interaction reduces the steady-state gain between MV_j and CV_i. This results in smaller feedback gain, but the loop would remain a candidate for implementation.
IV	+	+	+	$0 < \lambda < 1$	Interaction increases the steady-state gain between MV_j and CV_i. This could result is overly aggressive feedback unless the controller gain (K_C) is reduced for the multi-loop situation. The loop would remain a candidate for implementation.
V	+	-	- $\lvert\Delta CV_A\rvert < \lvert\Delta CV_B\rvert$	$0 < \lambda$	Interaction changes the sign of the steady-state gain! For stable feedback, the controller gain (K_C) would have to be modified to maintain negative feedback as the status (manual-automatic) of the interacting loops changes! **In essentially all situations, this loop pairing would be eliminated from consideration.**
VI	+	-	0 $\lvert\Delta CV_A\rvert = \lvert\Delta CV_B\rvert$	∞	Feedback control is not possible when other loops are in automatic. **This loop pairing would be eliminated from consideration.**

* Table includes only $\Delta C_A \geq 0$. Student exercise is to extend table for $\Delta C_A \leq 0$.
* These interpretations are applicable to processes that are stable (without control)
* All feedback controllers have an integral mode

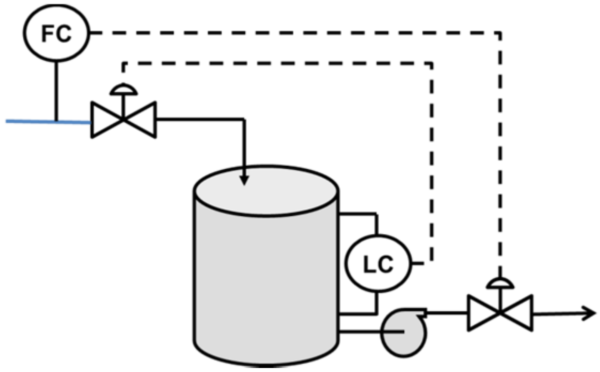


Figure 6.49a Example of nested control loops (relative gain = 0)

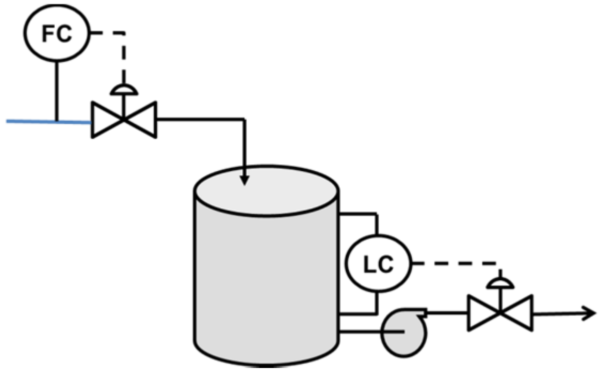


Figure 6.49b. Example with nested loops reconfigured.

Example 6.13 Interaction measure in distillation – You desire to control both product compositions in the distillation tower shown in Figure 6.50. The pressure and both level controllers have already been designed. You have performed experiments to determine approximate linear models for the composition control, with the results given in the following expression; note that the matrix contains individual input-output models.

$$\begin{bmatrix} XD(s) \\ XB(s) \end{bmatrix} = \begin{bmatrix} \dfrac{0.0747e^{-3s}}{12s+1} & \dfrac{-0.0667e^{-2s}}{15s+1} \\ \dfrac{0.1173e^{-3.3s}}{11.75s+1} & \dfrac{-0.1253e^{-2s}}{10.2s+1} \end{bmatrix} \begin{bmatrix} F_R(s) \\ F_V(s) \end{bmatrix} + \begin{bmatrix} \dfrac{0.70e^{-5s}}{14.4s+1} \\ \dfrac{1.3e^{-3s}}{12s+1} \end{bmatrix} X_F(s)$$

with

XD	=	***Light key in distillation***
XB	=	***Light key in bottoms***
F_R	=	***Reflux flow rate***
F_V	=	***Vaporized flow from reboiler***
X_F	=	***Light key in the feed***

Evaluate the interaction measure and discuss the influence of these results on the control loop pairing composition control.

The steady-state gains in the dynamic model can be used to evaluate the relative gain array, which is given in the following.

	F_R	***F_V***
XD	***6.09***	***-5.09***
XB	***-5.09***	***6.09***

We observe that only one of the possible two loop parings has positive relative gains. Therefore, the designer must select the positive pairing, following the recommendations in Table 6.6.

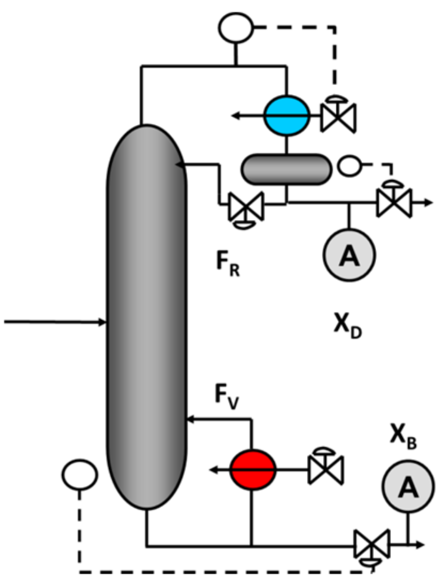


Figure 6.50 Distillation tower for Example 6.13.

6.4.4 Multi-loop systems: Control loop pairing

With the process design complete with sensors and final elements included, the multi-loop design decision becomes the pairing of measured controlled variables with manipulated variables, which can be final elements of set points in cascade designs. The approach usually taken is to eliminate possibilities that cannot provide adequate control performance. Then, evaluate the remaining for dynamic performance. The criteria used here are summarized in Table 6.7 and are discussed and applied in the following.

6.4.4a. Ensuring process controllability – This topic has been addressed in Section 6.4.1. The process must be controllable before the engineer proceeds to designing a control strategy.

Table 6.7. Criteria to determine loop pairing

• **Controllability**	Ability to control selected CVs with selected MVs
• **Integrity**	Ability for control system to function properly when one or more control loops is not functioning
• **Range**	Sufficiently large operating window when control system is compensating for disturbances
• **Important variables**	Performance of the most important variable(s) is the best possible, perhaps at the expensive of poorer performance of less important variable(s)
• **Expected scenarios**	Control performance is best possible for the disturbances that are expected to occur most frequently, perhaps at the expensive of poorer performance for other disturbances

6.4.4b. Ensuring design integrity –Control integrity is defined in the following.

A control system has integrity if, when one or more of the controllers is not functioning, the remaining feedback control system is stable without changing the sign of any controller gains.

When a feedback controller is not functioning, it is not adjusting the manipulated variable. This can occur when the controller is placed in manual (off) or when the controller remains in automatic (on) but the manipulated variable reaches an upper or lower bound.

Control designs with pairings on loops with positive relative gains satisfy this integrity criterion. Therefore, positive relative gain elements are required for good control design. A rigorous check requires that all sub-systems also have positive relative gains. For example, a

3x3 system must be paired on positive relative gains. Then, each 2x2 system resulting in one controller being "off" must have positive relative gain pairings. This checking procedure can involve many instances for a large number of controlled and manipulated variables.

Note that the system is not guaranteed to be stable with one or more controllers off. However, the system will be stable for retuned feedback controllers, with the retuning not changing the sign of the feedback controllers.

Example 6.14 applies the integrity requirement for distillation control.

Example 6.14. Ensuring integrity in Blending- We must control the composition and total blended flow rate from the blending process in Figure 6.46. We have already concluded in Example 6.11 that the process is controllable. The product specification is 5% component A. Select a design with good integrity.

We evaluate the relative gain matrix, which can be done using the steady-state gain matrix.

	F_A	F_S
x_{AM}	0.95	0.05
F_M	0.05	0.95

$$\boldsymbol{K} = \begin{bmatrix} 1 & 1 \\ 0.95 & -0.05 \end{bmatrix}$$

By applying the criteria in Table 6.6, we conclude that we cannot eliminate either loop pairing, because the relative gains for each are finite and positive. Therefore, both have acceptable integrity. Dynamics are important, so we consider the feedback dynamics and observe that the responses will be fast for all valves to all sensors (as long as the composition analyzer is close to the mixing point).

When the relative gain is near 1.0, the "effective process gain" and therefore, the controller gain remains relatively unchanged, whether the interacting loop in "on" or "off". However, if the relative gain is far from 1.0, the "effective process gain" changes significantly depending on whether the interacting controller is "on" or "off". Therefore, the controller gain yielding stable feedback with good performance is very different whether the interacting control is "on" or "off". For the case with relative gains far from 1.0, the controller gain must be adapted in real-time based on the status of the interacting controller. The adaptation can be achieved, but it is to be avoided unless absolutely necessary. Therefore, we select the pairing x_{AM} –F_A and F_M – F_S., which has a relative gain much nearer 1.0.

Note that these results depend strongly on the nominal steady-state operating conditions. Would the results change if the nominal operation were changed from 5% to 95% A in the mixture?

6.4.4c. Achieving control range – The range addresses an acceptably large operating window, so that the control system can compensate for expected disturbances and achieve the desired range of set points. This topic was addressed in Section 6.4.2; the extension of the design guideline presented here is that the control system should be able to achieve the required region of the operating window based on the actions of the process controllers.

If the original design does not provide adequate coverage of the operating window, the control design can be enhanced using split-range technology. Example 6.5 shows a control design with split range technology to ensure that the automatic controller can achieve the appropriate range of fuel gas manipulation.

6.4.4d. Emphasizing the most important variables – Often, one or more of the controlled variables are much more important than the remaining controlled variables. In this situation, the control design should be designed for fast, large range feedback for the more important controlled variables. This might result is somewhat poorer performance for controlled variables of lesser importance. A proper balance must be achieved by the design engineer.

Another approach for achieving better performance of selected controlled variables is to tune their feedback controllers more aggressively. This can be achieved if other feedback controller are tuned loosely, giving slower and poorer feedback control to the less important controlled variables. Again, a proper balance must be achieved by the design.

Example 6.15 Gaining fast control for important variable – In the fuel gas distribution system in Figure 6.51, two sources of fuel (fuel gas and vaporized fuel) are provided to supply the total demand for numerous consumers. The consumers demand flow rates independent of each other and of the fuel gas system. The control system must provide the quantity of fuel, and in this design, it must also regulate the heating value of the fuel to the consumers.

The control system balances the demand and supply by controlling the pressure in the distribution system (header). The heating value is controlled using an on-stream analyzer that adjusts one of the fuel sources. How should the two loops be paired? We base this decision on the need for fast pressure control, while the heating value can experience short-term deviations from its set point. Therefore, we control pressure by adjusting the faster manipulated variable, the fuel gas, and we design the controllers to control heating value by adjusting the remaining manipulated variable, steam to the liquid fuel vaporizer.

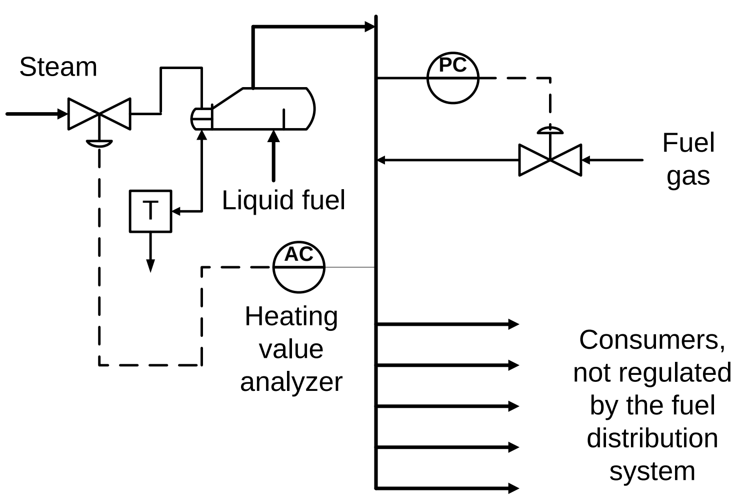


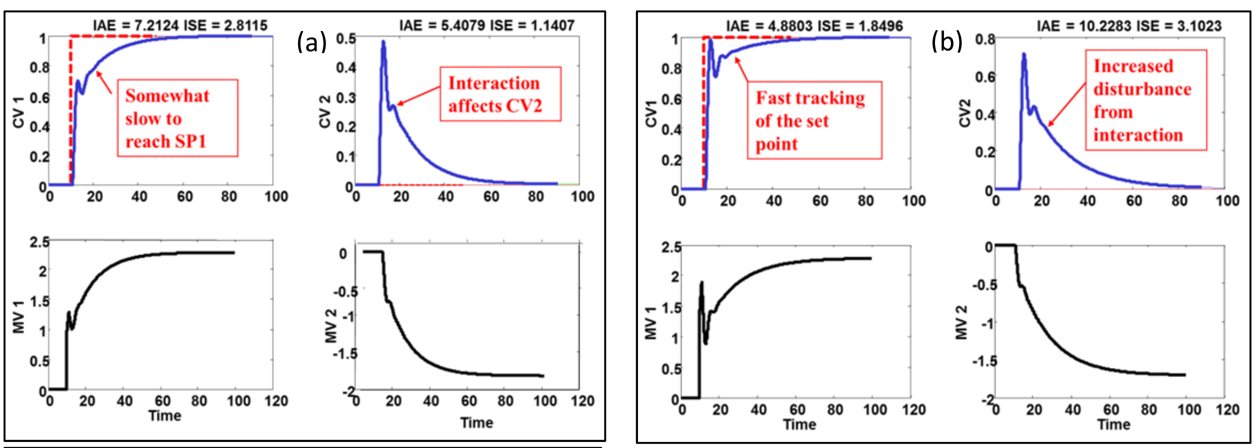
Figure 6.51 Fuel distribution system in Example 6.15

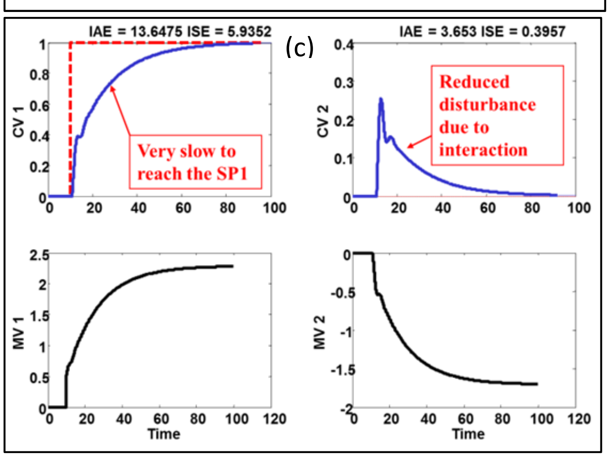
Example 6.16 Tuning matched to control goals – Let's consider a hypothetical two-variable process with the following dynamic model.

$$\begin{bmatrix} CV_1(s) \\ CV_2(s) \end{bmatrix} = \begin{bmatrix} \dfrac{1.0e^{-1.0s}}{1+2s} & \dfrac{0.75e^{-1.0s}}{1+2s} \\ \dfrac{0.75e^{-1.0s}}{1+2s} & \dfrac{1.0e^{-1.0s}}{1+2s} \end{bmatrix} \begin{bmatrix} MV_1(s) \\ MV_2(s) \end{bmatrix}$$

We desire to control both output variables, and we will pair the variables on their largest gains, CV_1-MV_1 and CV_2-MV_2. We encounter three common situations (i) CV_1 is more important, (ii) CV_1 and CV_2 are of equal importance, and (iii) CV_2 is more important. How can we adjust the control system to achieve one of these goals, while noticing that the dynamics are equal for both manipulated variables? The answer is controller tuning. The controller for the more important variable can be tuned tightly, with the interacting controller detuned to maintain stability. Note that simply increasing the controller gain of the more important while leaving the interacting controller(s) unchanged could lead to overly aggressive and even unstable behavior.

If the controlled variables are of equal importance, the two controllers are tuned to give approximately the same aggressiveness and performance. Typical dynamic performances for three situations are given in Figure 6.52. More details on this approach and case study, including the tuning, is available in Marlin (2000, Chapter 20).





(Note the differences in CV_2 scales.)

Figure 6.52 Multiloop PID tuning to match the controlled variables goals. (CV's plotted as deviation from initial steady state.) (Marlin, 2000)

a. CV_1 and CV_2 are equally important
b. CV_1 is more important
c. CV_2 is more important

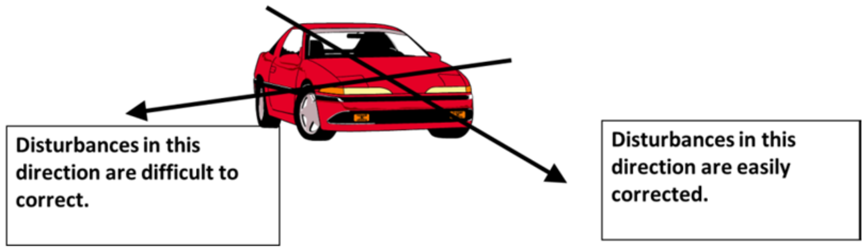


Figure 6.53 Directionality for feedback and two disturbances depicted for an automobile.

6.4.4e. Achieving good performance for expected disturbance scenarios– Many possible disturbances can affect a process, but often, one or a few disturbances occur most frequently at significant magnitudes to affect process performance. In single-loop control, the type of disturbance was not considered because the guidelines in Table 6.5 apply to all disturbances, e.g., feed rate, feed composition, cooling water, etc. A major difference of multivariable systems is that they have key characteristic not appearing in single-loop systems. While single-loop and multiloop systems are influenced by disturbance size and frequency, only multiloop systems are also affected by "direction".

To understand the concept, let's consider the automobile shown in Figure 6.53. If the automobile were pushed forward (or backward), the driver could easily compensate and return the automobile to it's the original position. However, if the automobile were pushed sideways, the driver would have to iterate forward and backward, slowly making progress returning to the original position. Disturbances of the same magnitude are easy or difficult to control based on the directions of the disturbance and feedback process. If they are "aligned", control performance is good. If they are orthogonal, feedback control is more difficult and performance degrades. Directionality analysis applied to the automobile also applies to process control.

Example 6.16 Distillation directionality – We will investigate the performance of the two distillation quality control strategies shown in Figure 6.54. The designs are two common approaches to achieve level and composition control in two-product distillation. We note that the material balance design has a relative gain for the two composition loops of 0.39, while the energy balance has 6.09. Clearly, the energy balance control has greater interaction, as measured by the relative gain metric. But, how do the designs compare for disturbances?

The most common unmeasured disturbance is feed composition. The dynamic responses for each control system responding to a step disturbance in feed composition are given in Figure 6.55. It is clear that the energy balance design has better performance, with lower total IAE (and lower ISE) and a shorter settling time.

The result contradicts s design rule sometimes proposed that the best control design relative gain values nearer to 1.0. In this case, and in others of importance in industry, the dynamic performance is best for the design when the feedback direction is aligned with the disturbance direction, whether or not the relative gain is closer to 1.0.

A full coverage of the directionality concepts is not possible in this process-design oriented chapter. The interested reader is referred to material on quantitative analysis of directionality in McAvoy et. al. (1985), Skogestad and Morari (1987) and Marlin (2000, Chapter 21).

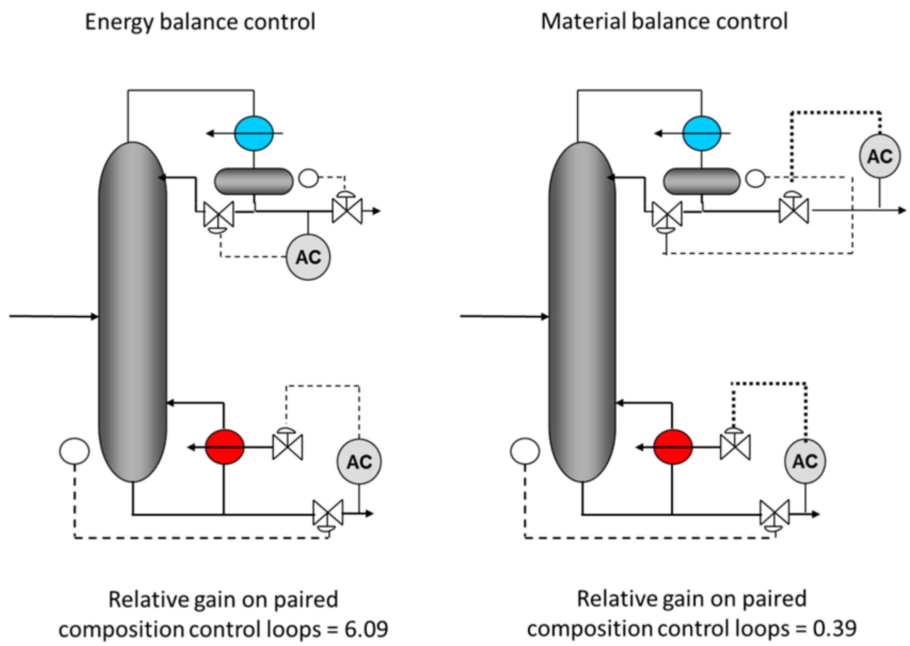


Figure 6.54 Candidate distillation control strategies for Example 6.16.

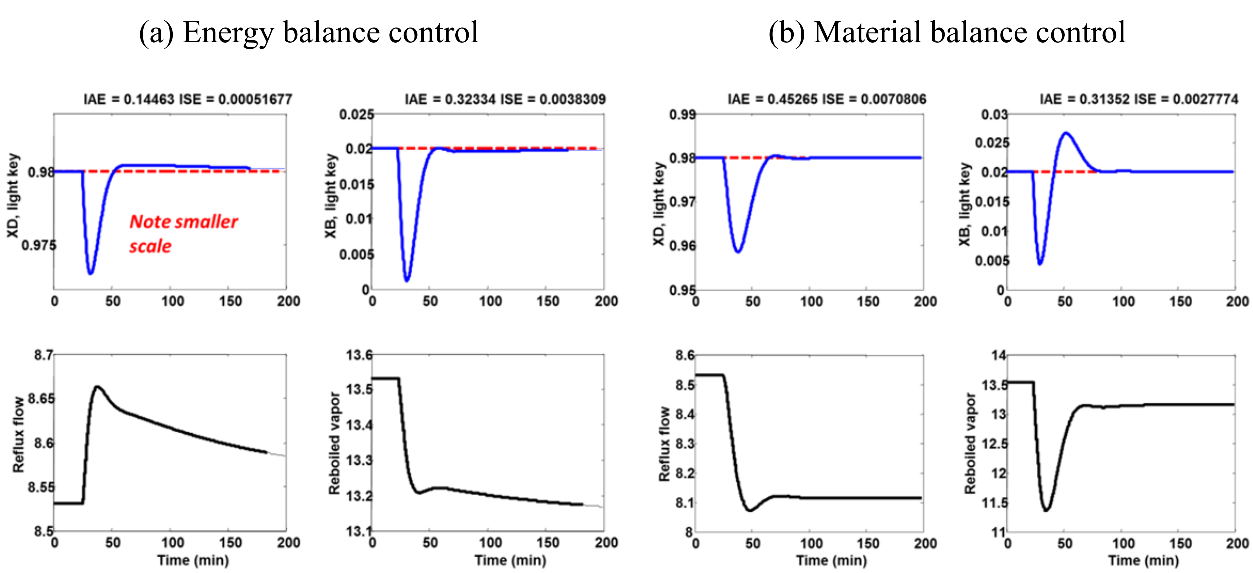


Figure 6.55. Dynamic behavior for two distillation tower control designs. (Marlin, 2000)

There are situations where engineers can encounter conflicts in applying the five guidelines for loop pairing in Table 6.7. Let's consider an example where a conflict among the guidelines occurs.

Example 6.17Conflicts in the control of a pyrolysis heater – Two basic control objectives for a fired heater are (i) process fluid throughput flow control and (ii) process fluid outlet temperature control. The manipulated variables are the feed flow valve and the fuel flow valve. (We will assume that adequate air is provided to the burner.) What is the proper loop pairing?

The conventional design is shown in Figure 6.56a. There is a causal relationship between the controlled and manipulated variables, the pairing is on positive relative gain elements, and the dynamics of the temperature control are usually adequate, although having minutes of delay and time constant. However, the conventional control does not provide extremely tight temperature control and cannot provide separate control for multiple passes (pipes). Furnaces with multiple passes operating at high temperatures require independent temperature control for each pass. The strategy in Figure 6.56b achieves the tight, multi-pass temperature control. However, it suffers a disadvantage. The causal relationship between the total feed flow controlled variable and the manipulated fuel requires the temperature controllers to be in automatic (on). Thus, the design in Figure 6.56b has "nested loops" and has a pairing on a zero relative gain. These characteristics are to be avoided if possible, but they can be tolerated if there is an overwhelming advantage. When applied, the design in Figure 6.56b should have a monitor that does not allow the flow controller to be in automatic unless all temperature controllers are in automatic.

The conventional design in Figure 6.56a is applied in essentially all heaters whose purpose is to raise a stream temperature. The unconventional design in Figure 6.56b is used by some practitioners for heaters with multiple passes where chemical reactions occur in the passes.

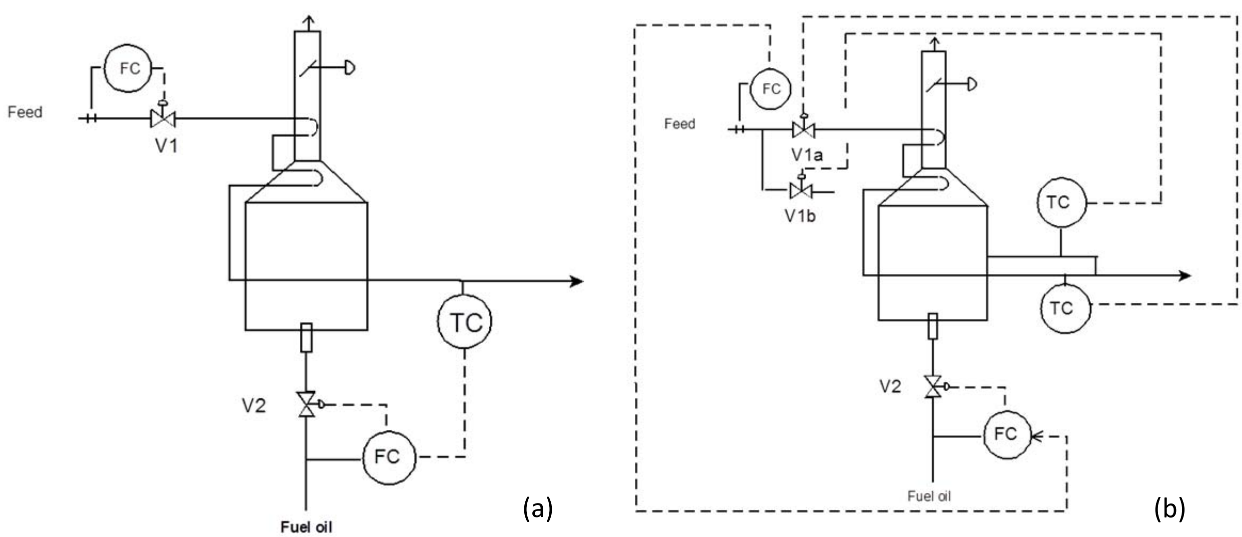


Figure 6.56 Two fired heater control strategies.
(a) Conventional with positive relative gain
(b) Unconventional with zero relative gain but fast, independent temperature loop dynamics for every pass

This example serves as a warning about guidelines. Good guidelines are often correct and give useful assistance to the engineer; however, guidelines are not fundamental principles. The engineer must understand the basis for the guidelines, know when they should be observed, and know when they should be violated, with care.

The presentation of multiloop control in this section has introduced many key issues and provided guidance on analysis methods and design technology. The reader should recognize that this is an enormous topic and full coverage would require a separate book. The reader can access helpful materials given in Additional Learning Resources. In should also becoming clear that control design requires the understanding of both process and control technology, and each design is based on a thorough definition of the pertinent issues, including performance goals, process behavior, relevant constraints, key disturbances, and instrumentation available. The next section provides guidance on a structured design procedure to integrate all pertinent issues.

6.5 Control System Design Procedure

As is often stated by professors, “If you don’t know where you are going, any path will do.” Certainly, this old adage applies to control system design. A great deal of important information is contained in the process material and energy balances and the process drawings. However, the complete definition of the required control performance (safety, economics, production rate, product quality, etc.) and the situations expected in the plant (disturbances, constraints, etc.) are not. Here, a method for defining the problem and steps for completing the design are presented and applied to a simple example, the flash separation process in Figure 6.57.

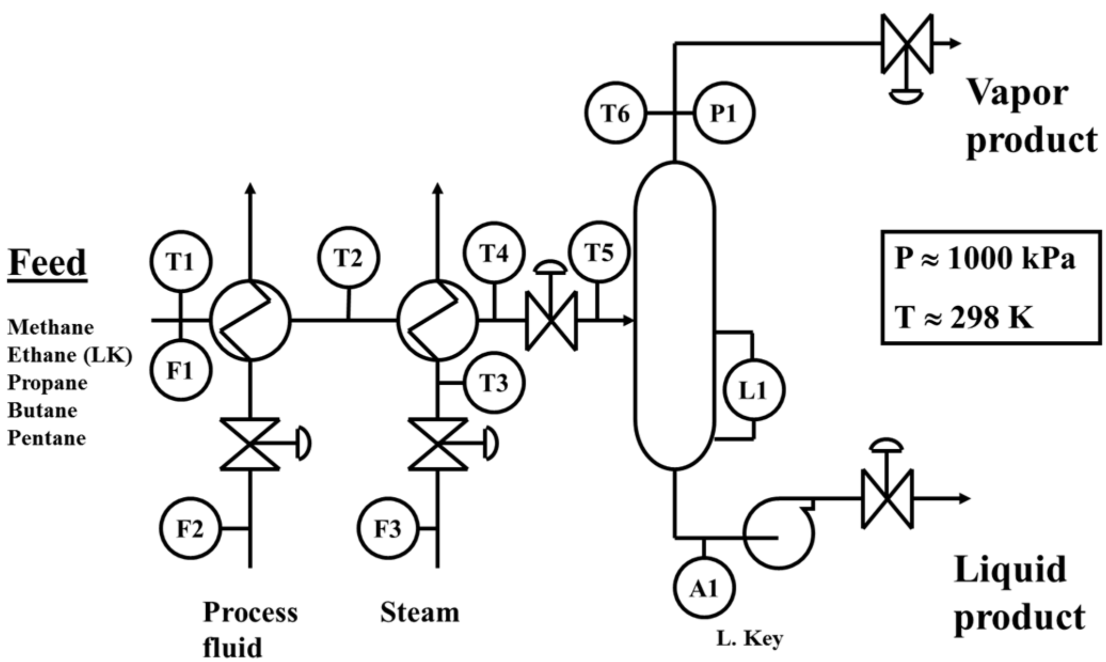


Figure 6.57 Flash process with preheat. (Marlin, 2000)

6.5.1 Defining the control design problem

The design definition is summarized in the control design form. The definition is a summary of business data, performance targets, and preliminary process flowsheet and equipment information. Business data includes feed, product and fuel values, the ranges for raw material properties, production rates, and product qualities, including multiple product specifications for similar materials. Performance targets involve safety and reliability targets (covered in more detail elsewhere in this monograph). Preliminary process design includes equipment type (e.g., centrifugal or positive displacement pump), equipment sizing, pressure rating and instrumentation selection. When designing controls, the engineer must re-evaluate all of these decisions and modify them as required.

A sample control design form is presented in Table 6.8 for the example process. The first entry in the table after the heading presents the control objectives using the seven categories proposed by Marlin (2000). The definitions should be as detailed as possible, giving, for example, sizes of disturbances, set point changes, and allowable deviations in key variables. The control objectives should not contain or imply design solutions; they should be based on desired operation for safety, product quality, profit, and so forth. Note that issues that do not fit the general categories can be documented in the table's last entry, "additional considerations".

The next two entries define the instrumentation, measurements and final elements. Clearly, these preliminary decisions will be evaluated and possibly improved during the control design.

The next category defines the constraints or limitations that influence dynamic operation. Some of the constraints result from equipment performance limitations. Other constraints result from bounds on operating conditions outside of which the process could be unsafe or produce useless products. Violations of "hard" constraints lead to unsafe operation or extreme economic loss, such as shutting down the process for repairs. Violations of "soft" constraints can be tolerated, at least for limited duration, but with economic loss; an example could be a high temperature in a reactor that would degrade the catalyst activity.

The following category defines the key disturbances, giving approximate ranges and frequencies of occurrence. Engineers accustomed to operability analysis will be able to identify these disturbances from knowledge of the integrated processes and the equipment being used.

Certainly, the process dynamics should be understood for control design. Dynamic models for all inputs and outputs would be ideal; however, empirical models cannot be determined because the process does not exist, and fundamental models often require excessive time and cost to develop. As a minimum, all key aspects of the dynamics should be noted, including open-loop unstable variables, variables affecting safety, very slowly responding variables, etc. For new processes, where experience does not exist, a fundamental analysis of dynamics using simulation might be warranted.

Table 6.8. Preliminary Control Design Form for the flash process in Figure 6.57 (Marlin, 2000)

TITLE:	Flash drum	ORGANIZATION:	Profit, Inc.
PROCESS UNIT:	Hamilton chemical plant	DESIGNER:	I. M. Learning
DRAWING :	Figure 6.57	ORIGINAL DATE:	January 1, 2011

CONTROL OBJECTIVES:

1) Safety of personnel
   a) the maximum pressure of 1200 kPa must not be exceeded under any (conceivable) circumstances
2) Environmental protection
   a) material must not be vented to the atmosphere under any (conceivable) circumstances
3) Equipment protection
   a) the flow through the pump should always be greater than or equal to a minimum
   b) cavitation in the pump should be prevented
4) Smooth, easy operation
   a) the feed flow should have small variability
   b) product flows should have modest variability, not greater in percentage than the feed flow variability, which includes production rate changes
5) Product quality
   a) the steady-state target of value 10 mole% of ethane in the liquid product should achieved for operating condition changes of +20 to -25% feed flow, 5 mole% changes in the ethane and propane in the feed, and -10 to +50 °C in the feed temperature.
   b) the ethane in the liquid product should not deviate more than ±1 mole % from its set point during transient responses for the following disturbances
      i) feed temperature experiences a step from 0 to 30 °C
      ii) feed composition experiences a step of +5 mole% ethane and -5 mole% of propane
      iii) feed flow set point changes 5% in a step
6) Efficiency and optimization
   a) the heat transfer should be maximized from the process integration exchanger before using the more expensive steam exchanger
7) Monitoring and diagnosis
   a) sensors and displays needed to monitor the normal and upset conditions of the unit must be provided. Typical faults for this process would involve sensor or valve failures that would affect the flash temperature and/or pressure.
   b) sensors and calculated variables required to monitor the product quality and thermal efficiency of the unit should be provided for longer term monitoring. One typical slowly changing aspect of this process would be heat exchanger fouling.

MEASUREMENTS:

Variable	Principle	Value, Sensor Range	Special Information
A1	chromatograph	10, 0-15 mole%	update every 2 minutes
F1	orifice	100, 0-200	
F2	orifice	120, 0-200	
F3	orifice	100, 0-200	
F4	orifice	45, 0-90	
F5	orifice	55, 0-110	
L1	delta pressure		range is lower half of drum
P1	piezo electric	1000, 500-1500 kPa	
T1	thermocouple	0, (-)50-100 °C	
T2	thermocouple	25, 0-100 °C	
T3	thermocouple	90, 0-200 °C	
T4	thermocouple	45, 0-200 °C	
T5	thermocouple	25, 0-100 °C	
T6	thermocouple	25, 0-50 °C	

MANIPULATED VARIABLES:

I.D.	Maximum capacity (at design pressures) (%open at design, maximum flow)
v1	100%, 100
v2	53%, 189
v3	50%, 200
v4	14%, 340
v5	52%, 106

CONSTRAINTS:

Variable	Limit values	Measured/ inferred	Hard/ soft	Penalty for violation
drum pressure	1200 kPa, high	P1, measured	hard	personnel injury
drum level	15%, low	L1, measured	hard	pump damage
Ethane in F5 product	10 ± 1 mole%, (max deviation)	A1, measured & T6, inferred	soft	reduced selectivity in downstream reactor

DISTURBANCES:

Source	Magnitude	Dynamics
feed temperature (T_1)	-10 to 55°C	infrequent step changes of 20°C magnitude
feed rate (F_1)	70 to 180	set point changes of 5% at one time
feed composition	±5 mole% ethane	approximately periodic changes (every 2-3 hr)

DYNAMIC RESPONSES: (input = all manipulated variables and disturbances)
(output = all controlled and constraint variables)

Input	Output	Gain	Dynamic model

(Determined empirically after the process has been started up)

Bottoms liquid level	- This variable is unstable and must be controlled.
Drum Pressure	- This is a critical safety and process environment variable that changes rapidly and can exceed limits quickly. Feedback control is required.
Flash temperature	- The temperature is not high enough to damage equipment. Good temperature control is essential to achieve the desired separation
Feed flow rate	- The feed flow rate can vary but should do so slowly so that the flash pressure and temperature do not deviate significantly from their set points
Bottoms % ethane	- This is the purpose of the equipment! In this process, the specifications allow little variation, so an analyser is provided for on-stream analysis

ADDITIONAL CONSIDERATIONS:

a) Liquid should not exit the drum via the vapor line
b) The composition analyzer is much less reliable than other sensors

The final entry addresses the important topic of monitoring and diagnosis, which should be addressed at the design stage to ensure that adequate sensors and sample locations are provided. Knowledge of equipment enables engineers to define the likely faults that result from equipment malfunction and human failure. Then, important sensors for identifying each fault could be determined using a cause-effect diagram.

This form may seem a bit pedantic, requiring excessive documentation for every decision, and in fact, most control designs are performed in practice without such extensive documentation. The form is used here because it provides an excellent structure for beginning engineers who after gaining proficiency, may perform the analysis without the form. However, even the most experienced engineers benefit from this type of documentation for complex designs.

It is important to recognize that experienced engineers can *sometimes* by-pass the Control Design Form (CDF) documentation, but they *always* perform a thorough analysis involving information and issues included in the CDF.

6.5.2 Designing the control system

Describing a design procedure is always difficult, because individuals solve problems using many good (and some poor) procedures that are tailored to the specific problem. A typical sequence of analysis steps are proposed here that should help the reader and inexperienced engineer address complex control design issues. As the individual gains experience, he/she can personalize the sequence.

The control design procedure shown in Table 6.9 begins by collecting basic information on the preliminary process design. Then, the control design form is completed to document the information required for the design. Next, the feasibility of achieving the goals is evaluated by ensuring that the design has the adequate number of degrees of freedom and adequate capacity, i.e., the operating window is large enough. The following step involves understanding the process and operational goals, approximate dynamics, constraints, disturbances, and so forth that influence the design. More quantitative analysis, such as relative gain evaluation, could be performed at this step. Then, the design is begun by defining controls in the following sequence.

- The overall flow and inventory control
- Each unit control
  - Flow and inventory
  - Process environment (pressure, temperature, flow ratios, etc.)
  - Product or intermediate stream qualities
  - Safety as required
- Optimization
- Monitoring and diagnosis
  - Short-term monitoring of fast-occurring situations
  - Long-term monitoring of process behavior

Table 6.9. A Control Design Procedure. Marlin (2000)

START: Acquire information about the process

(a) Process equipment and flow structure [Modify process and instrumentation]
(b) Operating conditions
(c) Product quality and economics
(d) Preliminary location of sensors and final elements

1. DEFINITION: Complete the Control Design Form

(a) Use checklists
(b) Sample questions [Modify objectives]
(c) Prepare a preliminary set of controlled variables

2. FEASIBILITY: Determine whether objectives are possible

(a) Degrees of freedom [Iterate]
(b) Select controlled variables and evaluate controllability
(c) Operating window for key operating conditions

3. OVERVIEW: Develop understanding of entire process to enable "look-ahead" in decisions

(a) Key production rate variables
(b) Inventories for potential control
(c) Open-loop unstable processes
(d) Complex dynamics (long delays, inverse response, recycle, strong interactions)
(e) Key product qualities
(f) Key constraints
(g) Key disturbances
(h) Useful manner for decomposing the analysis (and control design), if necessary and appropriate

4. CONTROL STRUCTURE: Selection of controlled and manipulated variables, interconnections (pairings in decentralized control), and relevant tuning guidelines

(a) Preliminary decisions on overall process flows and inventories
(b) Process segment (Unit) 1
(c) Process segment (Unit) 2
Control hierarchy (temporal decomposition) for every unit [Iterate]
1. Flow and inventory
2. Process environment [Modify control]
3. Product quality
4. Safety

(d) Integrate control designs as needed for good overall performance

5. OPTIMIZATION: Strategy for excess manipulated variables

(a) Clear strategy for improved operation, or
(b) Measure of profit using real-time data
(c) Sensors and final elements [Iterate]
(d) Minimize unfavorable interaction with product quality

6. MONITORING AND DIAGNOSIS

(a) Real-time operations monitoring
1. Alarms
2. Graphic displays and trends [Iterate]

(b) Process performance monitoring
1. Variability of key variables (histogram and frequency range)
2. Calculated process performances (efficiencies, recoveries, etc.)

FINISH: Completed specification, meeting objectives in step 1

(a) Process equipment and operating conditions
(b) Control equipment, sensors, and final elements
(c) Control structure and algorithms
(d) Tuning guidelines as needed, e.g., level control and interacting loops
(e) Safety controls and alarms
(f) Optimization
(g) Monitoring calculations

The design sequence does not necessarily involve a linear sequence of steps. As some design decisions are evaluated, the engineer could decide that either the process equipment design or previous control decisions should be modified. If so, iteration is required to modify previous decisions in a manner that enables the engineer develop designs that meet all goals. To perform the design with minimal iterations, the design engineer must be able to "look ahead" and anticipate later aspects of the design. For example, when designing the production rate and inventory controls, the engineer should not use valves for production rate control that are more appropriately used for important safety and product quality control.

6.5.3 Control design for the flash process

The control design procedure can be applied to the flash process in Figure 6.57. The entire procedure is presented in Chapter 24 of Marlin (2000). Here, a few key results are presented along with the final design. We begin by provisionally defining five controlled variables as the vessel pressure, liquid level, feed flow rate, flash temperature, and ethane composition in the liquid product. Next, we determine if a sufficient number of valves exist. We see that five valves are present, which is sufficient to control the five controlled variables. However, we must investigate further by evaluating the controllability using the steady-state gain matrix. (Note that the level is unstable without control and has no steady-state gain, but the analysis can proceed using the derivative of the level (McAvoy, 1983).)

$$\begin{bmatrix} F1 \\ T6 \\ A1 \\ P1 \\ dL_1/dt \end{bmatrix} = \begin{bmatrix} 0 & 0 & 2.0 & 0 & 0 \\ .0708 & .85 & -.44 & 0 & -.19 \\ -.00917 & -.11 & -.44 & 0 & .043 \\ .567 & 6.80 & 1.39 & 0 & -5.36 \\ -.0113 & -.136 & .31 & -.179 & -.0265 \end{bmatrix} \begin{bmatrix} v1 \\ v2 \\ v3 \\ v4 \\ v5 \end{bmatrix}$$

We evaluate the determinant of the gain matrix.

x 12

$$\begin{bmatrix} F1 \\ T6 \\ A1 \\ P1 \\ dL/dt \end{bmatrix} = \underbrace{\begin{bmatrix} 0 & 0 & 2.0 & 0 & 0 \\ .0708 & .85 & -.44 & 0 & -.19 \\ -.00917 & -.11 & -.44 & 0 & .043 \\ .567 & 6.80 & 1.39 & 0 & -5.36 \\ -.0113 & -.136 & .31 & -.179 & -.0265 \end{bmatrix}}_{\text{Det } [K_P] = 10^{-7}} \begin{bmatrix} v1 \\ v2 \\ v3 \\ v4 \\ v5 \end{bmatrix}$$

First, we observe that the determinant is essentially zero, from which we conclude that the five measured variables cannot be controlled using the five valves. We also observe that the first and second columns can be made identical my multiplying by a constant value of approximately 12. This indicates that valves *v1* and *v2* have the same effects on the measured variables. We will remove one manipulated variable from this analysis and as a result, must reduce the number of controlled variables by one. We note that valves *v1* and *v2* heat the feed, so that they have the same effect of flash temperature (T6), pressure, level and composition. The least important controlled variable is the flash temperature; it does not affect safety and is a "process environment" variable. The resulting gain matrix is non-singular, indicating that the process can be controlled (at least in a small region around the design point).

$$\begin{bmatrix} F1 \\ \\ A1 \\ P1 \\ dL_1/dt \end{bmatrix} = \begin{bmatrix} 0 & 2.0 & 0 & 0 \\ \\ -.11 & -.44 & 0 & .043 \\ 6.80 & 1.39 & 0 & -5.36 \\ -.136 & .31 & -.179 & -.0265 \end{bmatrix} \begin{bmatrix} \\ v2 \\ v3 \\ v4 \\ v5 \end{bmatrix}$$

Next, we determine if the process has the capacity to respond to defined changes in disturbance variables while maintaining safe operation and achieving the desired liquid product composition. For this small process, we can sketch the operating window in Figure 6.58. We see that the process has the capacity required for large changes in feed temperature and flow rate.

Since the process equipment is satisfactory, we move to the loop pairing. We want a design that functions properly when one controller is placed in manual or equivalently, when one valve saturates. Therefore, we evaluate the relative gain array below.

$$\begin{array}{c|cccc} & v2 & v3 & v4 & v5 \\ \hline F1 & 0 & 1 & 0 & 0 \\ A1 & 1.83 & 0 & 0 & -.83 \\ P1 & -.83 & 0 & 0 & 1.83 \\ dL1/dt & 0 & 0 & 1 & 0 \end{array}$$

The relative gain analysis indicates that there is only one loop pairing design for this process with positive relative gains for all control loops! Next, we quickly evaluate the dynamics for the loops indicated by the relative gain analysis. We see that the dynamics are favorable, from which we conclude that the valve that has the "strongest" effect on the variable also has fast dynamics. Therefore, we accept the design.

We still have some important variables to integrate into the design. We recognize that the flash temperature is important. In a previous section, we determined that the flash

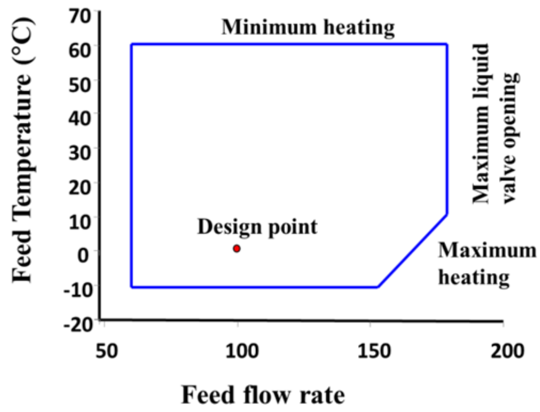


Figure 6.58 Operating window for the flash process. Marlin (2000)

temperature is a good inferential variable for the liquid composition. Therefore, we include T6 as a secondary variable in the analyzer feedback control loop. (This does not violate the controllability analysis because there is no specific value for this temperature; its set point is adjusted by the analyzer controller.)

In addition, we have an additional valve, *v1*. The valves *v1* and *v2* both heat the feed, but the costs of the two are different. The process heat integration is much less costly than the steam heat that requires fuel combustion. However, we need both exchangers to provide the capacity for changes in operating conditions. Therefore, we introduce split range control for the flash temperature that maximizes the use of preheat before opening the steam valve.

The completed design is shown in Figure 6.59, which includes additional equipment for safety. An example of a dynamic response to a feed composition disturbance is shown in Figure 6.60 that shows the successful regulation of all variables.

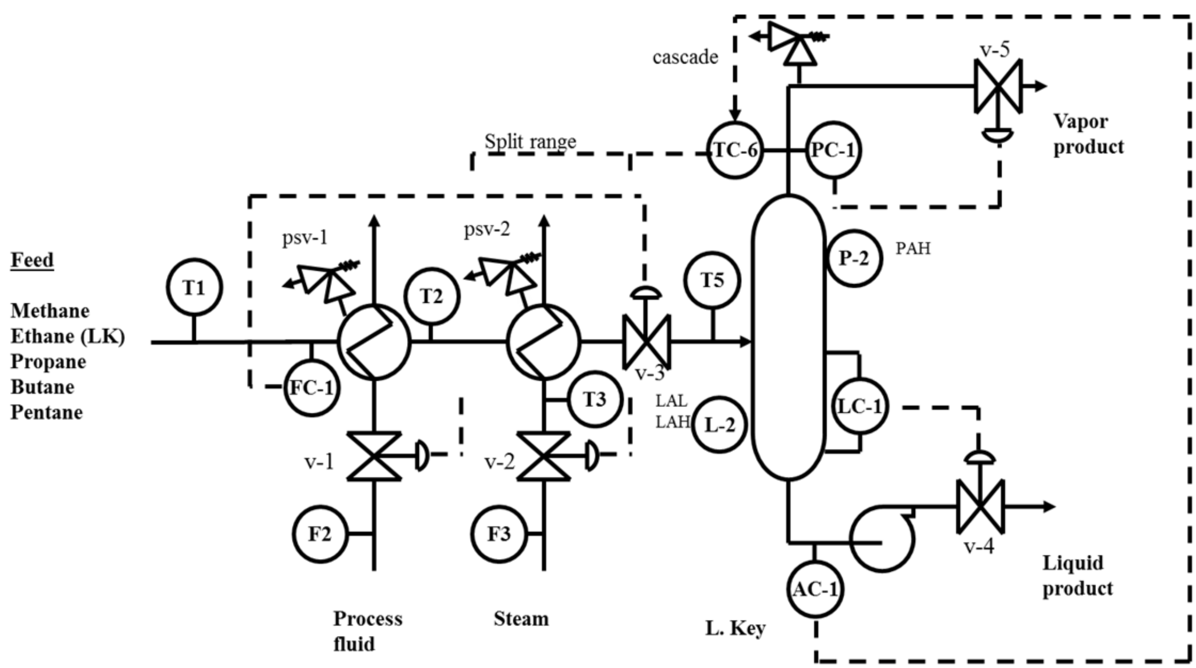


Figure 6.59 Control design for the flash process. Marlin (2000)

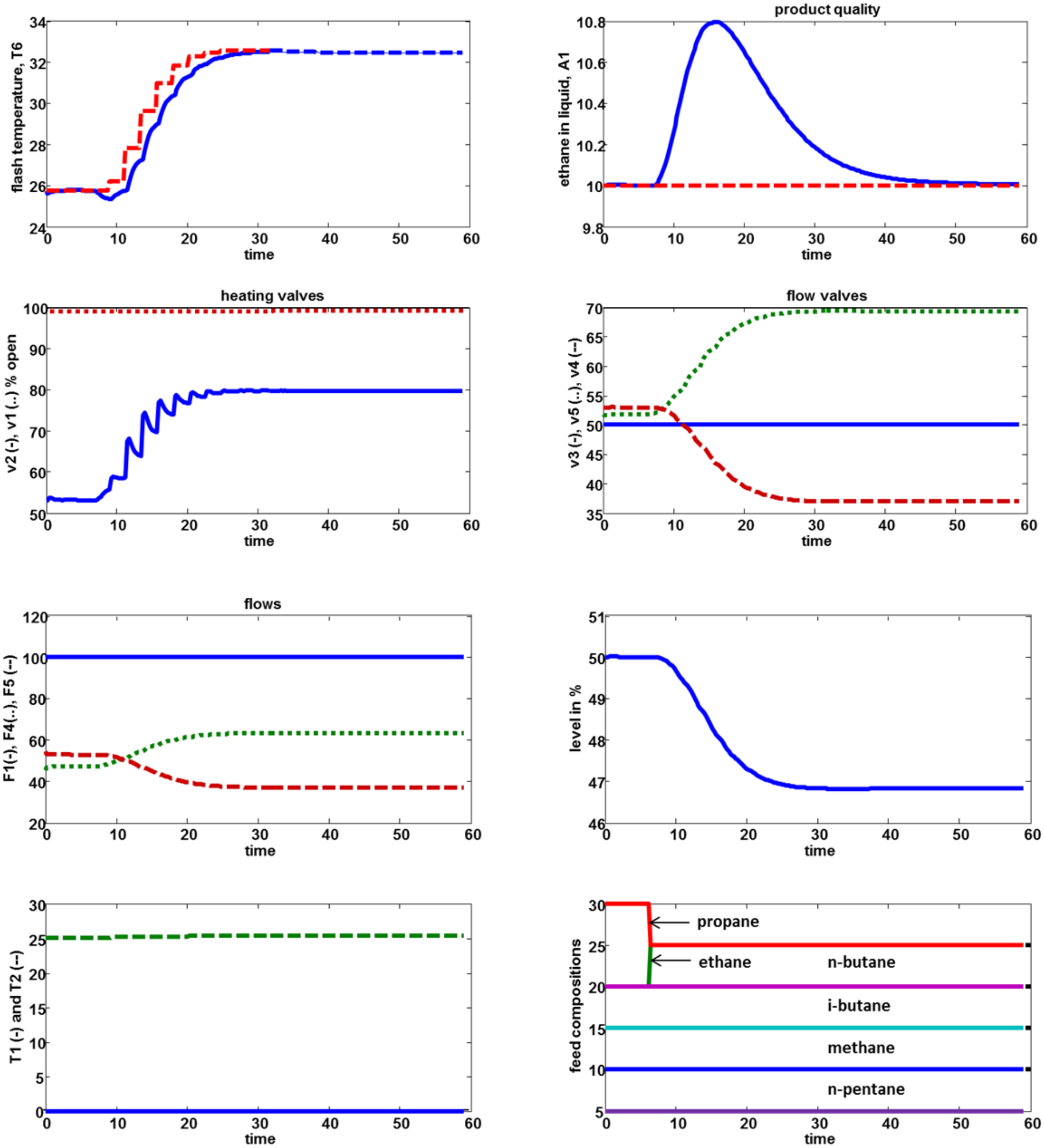


Figure 6.60 Dynamic response of the flash control system to a step disturbance in the feed composition. (Time in minutes, temperatures in degrees Celsius and compositions in mole percent.) Marlin (2000)

6.5.4 Unique features of plant-wide control

The definitions of plant-wide control problems often contain common features. Here, we discuss a few of the common features and control designs to achieve good performance. We do not introduce new basic control methods; we tailor the technology to achieve the demands of the process dynamics, safety, product quality, production rate, and profitability.

6.5.4.1 Production rate and inventory control

One important goal of any process operation is to ensure that material entering equals the material leaving, corrected as needed for some accumulation or depletion consistent with the size of inventory storage. The goal of production rate control is to ensure that the major flow variables match the production rate without human intervention. Several general issues that have to be considered for production rate control are discussed in the following.

- **Inventory control** – Most process plants contain inventories, so that materials flow from one intermediate inventory to the next. Therefore, production rate control cannot be achieved by a single flow controller; a series of flow and inventory controllers are required because setting one flow into or from an inventory does not ensure that the other (from or into) flow will match the desired production rate. For example, most liquid levels are pumped as shown in Figure 6.61. Changing the inlet flow rate to the vessel does not affect the flow rate exiting the vessel. The system is unstable, and feedback control of the level is required to ensure that the flows in and out are equal at steady state.

  In the other common example, the inventory of gas in a vessel is measured by the pressure as shown in Figure 6.62. A change in the flow in affects the vessel pressure that subsequently affects the flow out, so that the pressure is a stable variable without control. While this system would theoretically attain a steady state with the flows in and out matching, the large changes in vessel pressure are usually not acceptable due to the high cost for a high-pressure vessel. Therefore, pressure is controlled to maintain a safe condition in the vessel as well as to match flows in and out.

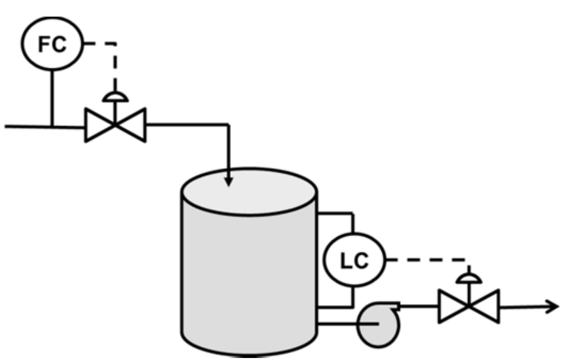


Figure 6.61 Liquid level control with pumped out flow and control

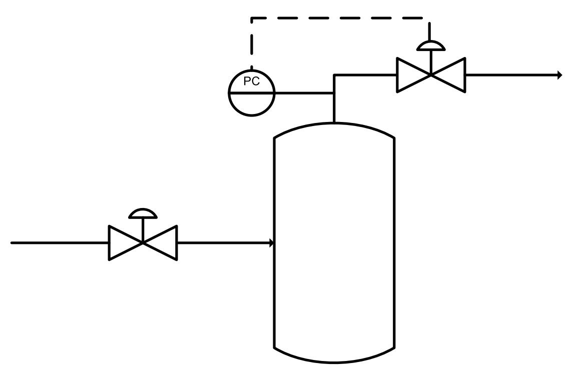


Figure 6.62 Gas inventory pressure control

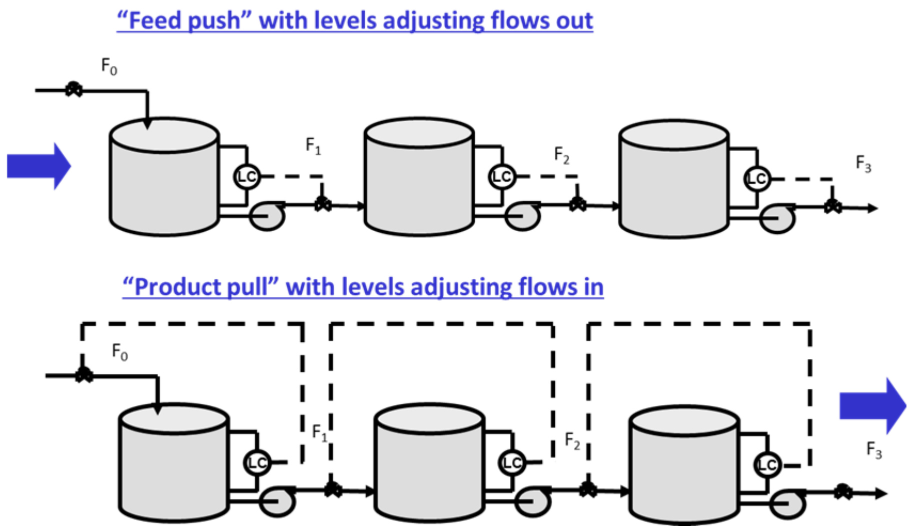


Figure 6.63 Production rate and level control.

- **Location of production rate controller** – In a series of flows and inventories, only one flow can be set independently by operating personnel; all other flows are determined by the inventory controllers so that material balances are achieved. Therefore, the engineer has to select the best location for the production flow controller. The most obvious choice is the feed of raw materials into the first unit as shown in Figure 6.63 referred to as a "feed push" design. (In the figure, each liquid level represents what might be a more complex process with inventory.) While this design results in little variability in the first unit, it allows more variability in later units and cannot rapidly achieve a specified product flow rate. A second choice could be the product flow rate as shown in Figure 6.64; this design provides tight product rate control but allows variability in upstream flows. This is often referred to as a "product pull" design and is used where the process must rapidly satisfy a variable demand. Other choices are possible within the process as shown in Figure 6.64. The unit with feed flow control is usually selected to be the most sensitive to flow variation, so that very low variability flow rate yields the best performance, such as high reactor yields, highest production rate or profit. This is often referred to as "bottleneck control".

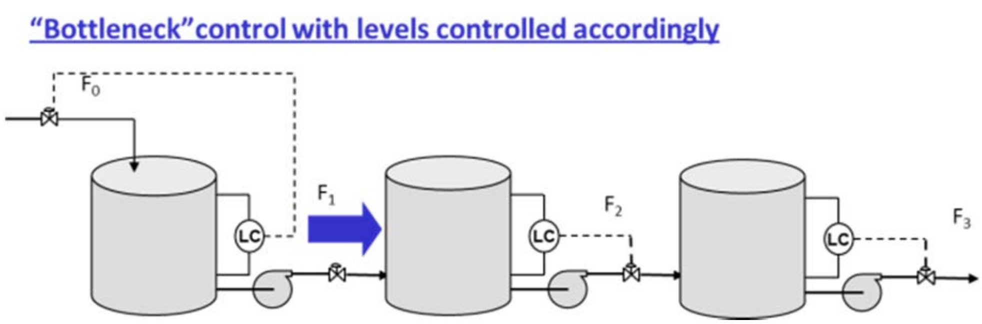


Figure 6.64 "Bottleneck" production rate and level control

- **Multiple flows** – Often, a plant processes several materials that must be supplied concurrently. Usually, these materials are provided in a specific ratio, and the ratio is adjusted based on feedback from process behavior, such as product yields or purity. A flow ratio design can achieve the desired flow rates with only one flow being independent and all others being supplied according to specific ratios.

  A caution on ratio control is noteworthy. Ratio control should not be applied to the output flows from reaction or separation processes. An example is shown in Figure 6.65. Controlling both product flows from a distillation tower is generally not acceptable, because the ratio of flows depends on the feed composition (and reflux ratio); this design will likely result in inventories completely draining or filling. At least one of the product flows for a distillation tower should be determined by an inventory controller.

- **Discontinuous units** – Many process plants consist of some continuous processes and some discontinuous processes. For example, the discontinuous process could be a chemical reactor, and the continuous process could be a separation sequence after the reactor. Plants are often designed as shown in Figure 6.66, with multiple discontinuous processes in parallel and some inventory between the discontinuous and continuous sections of the plant. The plant is operated so that one discontinuous process completes a batch and transfers its material to the inventory, which must have capacity to store the complete batch quantity. The flow rate to the continuous process is set to process, on average, the entire batch by the time that the next batch will be completed and transferred to storage. In this case, the inventory is *not controlled automatically*.

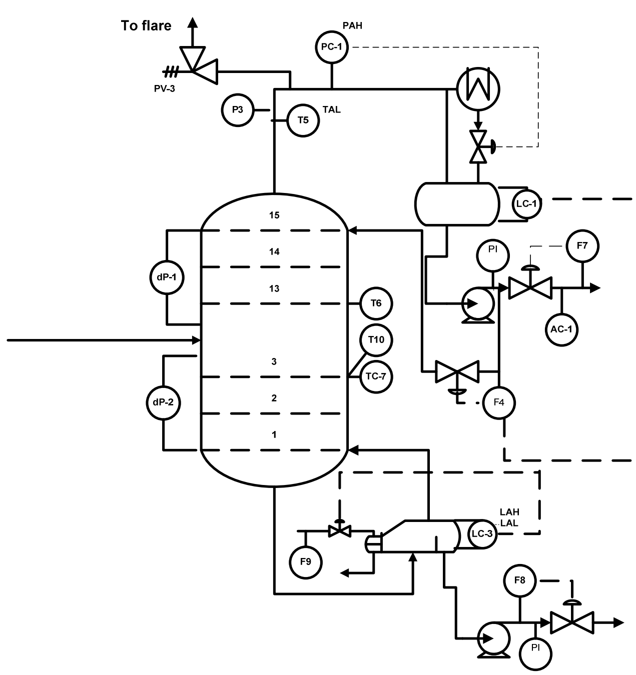


Figure 6.65 Distillation with both products determined by flow control (Not recommended)

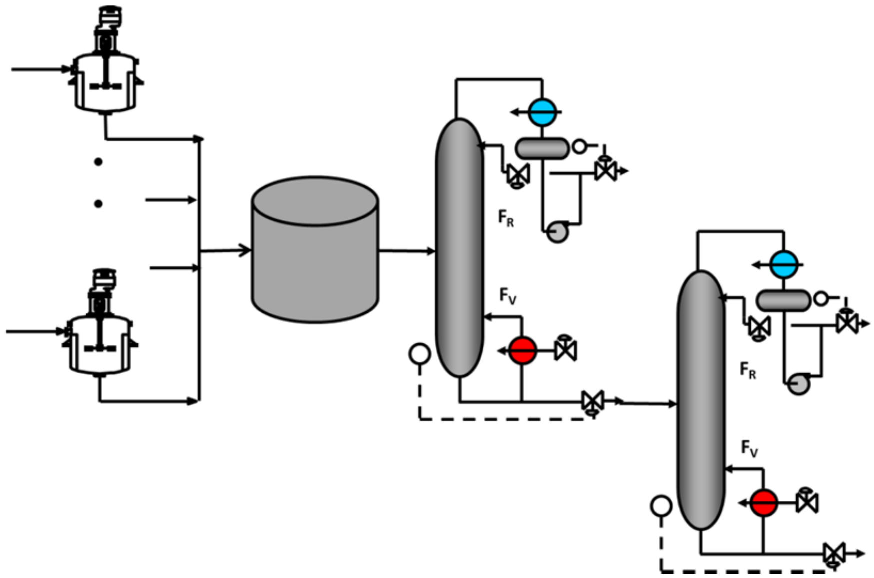


Figure 6.66 Batch reactors feeding continuous processes with intermediate inventory.

- **Raw materials and finished product inventories** - Whenever inventories are expected to vary for good plant operation, the inventories are not controlled automatically, and people monitor the inventories and redirect flows as required. Naturally, sensors and alarms are provided to aid the operations personnel. A typical situation involves raw material and finished product inventory. These inventories are very large, containing material for many days of plant operation when transfers to and from the plant occur infrequently.

6.5.4.2 Utilities control

An important, if often overlooked, aspect of plant-wide operations are the utilities generated and consumed in the plant. These utilities include, but are not limited to, steam, cooling water, compressed air, liquid and gas fuels, hydrogen, oxygen, heating or cooling fluids, and electricity. A utility is an entity that is required to produce the products; we exclude raw materials and consider them separately. The plant operation is simpler when the plant purchases the utilities from an outside company; however, external purchasing is often more costly because of the profit gained by the outside company. In addition, some by-products can be most economically used as utility streams. Therefore, a process plant often generates some or all of the utilities it consumes.

The operation of a utility process is different from a conventional process because it is required to satisfy the second-to-second demand of the plant. Some of the requirements include matching the production with consumption, responding very rapidly to changes in demand, operating over a wide range of generation, providing the utility reliably, and generating the utility at low cost. The importance of utilities becomes clear when we recognize that the entire plant depends on immediate supply of each utility to match every individual process unit requirement.

Some key issues in utility design, operation and control are discussed in the following.

- **Balancing production and consumption** – The utility system must match its generation (perhaps, including external purchases) to the total process consumption. Since many process consumers exist and they function independently, this can be challenging. Let's take steam as an example utility. One method for determining the consumption of steam would be to measure the flow rate of steam to every process consumer, heat exchanger, reboiler, steam turbine, and so forth, as shown in Figure 6.67. Then, we could ensure that the generation equaled the calculated consumption. However, this would be a *terrible approach* because errors in flow measurement would always result in an erroneous calculated value of total steam consumption. The pressure would increase or decrease until a major incident occurred in the plant.

  A better method is to rely on the steam system to perform the material balance, as shown in Figure 6.68. Whenever the flows into and out of the steam pipe are unequal, the pressure changes. Therefore, we use the pressure as an indication of how well the generation and consumption match. To automate this concept, we install a pressure controller that adjusts the steam generation. This ensures that the total (integrated) steam consumption and generation are equal and that deviations are corrected quickly, due to the fast response of the pressure in the small pipe volume.

- **Wide range of operation** – Each of the many units in the plant consumes the utility and might be operating at low or high production rates (and utility consumption rates), while some might be shut down. Therefore, the total utility rate can experience wide variations from day-to-day and very wide variations from month-to-month. In additional, process disturbances, including emergency shutdowns, can cause large variation over minutes. The challenge is to design a generating system that can operate over such a wide range *efficiently*. Again, we can consider steam as an example utility that is generated in one or

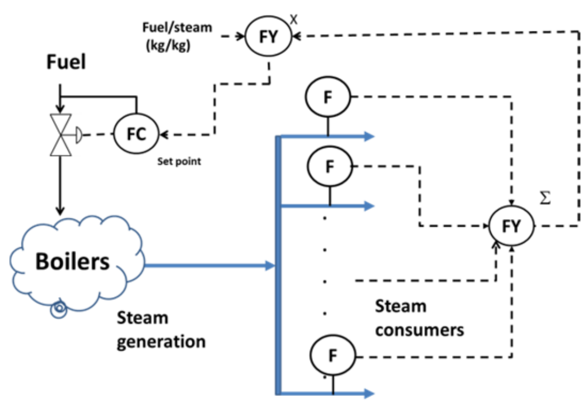


Figure 6.67 Steam balance by equating manipulated to measured flows (not recommended).

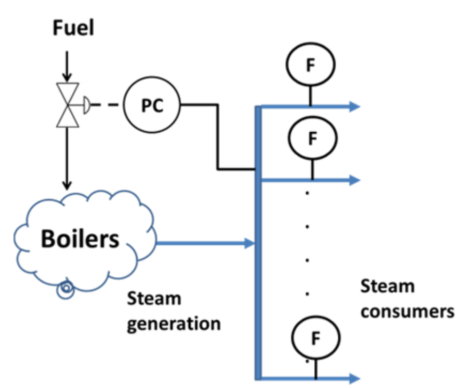


Figure 6.68 Steam balance by pressure control (recommended).

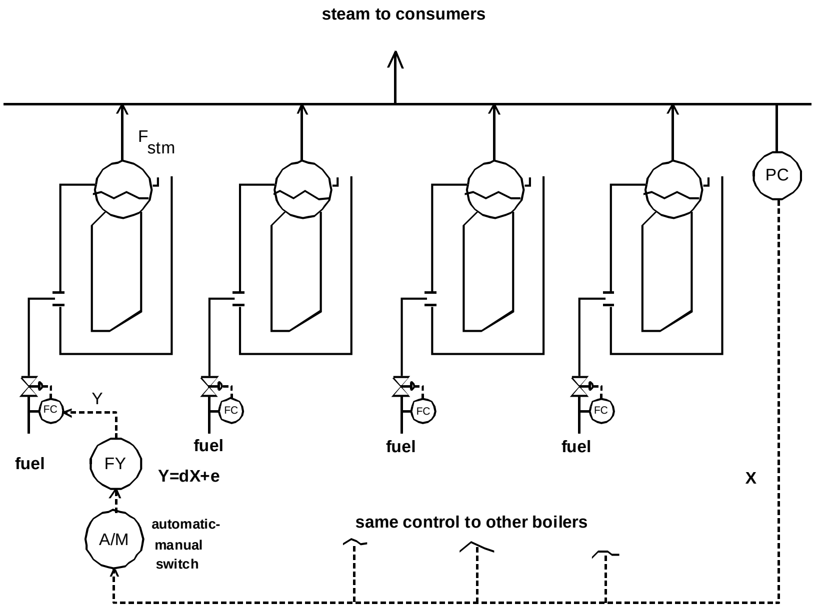


Figure 6.69 Typical mutli-boiler steam header control. The ratio (d) and bias (e) parameters can be adjusted independently for each boiler to achieve the highest efficiency. Marlin (2000)

more boilers heated by fuel combustion. A boiler can operate over a relatively wide range, perhaps 25-100% of its maximum capacity. However, a boiler's efficiency (heat to boil steam/total heat of combustion) is low at lower generation rates. Therefore, a typical design has several boilers that can be placed in operation when needed. As a result, a wide range of operation and a reasonable efficiency can be achieved concurrently. A typical control design is given in Figure 6.69.

- **Reliable utility generation** – The entire plant depends on utilities, so that reliability is essential. For steam generation, a boiler is a complex unit operating near material and combustion safety limits. Therefore, it is not unusual for a boiler safety and protection system to detect an incipient fault (e.g., potential loss of water circulation or loss of sufficient airflow) and shutdown the boiler one or more times per year. This situation would be disastrous if only one boiler were included in the design. In addition, most utility generators require considerable time for startup; for example, a boiler start from ambient temperature might require many hours. Therefore, more boilers are typically in operation than needed to satisfy the total demand, so that upon a single boiler failure the remaining boilers can satisfy the total plant steam demand. In the event of two boilers failing, the plant will have a "load shedding" plan that will stop some less important consumers and switch others to an alternative power source such as electricity. The load shedding is devised to maintain units in operation whose shutdown costs the most and requires the greatest time for restarting.

  In some instances, alternative sources of the utility increases the reliability of the utility system. For example, alternative sources of pipeline natural gas provide a backup for a utility that is designed to consume by-product methane and ethane in combustion systems. The backup sources are more expensive (if they were not, they would be the primary sources). Therefore, the control systems are designed to call on them only when needed. A typical fuel gas distribution for a process plant is shown in Figure 6.70.

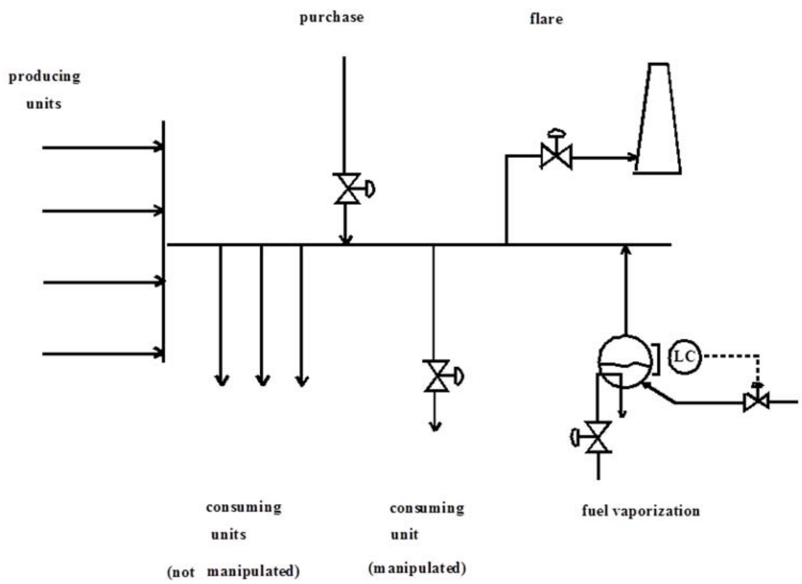


Figure 6.70. Typical fuel gas distribution system in a process plant. Marlin (2000)

Another method for providing reliability in a utility in the plant is to include an inventory that can be used to satisfy demand, at least for a short time until a fault is corrected and utility operation resumes. For many situations, this is not possible, but it is an option for some utilities such as air separation, where storage of liquid oxygen and nitrogen is possible. Naturally, liquefying and subsequently vaporizing the material increases the cost.

- **Rapid response to disturbances** – The feedback control systems for utilities using fast-responding variables can maintain most systems in balance. However, situations arise in which very large step disturbances occur that result in large deviations and potential utility system "collapse", where the generation is so much lower (or greater) than demand that the plant cannot remain in operation. For example, a large steam demand occurs where a unit is started up, and a steam excess occurs when a unit is shutdown. Naturally, the first approach to improve the dynamic behavior will be to slow the disturbance; however, this is not always possible or cost effective. In these situations, the control system must be enhanced with feedforward control.

- **Profitability** – Utilities involve fuel and power consumptions that have a major impact on overall plant profitability. Therefore, these systems need to be operated efficiently. As discussed, these systems typically have numerous parallel methods of generation and sometimes, external sources as well. The control system should be designed to utilize the least costly generation while satisfying the total demand.

In some instances, substantial improvements can be made by scheduling process operations to lower utility costs. For example, periodic operations such a material movements among storage locations that consume electrical power can be performed when the electricity prices are low, usually during the night.

6.5.4.3 Process recycle

Process integration through recycle of material and energy can provide substantial economic advantages. However, it is well recognized that integrated units can be more difficult to operate because a change in one unit propagates a disturbance to other units that ultimately returns to the originating unit. Thus, a poorly designed and operated plant with recycle can experience many disturbances that cycle through integrated units. Fortunately, proper design and control can substantially reduce the negative operability issues without losing the economic advantages of recycle.

The key complicating effect of recycle systems is the feedback occurring as part of the *process*. A process sketch and a block diagram of a simplified reactor process with recycle are shown in Figures 6.71a and 6.71b. The transfer function between a disturbance in the inlet temperature and the reactor outlet temperature is given below.

$$\frac{T4(s)}{T0(s)} = \frac{G_R(s)G_{H1}(s)}{1 - G_R(s)G_{H2}(s)}$$

Now, we will give specific transfer functions, which could be for an exothermic chemical reaction ($K_R = 3 > 1$).

Process without recycle	Process with recycle
$G_{H1}(s) = 0.40 \quad K/K$	$G_{H1}(s) = 0.40 \quad K/K$
$G_{H2}(s) = 0.0 \quad K/K$	$G_{H2}(s) = 0.30 \quad K/K$
$G_R(s) = 3/(10s+1)$	$G_R(s) = 3/(10s+1)$

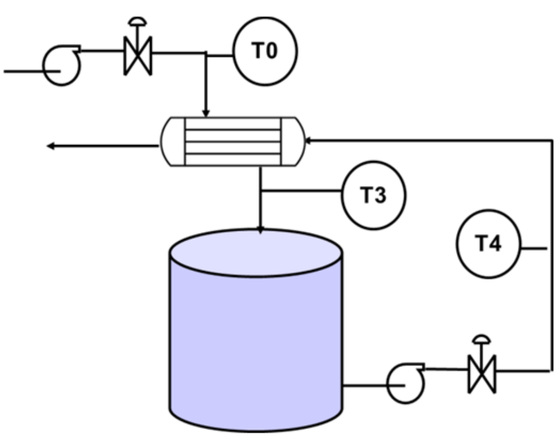


Figure 6.71a. Typical process with recycle.

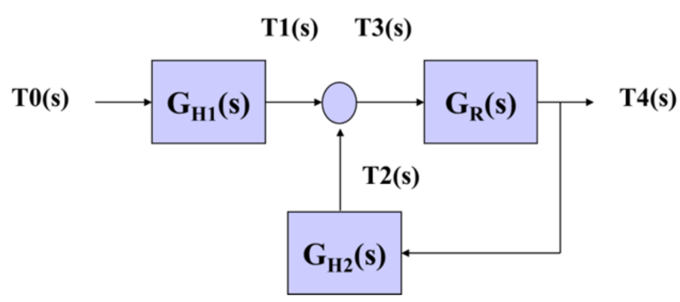


Figure 6.71b. Block diagram of the typical recycle process in Figure 6.71a

The question we first investigate is, “How does the recycle change the effect of a disturbance in T0?” The responses of two designs are given Figure 6.72 for a step change in the inlet temperature, T0. In the top figure, the response for a process without recycle is plotted, and in the bottom figure, the response for the recycle system is plotted. We note that the effect of the disturbance is much larger and slower in the recycle system. To understand why, we rearrange the overall transfer function to obtain the following.

$$\frac{T4(s)}{T0(s)} = \frac{K_{H1}K_R / (1 - K_{H2}K_R)}{\left(\frac{\tau_R}{1 - K_{H2}K_R}\right) s + 1} = \frac{K'}{\tau' s + 1}$$

We see that the effective gain and time constant are influenced by recycle, and as the recycle gain becomes larger (and $K_{H2}K_R$ approaches 1.0), the disturbance effect becomes larger. This is process feedback, but unlike the feedback provided by process control, this is *bad feedback*. When we look at the definition of feedback, we recognize the process feedback in this example as *positive feedback*, which tends to drive the output variable away from the desired value.

Process recycle systems almost always introduce positive feedback that magnifies the effects of disturbances and lengthens time to steady state.

Recycle is introduced into process designs to realize desirable efficiency effects, but it also introduces positive feedback in the process. The engineer should include control designs that provide the good efficiency while ensuring that the process has good dynamic performance A few approaches for the design and control of recycle systems are discussed in the following.

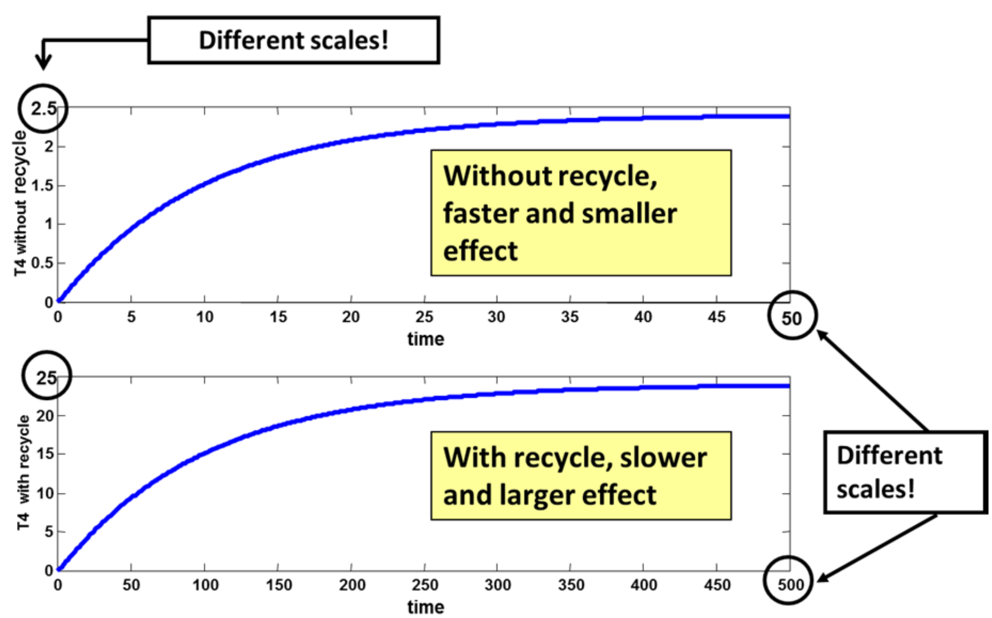


Figure 6.72 Dynamic response of two designs, with and without recycle. Marlin (2000)

- **Break the positive feedback** – A direct method for reducing the positive feedback from recycle is to adjust a compensating source or sink of the recycled entity (material or energy). As an example, let's consider the reactor with feed-effluent heat exchange in Figure 6.71a. The positive feedback results from the reactor effluent temperature affecting the reactor inlet temperature. This can be prevented (perfectly in steady state and with some deviation dynamically) by introducing an additional heat exchanger as shown in Figure 6.73, which would increase the capital cost. The heat exchanger heats the reactor feed to a fixed temperature. In this case, the heat recovered by the feed-effluent exchanger must be reduced and a heating fluid must be provided, which could result in higher operating costs.

  Applying this principle requires that some recycled entity not be recovered; however, most of the recycle is returned to the process. This type of design is very effective as long as the process and control dynamic response for the unit breaking the positive feedback is fast with respect to the other elements in the recycle system.

- **Smooth dynamics through inventory control** – In some processes, there is a long time between a change in the input to the process and the response of the recycle. For the example in Figure 6.74, a solvent is added to a reactor and is recovered, purified, and returned to the reactor. The recycle loop is lengthy, with delay between an increase in the solvent flow to the reactor and the solvent return. Therefore, the design in Figure 6.74 includes an inventory for storing solvent, which ensures that the reactor can be provided the appropriate amount without delay. The direction of level control involves manipulating the outflow (push system) for flows heading towards the intermediate storage tank and manipulating the inflow (pull system) for flows away from the intermediate storage tank. We note that the storage inventory is not controlled; it rises and falls to account for temporary changes in solvent processing rates. Level control in recycle has been discussed by Buckley (1974).

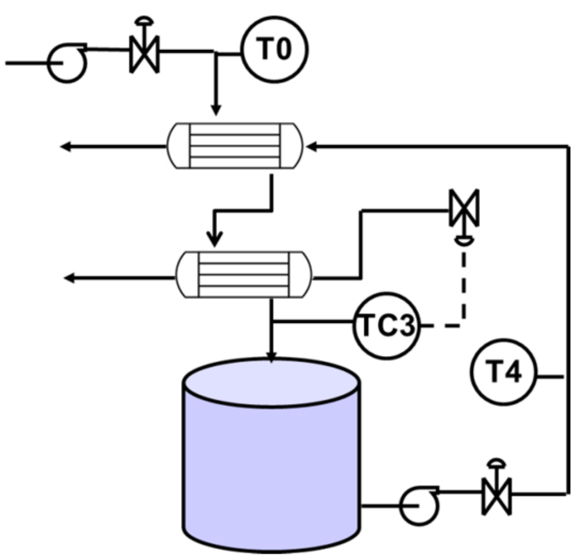


Figure 6.73 Recycle process with reactor inlet temperature control.

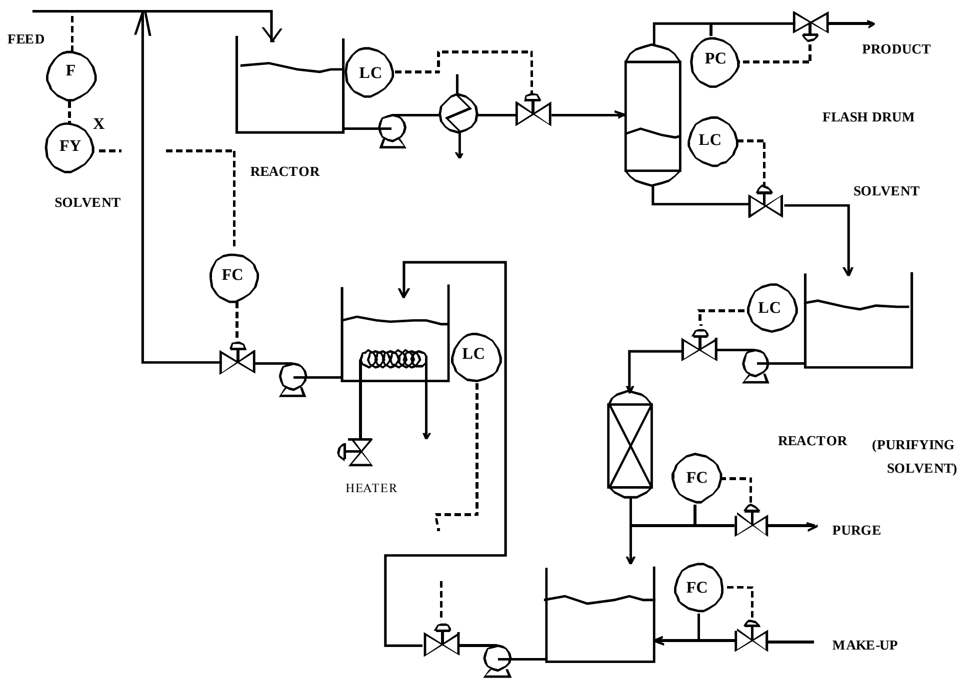


Figure 6.74 Solvent recycle process with uncontrolled recycle inventory. Marlin (2000)

- **Prevent component accumulation** – Material recycle systems provide the opportunity for a component to accumulate in the process. In an extreme example, a component might not exit the process in any product stream. In such a case, the component would accumulate in an unlimited manner, and upset the process operation. The classical solution is to introduce a purge stream from the process in a location where the accumulating material is at a high concentration.

  The purge will contain some of the accumulating material and will enable steady state to be achieved. Note that many other components exit in the purge as well, and if these are valuable, the purge stream should be processed in a separation unit and the valuable components returned.

  Component buildup is also possible in recycle systems with a chemical reactor in the loop. This topic is addressed in detail in Luyben et. al. (1999).

A.5.4.4 Partial control

Partial control is a concept closely akin to inferential control. Inferential control is generally confined to product quality in a single unit, so that all measured and manipulated variables are located in the unit. In contrast, partial control addresses a wider range of objectives, such as profitability. In addition, partial control may require adjustment of variables in integrated units. (As an aside, this general approach has also been termed “self-optimizing control.”)

Typically, the goal of partial control is profitability of the entire plant, such as the expression given below.

Profit = product sales revenues – raw material costs – fuel costs – power purchase
– changes in inventory (raw material, products, work in progress)
- catalyst and chemicals – cost of utilities

We note that important costs that are not affected by the control system, like capital, personnel, laboratory, and so forth are not included in the measure of "operating profit". In special circumstances, profit can be more concisely expressed as maximum selectivity or minimum energy or another simple goal, but the simplification should be investigated thoroughly before being accepted.

When designing partial control strategies, the engineer must determine surrogate variable(s) that, when controlled in feedback to a constant set point(s), result in process performance that closely approximates the desired goal, i.e., optimum profit. The surrogate variables are often termed "dominant variables".

Clearly, the advantage of controlling the surrogate dominant variables is much faster and simpler feedback control. An inherent disadvantage is the approximate relationship of dominant variables to the true profit. This disadvantage could be overcome by a higher-level profit controller that uses models and real-time data to reset the values of the dominant variables in a cascade design.

The design of the dominant variable(s) is a task for the engineer. To begin with, the dominant variable(s) must be controllable, as discussed previously in this appendix. In addition, they should provide fast feedback information about process performance, to ensure fast feedback control response. Finally, the variable(s) should be highly correlated with the ultimate control goal, such as product quality, profit, etc.

Controlling dominant variables at their set points will maintain the process in conditions yielding profit close to the maximum as disturbances occur.

Engineers can use two methods to determine an appropriate selection of dominant variables for partial control design. One method uses fundamental models, while the other uses historical data. The reader is referred to Luyben et.al. (1999) and Skogestad (2004) for details on partial control design. Some typical dominant control variables include the following.

- Ratio control of components to blending and reactors
- Ratio of utilities to feed rate to separation units
- Concentration of a limiting component in a chemical reactor

6.5.5 Final step - safety review

The final stage of the control design is a safety review. Remember that an improperly designed control system can lead to hazards and that even a properly designed control system introduces new faults that can lead to hazards. Therefore, a process safety analysis is performed after the process control system has been designed and shown on the P&I Drawings. Generally, a hazard and operability study (HAZOP) would be performed followed by layer of protection analysis (LOPA), if necessary. Also, changes to the process control design in an operating plant might have unforeseen consequences that compromise safety. Therefore, a management of change program must be performed for changes proposed after the initial plant-wide process safety analysis.

6.6. Unit operation control

Most process plants are composed of a complex interconnection of standard unit operations. Here, control of several of the more common unit operations is discussed. The emphasis is on the integration of process and control principles to select the best manner for control. In addition, many options exist in the design of the unit operation, and the impact of changes in process design on the control performance is addressed.

We observe that many factors influence control performance. In designing a control system, we desire a process design and control system that has

- **Range**: A wide range of operation
- **Gain**: A large steady-state gain, i.e., a strong effect
- **Dynamics**: A fast response
- **Linearity**: A nearly linear relationship between manipulated and controlled variable, so that a controller functions well with constant values for tuning parameters

6.6.1 Flow control

The simplest control loops with one measured, controlled variable and one manipulated variable. In this section, we will introduce issues and designs that are often misunderstood and will see correct designs.

Examples 6.18 to 6.22 address one of the simplest topics in design, regulation of flow in a closed conduit. Even this simple control system requires knowledge of the equipment performance. From many designs that could provide good flow control, one design often provides the highest energy efficiency. While learning these designs is useful, the more important lesson is that control design requires in-depth understanding of the process goals and equipment performance.

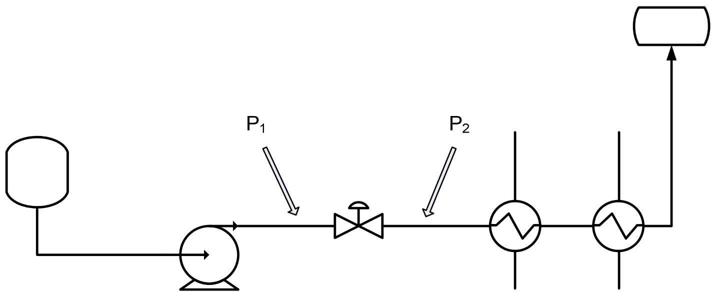


Figure 6.75 Flow process with centrifugal pump and variable resistance by valve.

Example 6.18. **Loop linearity using valve characteristic**: The most common method of flow control involves placing a variable resistance in a closed conduit. A typical system is shown in Figure 6.75 that includes a centrifugal pump and an automatic control valve. A proportional-integral controller can be used for feedback control.

- ***Range: A wide range of operation can be achieved, from zero flow to the maximum with the valve fully opened.***
- ***Gain: If a substantial percentage of the pressure drop occurs across the valve, the gain will be large.***
- ***Dynamics: The dynamics of flow measurement and valve actuation are fast, and the flow process is very fast. (How long after you open a faucet does it take the flow to increase?)***
- ***Linearity: A nearly linear relationship between manipulated and controlled variable is desired. The following analysis must be performed for each flow system, and a compensating non-linearity introduced into the control loop where needed.***

Next, we determine the relationship between the valve position and the flow rate, which can be derived by applying the Bernoulli equation to the flow system and ignoring small friction losses.

$$F = F_{\max} \frac{C_v(v)}{100} \sqrt{\frac{\Delta P_v}{\rho_{ave}}}$$

with

$C_v(v)$	***the inherent valve characteristic (v in percent open)***
F	***the volumetric flow rate***
F_{max}	***the maximum flow rate at 100% valve opening***
ΔP_v	***the pressure drop across the valve***
v	***is the valve opening in %***
ρ_{ave}	***is the average density of the fluid***

Note that inherent characteristic is the relationship between the valve stem position (v), which is adjusted by the controller, and the flow rate through the valve at <u>constant</u> pressure drop (ΔP_v) across the valve. Control valves are manufactured with various inherent characteristics, and a few common examples are given in Figure 6.76. The manufacturers achieve various functions for inherent characteristic by modifying the relationship between the stem position and the opening between the valve plug and seat. Since the manufacturer cannot know whether or how much the pressure drop

changes with flow, it is the task of the design engineer to select the proper characteristic for a particular flow system.

The key pressures in the typical flow system in Figure 6.75 are shown in Figure 6.77 as a function of the flow. The top curve is a typical head curve for a centrifugal pump; the design engineer can acquire the curve matching the installed pump from the pump manufacturer. The bottom curve is the "system curve" and can be determined by calculating the pressure drops as a function of flow using friction factor correlations.

We note that for the example process in Figure 6.75 the pressure drop across the valve decreases with increasing flow rate; see Figure 6.77. For good control performance, we desire a constant process gain between the adjusted control valve stem position (v) and the controlled variable (F). We can achieve a (approximately) constant gain by selecting a function for the valve characteristic (C_v) that results in the same change in flow rate for each one percent change in stem position, independent of the starting value of the stem position.

$$F = F_{\max} \frac{C_v(v)}{100} \sqrt{\frac{\Delta P_v}{\rho_{ave}}} \approx \beta\, v \qquad \text{with } \beta \text{ a constant}$$

The desired function is selected from standard characteristics provided by valve manufacturers. For this example, the proper characteristic is equal percentage, which yields a nearly linear process gain between the valve position and the flow rate.

The characteristic for every control valve must be determined considering the relationship between the adjusted valve stem and the measured controlled variable for the specific process and operating conditions.

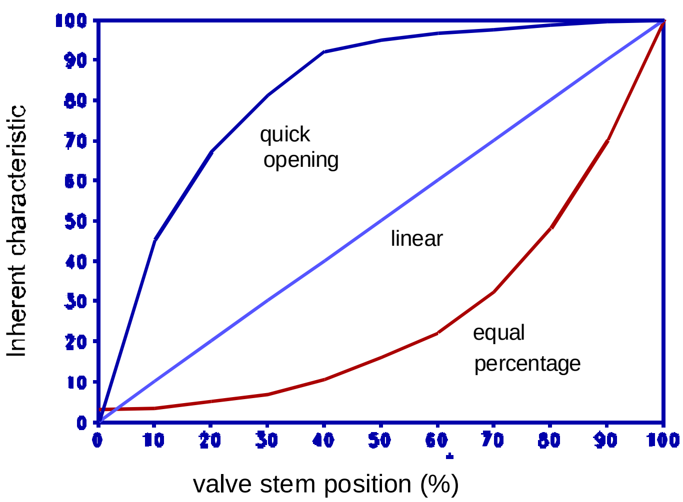


Figure 6.76 Typical (inherent) valve characteristics

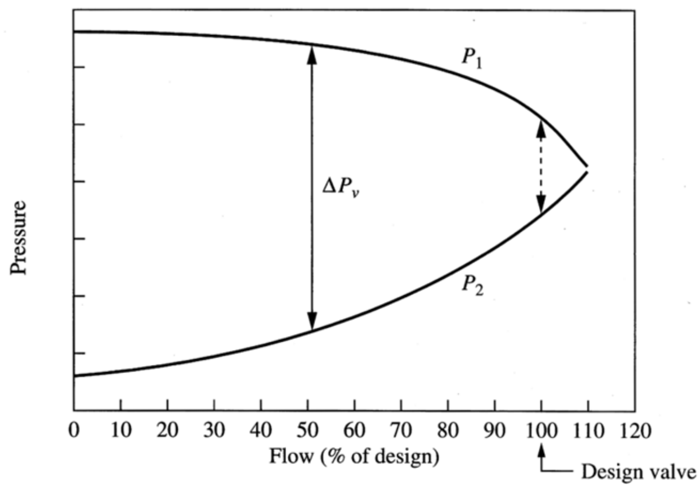


Figure 6.77 Valve pressure drop for the process in Figure 6.75 Marlin (2000)

The relationship between stem position and controlled variable will differ from loop to loop, so the engineer needs to evaluate each process to determine the appropriate linearizing characteristic. Also, note that the controlled variable does not have to be flow rate; it could be temperature, pressure or other measured process variable.

Example 6.19. Flow control. Different flow control designs are shown in Figure 6.78. The sensor is an orifice meter; the material is a liquid at its bubble point; and the goal is to control the fluid flow rate. Determine which, if any, of the designs is acceptable

a. This design has a serious flaw. A valve is located in the pump suction, which introduces a significant pressure drop before the pump. This design will likely lead to cavitation in the pump. Valves should be placed in pump outlets, not in suction lines.

b. This design is likely OK. But, it would be better to locate the orifice sensor before the valve, where the pressure is higher and the likelihood of cavitation in the sensor is lower.

c. This design does not have pump, which is proper if the pressure in the vessel is sufficient for the flow to the downstream unit; we will assume that this is the case. The liquid at the bottom of the drum is at its bubble point, which could vaporize in an orifice meter, due to the pressure decrease where the liquid velocity increases. Therefore, the orifice meter must be located sufficiently below the vessel to provide a head that can prevent cavitation.

d. This design has a serious flaw. The orifice meter is located after the valve, which introduces pressure drop. There is a high likelihood of flashing in the sensor.

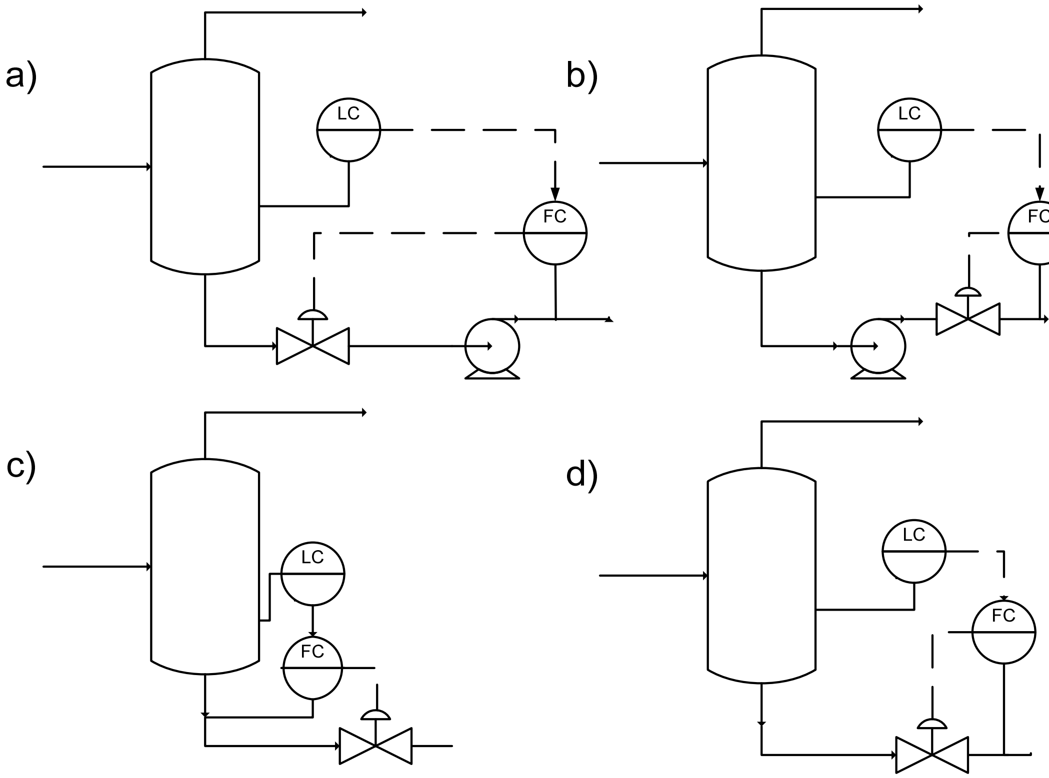


Figure 6.78 Possible flow control designs for the liquid from a flash drum.
(Orifice meters used in all designs)

Example 6.20 Two possible flow control designs are given in Figure 6.79. The fluid is far from its bubble point. Which would you recommend?

The key question for feedback control is, "Does a causal relationship exist between the manipulated and controlled variables?" For both designs, a change in valve opening will affect the measured flow rate. Therefore, either of the designs will function. As stated in the previous example, the flow meter is typically placed at the location of higher pressure to prevent cavitation.

Example 6.21 Energy consumption. The generic flow loop in Figure 6.80 would involve some pressure drop across the sensor and valve. In some processes, especially high pressure gas, the cost of these pressure drops can be substantial. What design steps can be taken to reduce the pressure drops and reduce the cost of energy?

Sensor: The pressure drop is affected by the choice of sensor technology. For example, the following list ranks some common sensors based on their non-recoverable pressure drop.

- ***Orifice*** ***(highest pressure drop)***
- ***Venturi***
- ***Pitot tube*** ***(lowest pressure drop)***

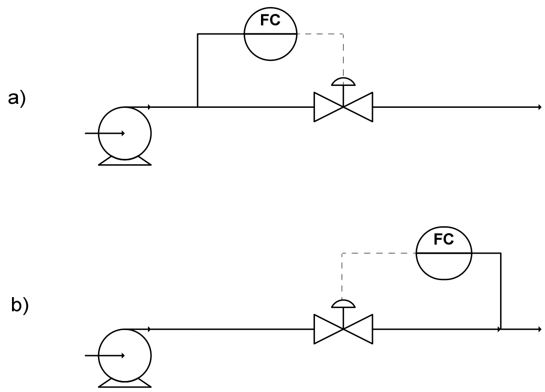


Figure 6.79 Two possible flow control designs.

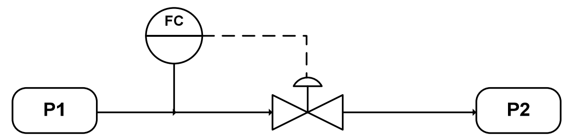


Figure 6.80 Flow control loop.

***<u>Valve:</u>* The choice of valve body affects the pressure drop across the valve. For example, the following list ranks some common valve bodies based on their non-recoverable pressure drops.**

- ***Globe*** ***(higher pressure drop)***
- ***Butterfly***

***<u>Valve sizing</u>*: The "size" of the valve refers to the orifice opening with the stem in the 100% open position. Valve manufacturers document the options for each valve type, and naturally, only a discrete number of options are available. The valve size affects the system (lower) curve in Figure 6.77, with a larger valve size or capacity having a lower pressure drop. With a lower pressure drop, the engineer can select a pump (upper) curve that requires less energy. However, the equipment must be able to process the desired flow rate, which requires some minimum capacity, investment and energy consumption. A general guideline is that the pump-valve-pipe combination should process the design rate at ~70% valve opening. One should use such a guideline with caution – it certainly would not apply in the case of reactor cooling water where perhaps triple the normal flow rate is required when the temperature of the reactor contents is too high and a runaway is possible. Engineers need to understand the range of variation to be achieved by the equipment before designing the equipment!**

***<u>Pump driver</u>*: The pump's speed of rotation of the centrifugal pump affects the top curve in Figure 6.77. A variable speed pump driver can achieve the desired flow rate by changing the speed without a control valve. The key advantage of this design is lower energy consumption. Variable speed drivers can be steam turbines or variable speed electric motors. The choice is between (a) the simpler, lower capital cost, and higher operating cost constant speed pump or (b) the more complex, higher capital cost, and lower operating cost variable speed driver. Economics provides the basis for the choice.**

Example 6.22. Two valves in a pipe: The flow system is Figure 6.81a has two valves in one pipe. The pump is a centrifugal pump with a constant speed driver. What can be controlled in this process?

Let's consider the flow rate first; how many flow rates can be controlled independently? By material balance, we note that the flow rates at all sensors are identical. Therefore, we conclude that no more than one flow rate can be controlled. Since there is a causal relationship between valve opening and the flow for each valve, either valve may be used as a manipulated variable to control flow.

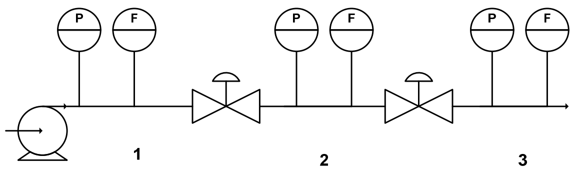


Figure 6.81a. Pipe with two valves.

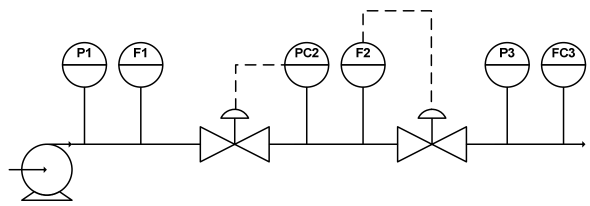


Figure 6.81b. One solution for control using two valves in one pipe.

Now, we will consider the pressures? Let's assume that flow is being controlled and consider three locations, numbered 1 to 3 in Figure 6.81a.

1. ***The pressure at the outlet of the centrifugal pump depends upon (i) the suction pressure and (ii) the pump head, which in turn depends on the flow rate (as defined by the pump performance curve). When the flow is determined, the pressure is dependent. Therefore, it cannot be controlled when the flow is controlled.***
2. ***The pressure between the two valves depends upon the relative amount of pressure drop occurring at each of the valves. Note that the total resistance to flow depends on the flow rate but that the same flow rate can be achieved by different combinations of the two valve openings. Therefore, the flow rate and intermediate pressure can be controlled independently using the two control valves.***
3. ***The pressure downstream of the two valves depends on the downstream process and cannot be controlled by adjusting the two valves available in this example.***

One solution for this problem is shown in Figure 6.81b. This design achieves a pressure reduction or "letdown" along with flow control. The pressure letdown could be desired when a high-pressure source is used in a process with lower pressure equipment. One common example is natural gas distribution in a city. The distribution network would be at high pressure because of the long distances, but local gas users would not like to invest in expensive equipment required by high pressures. Therefore, pressure letdown allows local users to have equipment appropriate for lower-pressure operation.

6.6.2 Heat exchanger control

Nearly every process plant involves heating and cooling, and many unit operations require heat transfer, e.g., distillation reboilers and condensers. Therefore, we begin the coverage of unit operations with some common heat exchangers. Let us begin with a counter current shell and tube heat exchanger shown in Figure 6.82. A simplified model of the exchanger is given in the following.

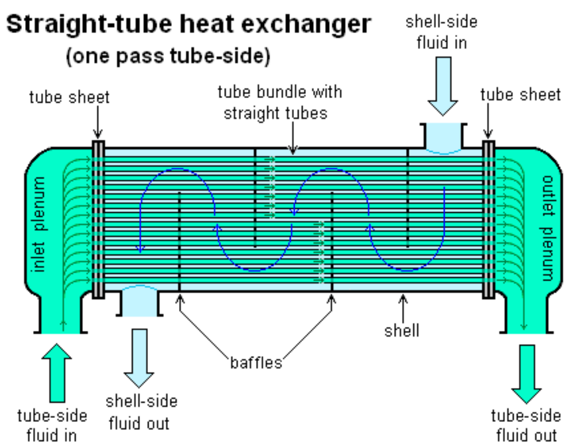


Figure 6.82. Schematic of typical shell and tube countercurrent heat exchanger. (Padleckas, 2012)

$$
\begin{aligned}
Q &= F_{Hot} C_{p_H} (T_{Hout} - T_{Hin}) \\
Q &= F_{Cold} C_{p_C} (T_{Cout} - T_{Cin}) \\
Q &= UAY(\Delta T)_{lm} \\
U &= \frac{1}{1/h_i + x/k + 1/h_F + 1/h_0} \\
h_i &= \alpha_i F_i^{\beta_i} \\
h_o &= \alpha_o F_o^{\beta_o}
\end{aligned} \tag{6.8}
$$

where

Q = heat transferred
F = flow rate
C_p = heat capacity
U = overall heat transfer coefficient
h = film heat transfer coefficient
Y = correction factor for non-exact counter current flow patterns
T = temperature
k = thermal conductivity
x = wall thickness
α, β = empirically determined coefficients
Subscripts F, i and o represent fouling, inner and outer, respectively

Example 6.23. **Basic heat exchanger control** - Let's see how some designs stack up with the desired behavior. We begin with a heat exchanger without phase change. The first design we will consider is the most obvious, in which one stream is adjusted to control the temperature of the other stream, as shown in Figure 6.83.

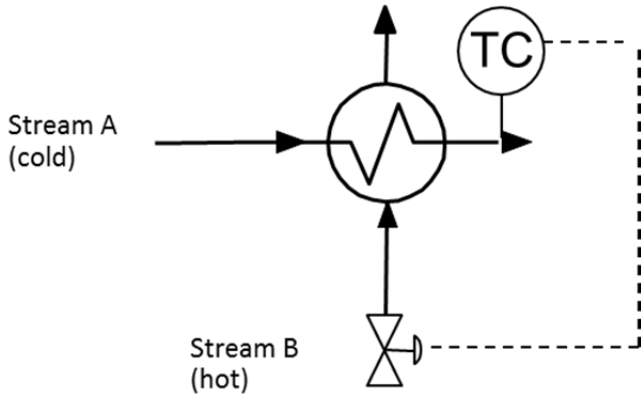


Figure 6.83. Heat exchanger with one flow rate manipulated to control the other stream's outlet temperature.

The design is evaluated in the following.

- ***Range: The design has a wide range, as the manipulated flow can be adjusted from zero (no heat transfer) to its maximum (maximum heat transfer)***
- ***Gain: The gain changes dramatically over the range of manipulated flow. At low flows, the gain ($\Delta T/\Delta F$ at steady state) is very large, but at high flows, the gain becomes very small.***
- ***Dynamics: The dynamics are reasonably fast, but the time to steady state involves heating or cooling all liquid and metal of the heat exchanger.***
- ***Linearity: As discussed under the gain, the process gain will be highly non-linear. This will require some compensation in other elements to achieve a nearly linear control loop. Options exist for non-linear compensation in the final element (control valve characteristic) or in the controller gain (K_C), which can be modified through a more complex algorithm.***
- ***Other considerations: The fouling rate is usually higher at low flow rates leading to high exit temperatures, especially when the adjusted stream is cooling water. Therefore, low flows through a heat exchanger should be avoided.***

Generally, the design in Figure 6.83 is used when tight control is not required and fouling is not an issue because of the materials and operating conditions.

Example 6.24. **Faster Heat exchanger control** - Now, let's consider a slightly more complex process in Figure 6.84 with an adjustable by-pass around the exchanger. The control system adjusts the ratio of fluid through the exchanger and through the by-pass to influence the heat transfer.

This design, which is used extensively in practice, is evaluated in the following.

- ***Range: The design has a wide range, as the manipulated flow can be adjusted from zero by-pass (maximum heat transfer) to all by-pass (no heat transfer)***
- ***Gain: The process is basically a mixing process. Therefore, the gain effect on the temperature is strong.***

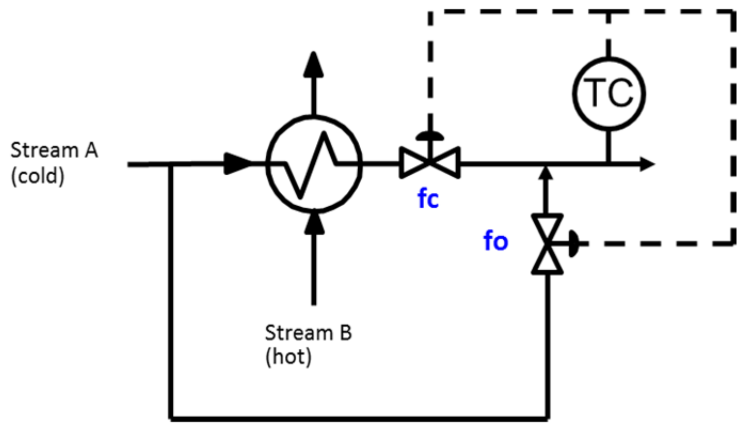


Figure 6.84. Heat exchanger bypass manipulated to control the stream's outlet temperature.

- ***Dynamics:** The dynamics of the mixing process is very fast, much faster than the heat exchanger itself.*
- ***Linearity:** The linearity depends upon the relationship between the valve adjustment and the flow. This will not be exactly linear, but it will be closer to linear than the previously considered design.*
- ***Other considerations:** Limitations can be imposed to prevent a low flow through the exchanger. The by-pass adjustment can be effected by either (i) one three-way valve or (ii) two valves with opposite failure positions. The three-way valve is less expensive and does not ensure tight closure. The two-valve design is used for large flows (pipes) and where (nearly) complete shut off is desired.*

Example 6.25 Manipulate the heat exchange area - Now, let's consider a heat exchanger with phase change where a liquid boils on the shell side. This example will enable us to evaluate novel methods for influencing heat transfer. In the first design, we will consider changing the area! Naturally, it is not possible to change the area of the tubes, but we can influence the area available for heat exchange. The heat transfer coefficient on the shell side is very large for the area where boiling occurs, and it is much smaller for the area where the vapor contacts the tubes. As a first approximation, all heat transfer takes place across the area where boiling occurs. Therefore, the heat transfer is strongly affected by the liquid level in the shell.

One control system is shown in Figure 6.85, and it is evaluated in the following.

- ***Range:** The design has a wide range, since the level can vary from zero (no boiling; no heat transfer) to full (maximum boiling and maximum heat transfer)*
- ***Gain:** Changing the effective value of UA has a strong influence on the heat transfer.*
- ***Dynamics:** The dynamics are moderately fast. Increasing the level can be achieved by increasing the liquid flow to the shell, but decreasing the level requires a longer time to boil more liquid than is entering the shell.*
- ***Linearity:** If the shell had straight sides, the area would be linear with the level, but the shell is typically a horizontal cylinder. The relationship between level and heat transfer is close to linear near the middle of the shell and deviates strongly at the extreme ranges of very low and high level.*

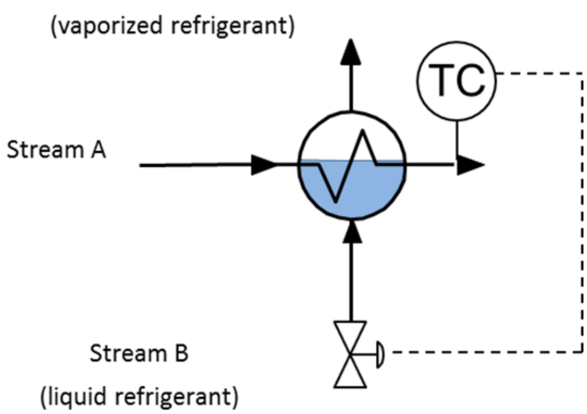


Figure 6.85. Process and control design that influences exchanger duty by influencing the liquid level in the exchanger.

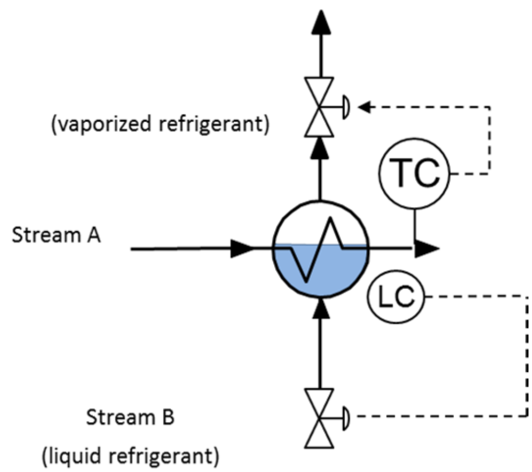


Figure 6.86. Process and control design that influences exchanger duty by influencing the boiling pressure.

- ***Other considerations: Often, this design is used with refrigerants, where the vapor returns to a compressor. In such situations, liquid carryover in the vapor could damage the compressor.***

The design in Figure 6.85 is very commonly used.

Example 6.26 Manipulate the exchange temperature difference - Let's evaluate an alternative in which the temperature is used to affect the heat transfer! The boiling temperature of the bubble-point liquid in the shell depends on the shell side pressure of the heat exchanger.

A design is given in Figure 6.86, and it is evaluated in the following.

- ***Range: The design has a moderate range, because operating at high pressures requires a more expensive shell design.***
- ***Gain: Changing the boiling temperature has a strong influence on the heat transfer.***
- ***Dynamics: The dynamics are fast, which is the major advantage for this design.***
- ***Linearity: The relationship between pressure and heat transfer are not linear, but not strongly non-linear.***
- ***Other considerations: Often, this design is used with refrigerants where the vapor returns to a compressor. In such situations, the extra pressure drop across the vapor line valve increases the work by the compressor; this can be a decisive negative factor in selection of a control system.***

In general, the design in Figure 6.85 is preferred for heat exchangers with a boiling refrigerant because of its lower work requirement; the design in Figure 6.86 is used when fast dynamics are essential.

Example 6.27 Steam condensation in an exchanger - Now, let's consider the very common heat exchanger using steam as the heating source. Steam near saturation is provided to the exchanger to heat the other fluid. The steam condenses at nearly constant temperature to provide the energy being transferred.

We could use the principle discussed above by influencing the area for steam condensation, as shown schematically in Figure 6.87a.

- ***Range: The range is large because the liquid can cover from none (maximum exchange) to all (no exchange) of the surface.***
- ***Gain: The gain would be large.***
- ***Dynamics: The dynamics are not symmetric. To increase the area for condensation, the valve can be opened to drain liquid, which would be fast. However, to decrease exchange the valve is closed and the water must accumulate from the condensing steam, which could be slow.***
- ***Linearity: From the discussion on dynamics, it is clear that the response is non-linear.***
- ***Other considerations: A minor advantage for this design is a smaller valve in the liquid line.***

An alternative steam exchanger design shown in Figure 6.87 drains all condensate so that no area is covered by water. The condensate in collected in a steam trap that periodically releases the condensate to return to the boiler, without allowing steam to escape. There are many steam traps using various principles; one float-type design is shown schematically in Figure 6.88.

- ***Range: The range is large because the steam valve can be adjusted from fully opened to closed.***
- ***Gain: The gain would be large.***
- ***Dynamics: The dynamics are symmetric. All of the area is available for condensation, and the transfer depends upon the steam available. The dynamics are moderately fast, although the liquid and metal of the exchanger influence the dynamic response.***
- ***Linearity: The linearity of the system depends on the relationship between the valve position and the steam flow, which can be made nearly linear by the selection of the proper valve characteristic.***
- ***Other considerations: Minor disadvantages for this design are a larger valve in the steam line and the maintenance required for the steam trap.***

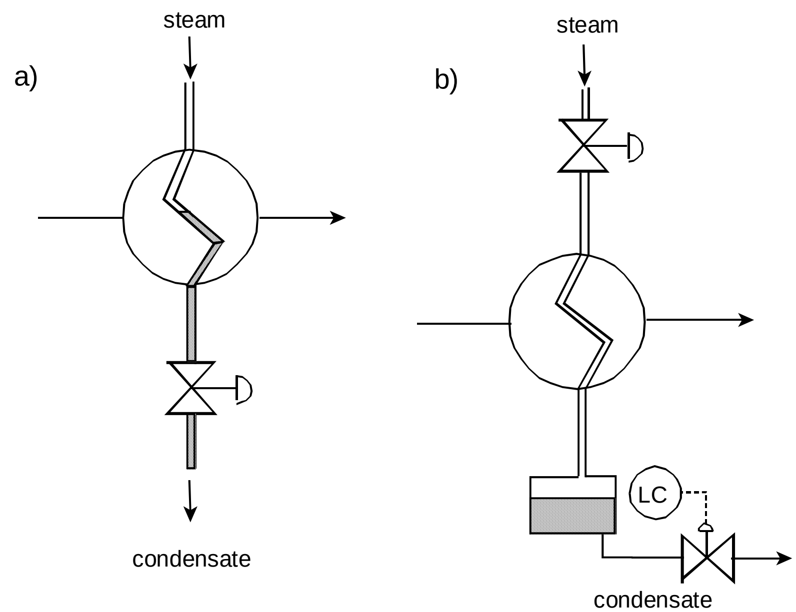


Figure 6.87 Control for a steam heat exchanger. (a) manipulating the condensate flow and (b) manipulating the steam flow. The condensate collection in (b) is typically achieved using a steam trap. Marlin (2000)

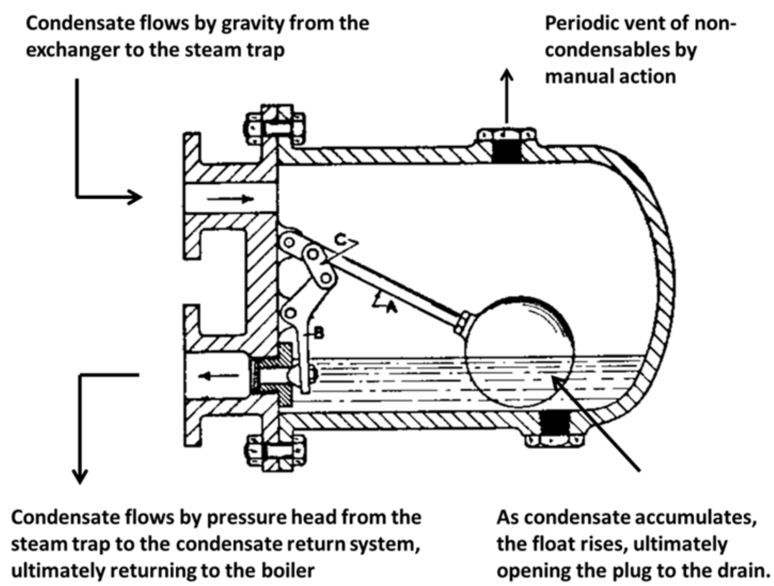


Figure 6.88 Typical float-type steam trap. (Sugartechnology, 2012)

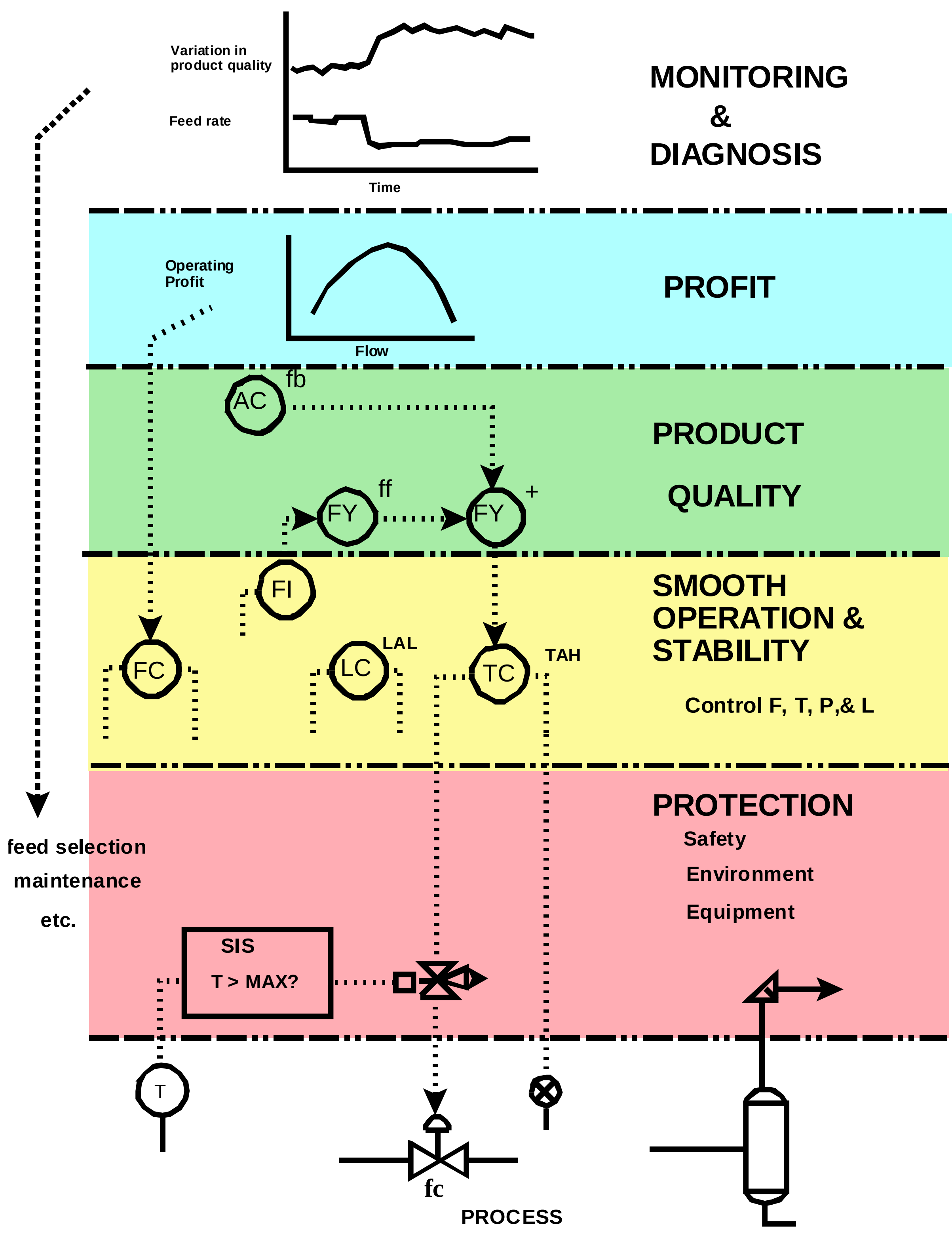
6.7. Conclusions

We will consider our progress in learning process control by considering the control hierarchy shown in Sidebar II. This hierarchy provides an overview of plant operations decision making with several categories based on a temporal decomposition. The fastest responses are required at the lowest level, with each subsequent level involving lower frequency decisions. The location of the decisions in the hierarchy is based on the importance of the decision and the process dynamics of the part of the process being considered. Let's briefly consider each level, discussing topics covered and topics for future study.

Level	Topics covered	Topics for future study
A problem-solving approach (all levels)	In **Section 6.5**, a systematic control design problem definition form is provided. A flowchart for design is given. Some plant-wide issues are presented.	Plant-wide control design remains a fertile topic for investigation, and likely will for a very long time.
The process Process equipment and structure, along with the control equipment, strongly influences dynamic behavior	The influence of the process design on control performance is presented in **Section 6.3.** Instrumentation is addressed in **Appendix A**.	This topic provides opportunity for innovation, especially with highly non-linear processes and newly developing process technology.
Protection This level addresses personnel safety, equipment protection and appropriate processing of effluents.	**Chapter 5** on Safety addressed the safety hierarchy, including the basic process control system. In this chapter, further examples of basic process control are covered.	Inherently safe process design Quantitative safety analysis • Fault tree analysis • Reliability analysis • Computing structures for highly reliable process control systems
Smooth Operation and Stability This level provides a consistent process environment and is critical for successful production rate and product quality control.	In **Section 6.2**, classical control methods beyond single-loop were introduced, including cascade, feedforward, signal select, split range, and inferential,. Some examples of unit-operation control are given in **Section 6.6**.	Multiloop control remains the basis for much control at this level. • Further analysis of multiloop control • More examples of control for unit operations • Greater depth in partial control concepts
Product Quality Direct measurement and control of product quality is essential.	Multiloop control techniques introduced in **Section 6.4** is often required because of strong interactions among quality variables.	A major advance in process control has been achieved through Model Predictive Control (MPC) that provides a centralized or coordinated

	 Inferential control is introduced in **Section 6.2.5**.	method for controlling numerous variables simultaneously. In some instances, quality cannot be measured in real-time. Inferential estimates play an increasingly important role, and multivariate statistical methods have proven successful in industrial applications.
Profit Plant operation has a strong influence on profitability.	Methods for determining the best operating conditions have not been addressed here. **Appendix B** gives a good insight into the economic benefits for process control and gives methods for estimating benefits for project development.	This is a major area of innovation in process control. Controls can be designed to "approximately" follow a high-profit strategy, see • Partial control • Self-optimizing control Models can be used in real-time to predict the best operating conditions, see • Real-time optimization • Optimal blending
Monitoring and Diagnosis People play a crucial role in plant automation. They must monitor, diagnosis and intervene to correct faults.	**Chapter 9** on Troubleshooting addresses this issue directly for short-term faults that need correction.	Every process must be monitored for performance to measure the following. • Performance (efficiency, yield, energy consumption, etc.) • Equipment status and likelihood of future fault • Need for maintenance (e.g., heat exchanger cleaning)
Plant planning and scheduling (Not shown in Sidebar) Important decisions include raw material contracts and purchases, daily and weekly production schedules, plant inventory management, and product dispatch	This issue is not addressed in this chapter.	Longer-term operations decisions require detailed economic calculations involving process performance predictions and analysis of dynamics. The dynamics include inventories, shipping, and responses of multiple companies in a supply chain. These are challenging and interesting problems for engineers.

Sidebar II Overview of Control Hierarchy



This brief summary of the control hierarchy and topics covered in operability demonstrates that (i) you have made considerable progress in building skills and knowledge and (ii) you have the opportunity to learn and to innovate. Process control is not a "solved problem", requiring mere application of known solutions and calculations. In fact, process control is a toolbox of technologies and methods that can be tailored to achieve good dynamic performance for a seemingly unlimited number of process structures.

You are now ready to apply process control to challenging new problems not discussed here or solved before by anyone!

Additional Learning Topics and Resources

For an **outstanding** public-domain reference for instrumentation, see the following reference.

Kuphalt, T (2012) Lessons in Industrial Instrumentation,
http://www.openbookproject.net/books/socratic/sinst/

The basic control topics in Sections 6.1 and 6.2 are covered more thoroughly in many process control textbooks. At the risk of appearing immodest, I would recommend the following.

Marlin, T. (2000) Process Control, Designing Processes and Control Systems for Dynamic Performance, McGraw-Hill, New York

Multiloop control is the basic approach for regulatory process control. While the principles are straightforward, developing "industrial strength designs" requires expertise. The following references give the insight into the importance of disturbance and process direction.

Skogstad, S. and M. Morari (1987) Effect of Disturbance Directions on Closed-loop Performance, IEC Res., 26, 20292035.
Stanley, G., M. Marino-Galarraga, and T. McAvoy (1985) Short-cut Operability Analysis: 1. The Relative Disturbance Gain, IEC Proc. Des.Devel., 24, 1181-1188

A control valve depends on an actuator to provide the force for movement and can have the additional component of a positioner to improve the precision of movement. The following Web sites have additional information.

http://www.spiraxsarco.com/resources/steam-engineering-tutorials/control-hardware-el-pn-actuation/control-valve-actuators-and-positioners.asp
http://iamechatronics.com/notes/general-engineering/436-final-control-elements-control-valves?start=10

Engineers build understanding and expertise by learning best practices. There is much to learn from reviewing designs of common unit operations, such as boilers, firer heaters, distribution networks, refrigeration cycles, compressors, turbines, pumps, and so forth. The following references give an introduction to many practical control designs.

Drieger provides designs for several industrial unit operations http://www.driedger.ca/
Liptak, Bela (1999) *Instrument Engineer's Handbook, 3rd Edition, Process Control*, CRC Press, Boca Raton
Luyben, William, Bojrn Tyreus, and Michael Luyben, (1999) *Plantwide Process Control*, McGraw-Hill, New York.
Dukelow, S. (1986) *The Control of Boilers*, ISA, Research Triangle Park, N.C.

Whenever we perform an engineering task, we should ask, "How can we monitor the performance?" Process control systems are functioning essentially continuously for years, and plant personnel are overwhelmed by the data generated every 200 ms for each of the thousands of control loops in a plant. Fortunately, technology is available to monitor and diagnose operating control systems without interfering with their normal functioning.

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Appendix 6.A Understanding Frequency Response

An important factor in dynamic performance is the frequency at which disturbances occur in comparison to the process dynamics. This topic is more exactly analyzed in using frequency response, but chemical engineers have difficulty with frequency response calculations. (Setting s=jω in the transfer function and finding the amplitude and phase angle of the resulting complex number; do you remember that?) Here, we will consider the issue qualitatively in the time domain. We will build insights that are used in analyzing process systems and complement detailed frequency-response calculations.

Disturbances occur because of changes to surrounding equipment and varying sources of utilities and raw materials. These are more or less random and generally do not follow patterns; however, they generally occur over a narrow range of frequencies. For example, the fluctuation of feed composition to a distillation tower can be caused by imprecise control of an upstream unit, and the fluctuation might involve oscillations with periods of 10 to 30 minutes. To simplify the analysis here, we will consider a sine disturbance, while recognizing that it is only an approximation for disturbances that change roughly periodically with time.

We will consider the performance of the output variable to sine inputs. To gain some basic understanding, let's consider the behavior of the concentration at the outlet of a mixing tank to a sine variation in the inlet flow concentration. The explanation follows from the drawing in Figure 6.A.1, which shows two disturbances, each with a different sine frequency in the input concentration but with the same amplitudes. The maximum effect of the input change on the tank concentration depends on the amount of "difference in" the component A that is introduced during a half period. Naturally, the + and - deviations from the mean cancel, so that the net effect of the average output is zero. The size of the input integral highlighted in crosshatched red determines magnitude of the transient deviation in the output. For the periodic disturbance, we note that t**he larger the integral, the larger the output magnitude.** Clearly, as the frequency increases, the integral (crosshatched red) decreases, so that the output magnitude decreases.

We can now consider the behavior of a typical process without and with control in response to a sine disturbance.

- **Without control** – How would a process respond to a sign disturbance? The response of the process without control to a sign disturbance is shown graphically in Figure 6.A.2a. We observe that at very low frequencies (relative to the time constant, so that $\tau\omega \ll 1$) the output magnitude is equal to the product of the system gain and input magnitude, i.e., the dynamics have little effect on the disturbance. In addition, we observe that at very high frequencies (relative to the time constant, so that $\tau\omega \gg 1$) the output magnitude becomes very small with respect to the product of the system gain and input magnitude, i.e., the process dynamics attenuate the disturbance without control! The results are displayed in a Bode plot in Figure 6.A.2b.

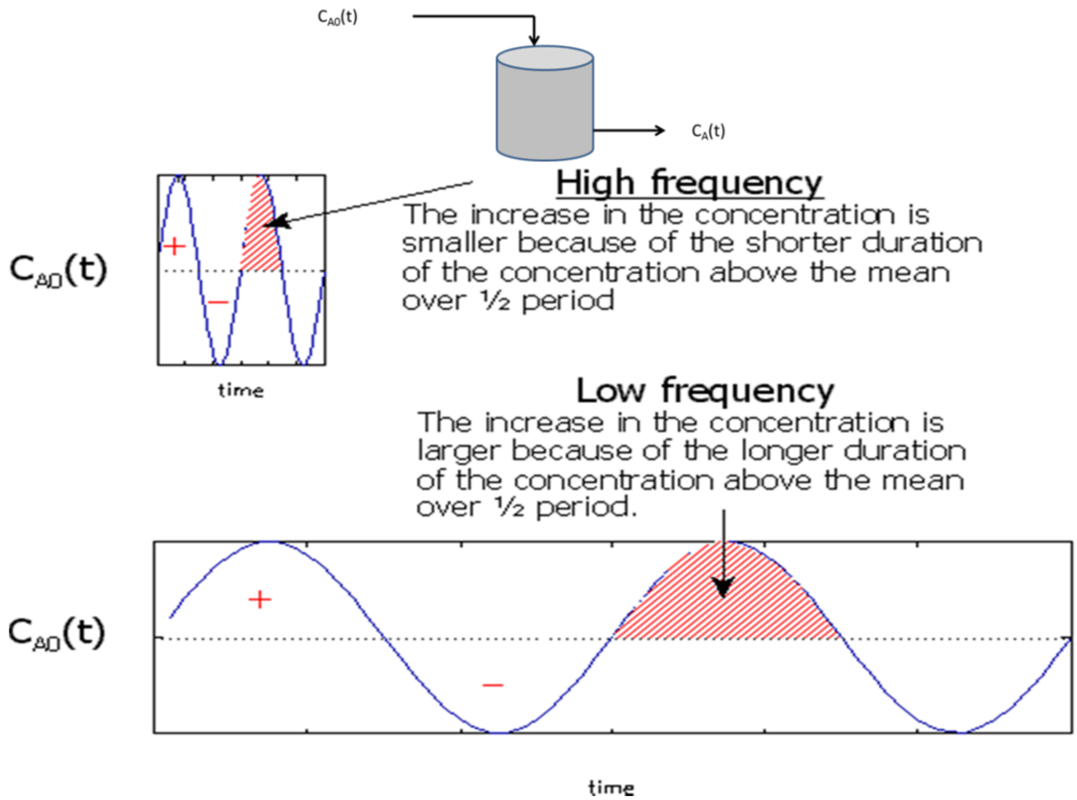


Figure 6.A.1. Effect of input frequency on the tank concentration

With control - How would a process under feedback control respond to a sign disturbance? The response of the process with feedback control to a sign disturbance is shown graphically in Figure 6.A.3a. At very high frequencies, we know that the process itself, without feedback, decreases the effects of disturbance. What happens at low frequencies? We observe that at low frequencies (relative to the time constant, so that $\tau\omega << 1$) the output magnitude is very low. This behavior results for the effectiveness of feedback, which has lots of time to respond to the slowly changing disturbance. We also observe that neither the process itself nor the feedback controller are effective in reducing the effect of the disturbance near the frequency where $\tau\omega_c \approx 1$, i.e., the resonant frequency. The results are displayed in a Bode plot in Figure 6.A.3b.

The observant reader might be wondering, "Since perfect sine waves don't occur in chemical plants, is any of this useful?" The answer is, "Yes!" Even though perfect sine waves do not occur, many disturbances can be characterized by a range of frequencies. Combined with knowledge of the feedback dynamics, an understanding of frequency response can be applied to predict the location of the dominant disturbance frequency in the major regions of system behavior shown in Figure 6.A.3b.

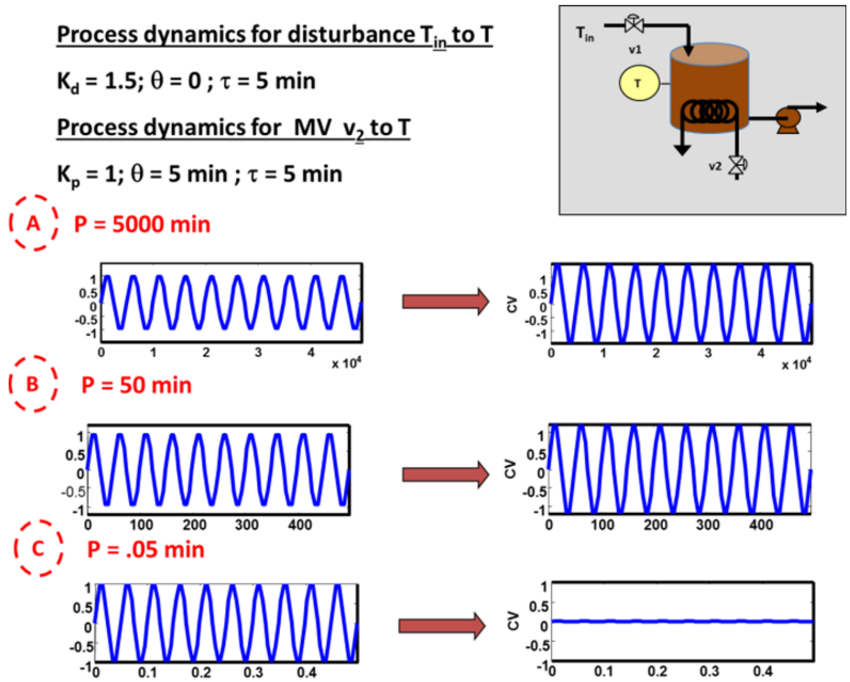


Figure 6.A.2a The effect of a sine disturbance on the process with control at three frequencies.

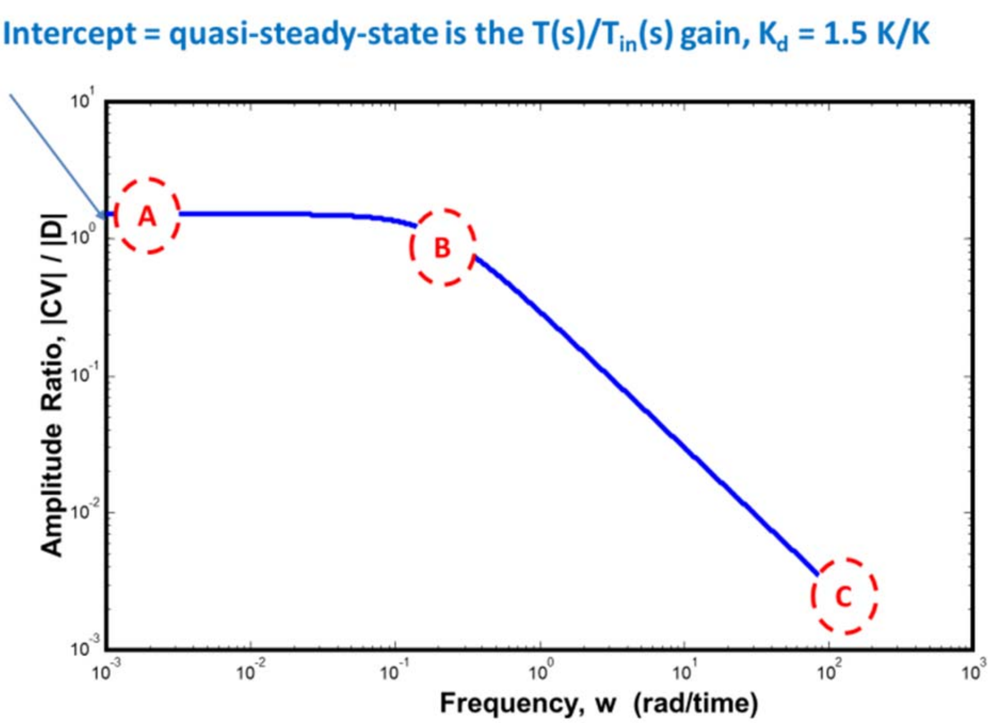


Figure 6.A.2b Bode plot of the amplitude ratio for a disturbance response with control. (CV = controlled variables, D = disturbance)

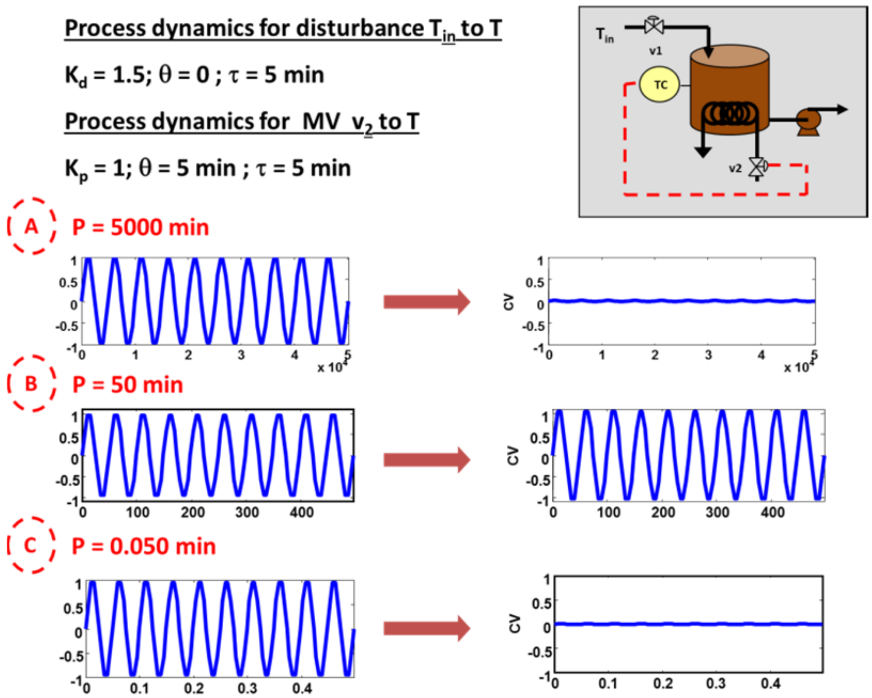


Figure 6.A.3a. The effect of a sine disturbance on the process with control at three frequencies.

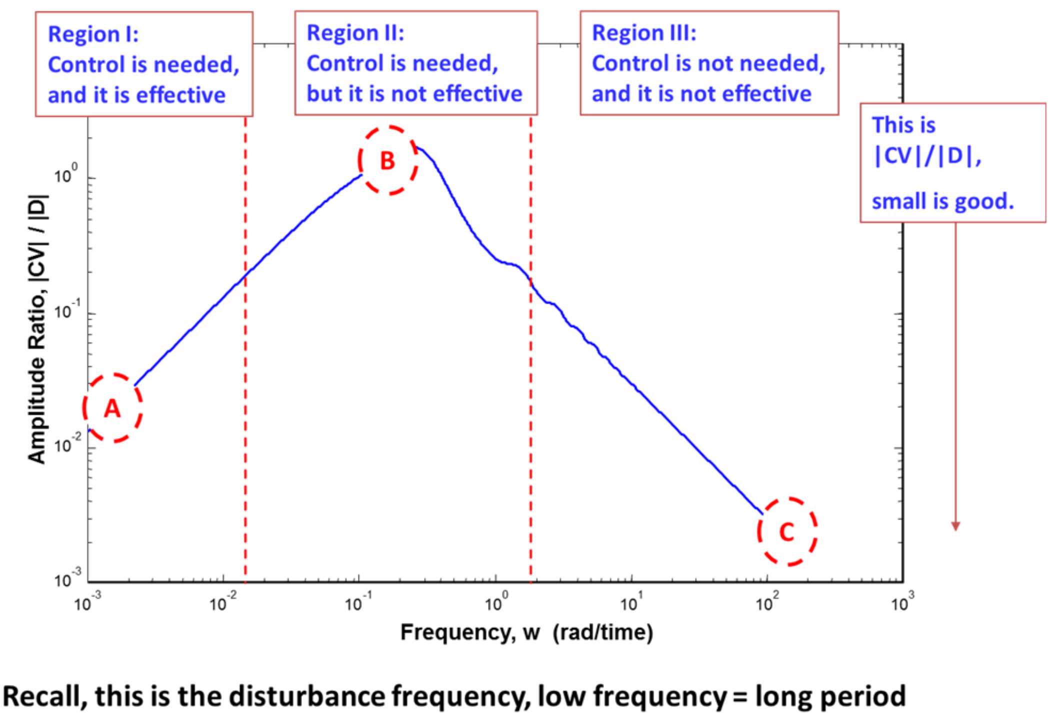


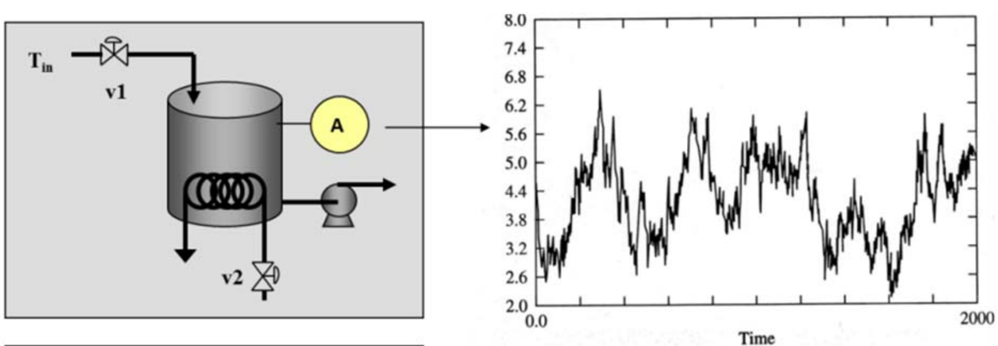
Figure 6.A.3b. Bode plot of the amplitude ratio for a disturbance response with control. (CV = controlled variables, D = disturbance)

Let’s consider the example in Figure 6.A.4, which shows a process, gives its open-loop dynamics and shows some characteristic data for a variable without control. The major design questions are, “Can feedback control reduce the variability of the output variables, and if so, by how much?” We note that the data has a very long period (very low frequency) for the dominant variability compared with the feedback dynamics. In addition, we note that the data contains some high frequency components that are of lower amplitude. We would predict the following.

- For the lower frequency variation, a large fraction of the uncontrolled amplitude can be reduced by feedback control
- For the higher frequency variation, very little (if any) of the uncontrolled amplitude can be reduced by feedback control

This prediction is confirmed by the data in Figure 6.A.5, which shows the performance with feedback control.

The preceding analysis was qualitative. However, a more reliable quantitative method for predicting future control system behavior is available. For example, see Harris (1987) and Jelali (2006). These quantitative methods are much preferred for engineering practice, and software products are available to perform the calculations. In addition, they have the advantage of being able to use original data with feedback control to predict possible performance improvements with modified controls.



Feedback dynamics are:

$$\frac{A(s)}{v(s)} = \frac{1.0e^{-2s}}{2s+1}$$

Data without feedback control

Figure 6.A.4. Typical process operating data without control.

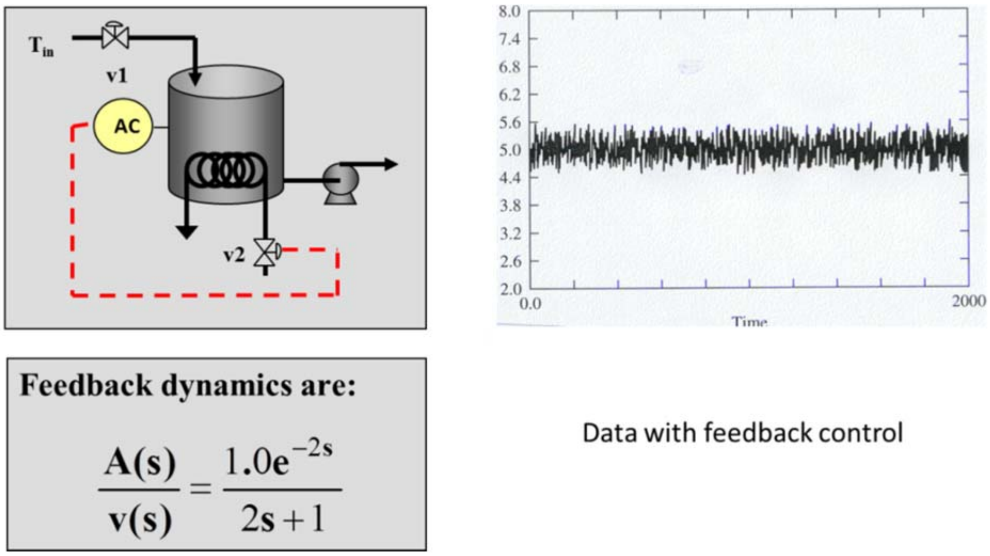


Figure 6.A.5. The reduction in variability achieved through feedback control. The process and disturbances are the same as shown without control in Figure 6.A.4. Marlin (2000)

Appendix 6.B Benefits from Process Control

6.B.1 Introduction

Process control is implemented to achieve many goals. The primary goal is safety. The design of the safety hierarchy is addressed in Chapter 6. Other control strategies are required by various regulations, e.g., by equipment and process technology vendors, or to satisfy government regulations, such as effluents to the environment. However, many control strategies are not essential and therefore can be considered "optional" methods for achieving product quality, production rate, and equipment protection. They must be justified based on economic criteria. The economic analysis of a control system is addressed in this appendix.

Process control achieves benefits by reducing variability. Perhaps, it is better to state that in most cases, the variability is transferred from very important variables to lesser important variables. For example, the original data without product quality control in Figure 6.B.1a shows high variability in the analyzer measurement and no variability in the coolant flow. When feedback control is implemented as shown in Figure 6.B.1b, there is much less variability in the product quality but much more variability in the coolant flow. This is why we say that the variability has been "transferred". Naturally, we design feedback control to transfer the variability from variables where it is costly to variables where the cost is extremely low.

One example of where variability is removed without transfer is the reduction of variability due to time constants in a process. For example, variability in a liquid feed property (composition or temperature) can be reduced significantly through mixing in a vessel. In addition, the flow rate variability can be reduced by averaging level controller tuning. (For averaging level control, see Marlin, 2000).

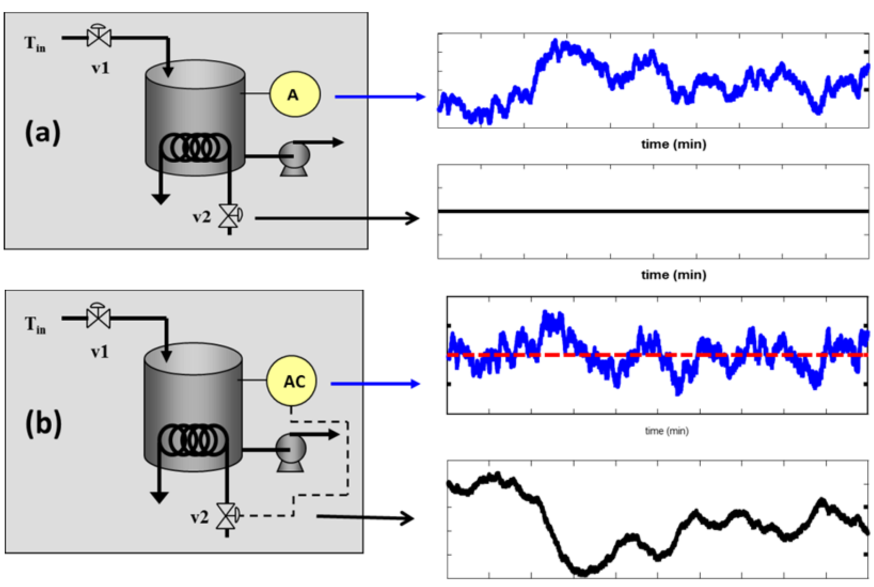


Figure 6.B.1. Variability and feedback control.
(a) without feedback control, (b) with feedback control

6.B.2 Benefits Calculation

Now, let's introduce the basic equation for estimating control benefits.

$$B = \Delta V(D_0, D_1) * M * {}^{SF}/_{100} * T \tag{6.B.1}$$

where

B	=	benefits for the change in dynamic performance ($/y)
D_0	=	original (base case) distribution in controlled variable (histogram)
D_1	=	distribution of the controlled variable under new control strategy (histogram)
$\Delta V(*,*)$	=	Improved economic process performance at the base case operating conditions ($/h)
M	=	correction factor for other operating conditions, e.g., production rate, (dimensionless)
SF	=	service factor, the fraction of time that the control strategy is improving the process performance (dimensionless)
T	=	Time when the control strategy should be in service (h/y)

A few important issues are worth noting. First, the calculation does not use average values of variables; it uses the distributions of controlled variable values. Since process control effects an improvement by reducing variability, this use of distributions before and after control seems reasonable. Unfortunately, this correct policy is not always followed in practice, which can lead to erroneous benefit predictions. Second, the benefit can be influenced by other variables in the process. For example, many benefits are proportional to production rate; an example would be energy saved in a distillation tower. Third and finally, the benefits depend on the amount of time in the year when the control strategy is functioning.

The distribution of the key controlled variable is used in the benefit calculation. The distribution is easily determined from plant data, as shown in Figure 6.B.1. With the data represented by the distribution, we can readily evaluate the average process performance using the following equation, as shown schematically in Figure 6.B.2.

$$PR_{ave} = \sum_{i=1}^{m} F_j * PR_j \tag{6.B.2}$$

with

PR_{ave} = average process performance
PR_j = process performance in interval j (evaluated at mid-point of interval)
F_j = fraction of data in interval j
m = number of intervals

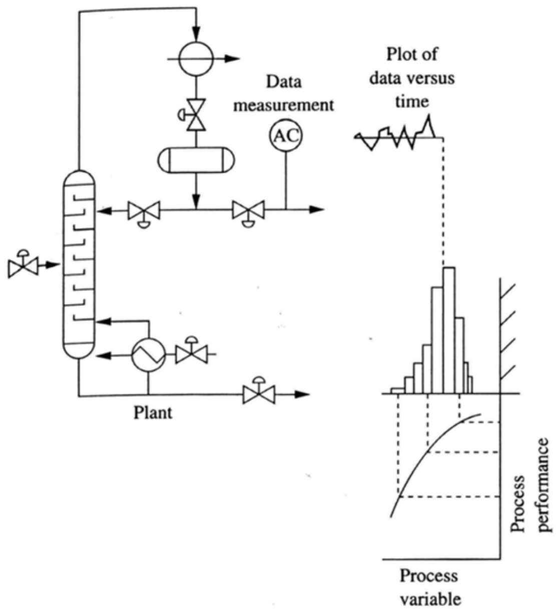


Figure 6.B.2 Evaluating variable distribution from trend data. Marlin (2000)

Note that the process performance is typically expressed as a physical variable, such as energy/throughput, percent yield, and so forth. Then, a second relationship is employed to translate the change in process performance to a change in economic performance, i.e., the variable V in equation (6.B.1). This is shown in the following equation.

$$\Delta V = \Delta PR^* I_v = [(PR_{ave})_{base\ case} - (PR_{ave})_{improved}]\ I_v \tag{6.B.3}$$

with

$(PR_{ave})_{base\ case}$	=	the valuable of process performance with base case operation
$(PR_{ave})_{improved}$	=	the valuable of process performance with improved process control
I_v	=	the incremental value of the change in average process performance
PR	=	process performance

Note that this expression includes a constant incremental value, independent of the magnitude of the change in process operation. In some situations, the incremental value depends on the magnitude of change, and equation (6.B.3) can be easily modified to reflect the variable incremental value when calculating the change in economic performance.

Much more detail is important in this calculation procedure, but before addressing further detail, two examples are presented.

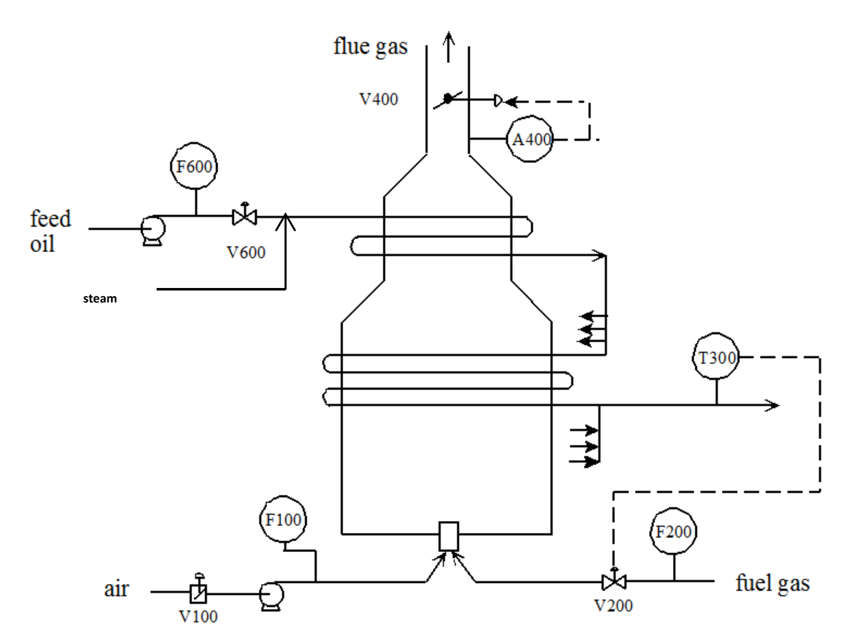


Figure 6.B.3 Ethane pyrolysis fired heater/reactor.

Example 6.B.1 Improved temperature control of a pyrolysis fired heater – A pyrolysis reactor "cracks" hydrocarbon feed to lighter products, with the primary goal of producing olefin products for subsequent use in polymerization reactors. Here, we will consider a single heater that processes ethane feed shown in Figure 6.B.3.

To determine the economic performance, the engineer requires the variable distribution and the relationship between the variable and the process performance. Data with the initial temperature control is shown in Figure 6.B.4a. (This data was previous considered briefly in Figure 6.44.) The process performance correlation is also plotted in Figure 6.B.4.

Applying equation (6.B.2) to the data in Figure 6.b.3a, the average ethane conversion can be evaluated to be 54.6%.

Several methods are available to predict the performance with improved control, and these will be discussed subsequently in this appendix. For the purposes of demonstrating the calculation, we will assume that the prediction in Figure 6.B.3b can be achieved. Applying equation (6.B.2) to the data in Figure (6.B.3b), the average ethane conversion can be evaluated to be 59%. We conclude that the change in process performance is the following.

$$\Delta P = (P_{ave})_{base\ case} - (P_{ave})_{improved} = 59\text{-}54.6 = 4.4\%$$

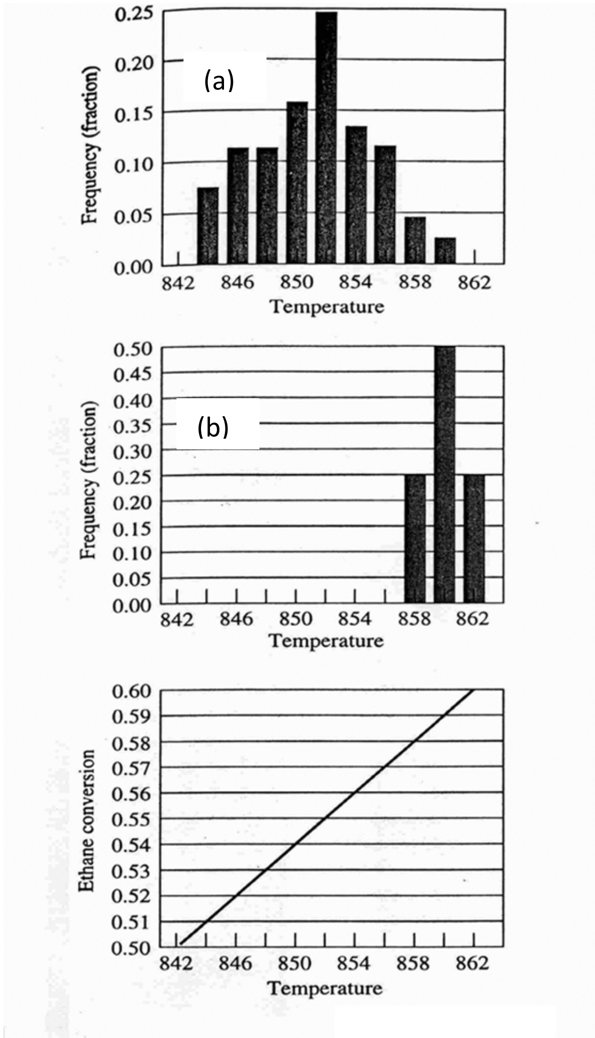


Figure 6.B.4 Distribution data and process performance correlation for Example 6.B.1.
a. Base case distribution, b. Distribution with improved control Marlin (2000)

To determine the change in economic performance, the incremental value of the change in feed conversion is needed. This value requires a simulation of the entire plant, because the unreacted ethane flows through the entire plant, requiring considerable energy, and is recovered and recycled to the pyrolysis reactor, where it can react again. We will take the commodity prices of ethane and ethylene in \$/kg to be 0.286 and 0.925, respectively. If we assume that the benefit is 30% of the difference in feed and product prices (based on yields including recycle), the value of the change in conversion is evaluated in the following.

Value of increased conversion/kg = $(\Delta PR/100)*I_v$ = (0.044)(0.639)(.30) = \$0.00843 \$/kg

ΔPR = change in yield due to improved control
I_v = incremental value

The total value depends on the production rate. A typical industrial pyrolysis heater can produce about 548 m-Ton/day of ethylene, requiring about 913,000 kg/day of ethane. Therefore, the economic improvement can be calculated to be the following.

$$\Delta V = M\ (kg/h) * \Delta PR\ (\%\ feed)/100 * I_v\ (\$/kg) = (0.00843\ \$/kg) * (0.913x10^6\ kg/day)$$
$$\Delta V = 7.7\ x10^3\ \$/day$$

Normally, we evaluate process economics using annual values. This reactor will be operating continuously when the plant is in operation. However, it requires periodic maintenance to remove by-product coke, and during this maintenance, the productive reactions do not occur. The maintenance will require one day, and eight of these maintenance decokes will occur every year. In addition, the plant will shut down once per year for five days for maintenance. Summing these periods when the control system cannot contribute improvements, one obtains the following.

$$SF/100*T = 365 - 8 - 5 = 352\ days/y$$

Note that this analysis assumes no unplanned shutdowns. The plant history could be reviewed to determine whether unplanned shutdowns contribute a significant a significant loss in operating time.

All of the terms can be combined to determine the profit realized through improved temperature control of a pyrolysis fired heater. The result is given in the following.

$$B = \Delta V * SF/100 * T = (7.7\ x10^3\ \$/day) * (352 day)$$
$$B = 2.7\ x\ 10^6\ \$/y$$

This is an enormous economic benefit! It would certainly justify considerable engineering effort to achieve the reduced variability in temperature control. Note that the base case is poorer than would likely be observed in practice today. But, we can conclude that the benefit for operating closer to the high temperature limit is greater than four hundred thousand dollars per year per degree Celsius. Naturally, these results depend strongly on the costs of materials and the market conditions, i.e., whether it is possible to sell the extra production.

We note that the reduction in variability in Example 6.B.1 was accompanied by a change the controller set point that increased the average temperature. In this example, a simple reduction in variability without the change in set point would not have improved performance. In general, the engineer must decide whether a set point adjustment is appropriate. Let's consider an example where the set point is not adjusted.

Example 6.B.2 Flue gas excess oxygen – Naturally, air is mixed with fuel in the burner of a combustion process. Sufficient air for complete combustion is essential so that no combustible material exists in the flue gas, because the flue gas with combustibles could mix with leaking air and explode. However, any air not required for complete combustion leads to inefficiency; it is heated and exhausted to the environment via a smoke stack. As a result, excess air leads to a waste of fuel. Therefore, the goal is to ensure a slight excess of air at all times but to keep the excess small. This situation is shown in Figure 6.B.5, which depicts the efficiency curve and the constraint of minimum excess oxygen in the flue gas.

The distributions are given in Figure 6.B.6 for the excess oxygen data for the base case, with higher variability, and the improved case, with lower variability. We can apply equation (6.B.2) to determine the efficiency for each distribution, and then, we can determine the increase in efficiency as given in the following.

$$\Delta PR = (PR_{ave})_{improved} - (PR_{ave})_{base\ case} = 87.7\text{-}86.8 = 0.90\%$$

To determine the change in economic performance, the incremental value of the change in efficiency is needed. The incremental value depends on the fuel cost and the base case consumption of fuel. We will use a 5 $/GJ cost of fuel and a base case fuel to the boiler of 100 GJ/h. Therefore, the economic improvement can be calculated to be the following.

$$\Delta V * M = \Delta PR \text{ (\% efficiency)}/100 * M \text{ (GJ/h)} * I_v \text{ (\$/GJ)} =$$
$$(0.9/100)*(5)*100 = 4.5 \text{ \$/h}$$

Normally, we evaluate process economics using annual values. We will use 340 days operation to allow time for maintenance shutdowns. Therefore, the annual benefit would be the following.

$$B = (4.5 \text{ \$/h}) * (24 \text{ h/day})*(340 \text{ day/y}) = 37 \text{ k\$/y}$$

This is a substantial savings by simply improving and existing control system.

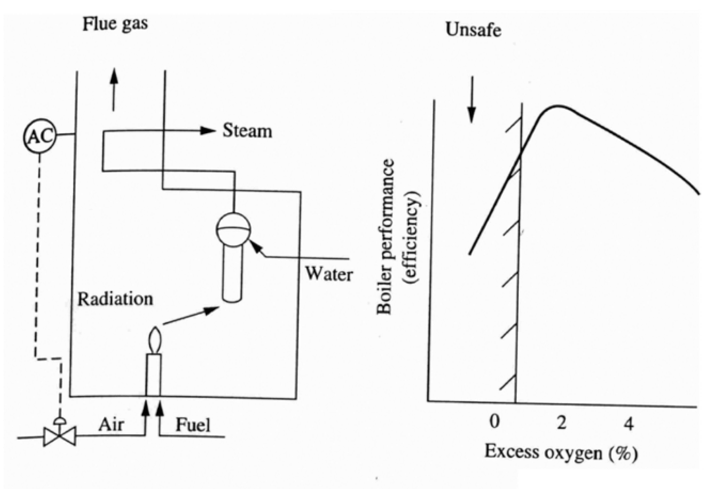


Figure 6.B.5 Boiler and efficiency curve versus excess oxygen in the flue gas Marlin (2000)

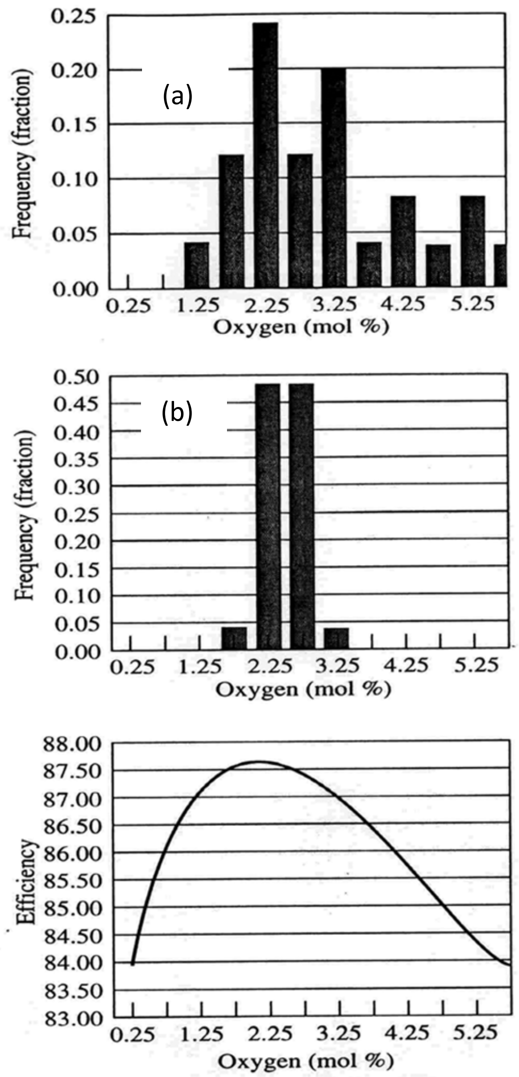


Figure 6.B.6 Distribution data and process performance correlation for Example 6.B.2.
a. Base case distribution, b. Distribution with improved control. Marlin (2000)

Note that the base case and improved distributions have essentially the same set points. In this case study, the entire benefit was realized by reducing the distribution around the same average value. If the engineer incorrectly evaluated the base case data using the average value of excess oxygen, the incorrect conclusion would have been that no improvement in efficiency was possible. This example highlights the importance of using the distribution of dynamic plant data when evaluating process performance.

The method just demonstrated evaluates the performance and profitability of a process based on the distribution of values for a key process variable. Like all engineering calculations, it has specific advantages and is based on assumptions that must be recognized by the engineer. Therefore, some key aspects of the method will be presented in the following.

6.B.3 Issues in the Benefits Calculation

Many potential pitfalls exist when applying the benefits calculation approach. Eight of the more common pitfalls are discussed here to help the reader avoid committing these errors. The most challenging issue, predicting the future performance, is addressed in the next section.

6.B.3a. Base case data – The calculations for the base case use historical data from the process. The historical data must be "representative". Representative data should characterize the behavior of the process in the future if no changes were made to the control system. Also, the variability in the data should be caused by factors that can be influenced by the feedback control being proposed.

Let's consider the sample data in Figure 6.B.7. The data includes typical variability and two major deviations from normal operation, one due to a process shutdown and one due to an equipment limitation. The equipment limitation could have been maintenance on a pump, heat exchanger, etc. The variability over the period is strongly affected by these major deviations. But, will the proposed control system prevent the shutdown or the maintenance event? If not, the base case data must exclude these events.

Obtaining data is very easy with digital control systems that store historical data for all variables. However, the example just discussed points out the dangers of extracting data from history without carefully reviewing it. If all of the data were used to characterize the base case, the improvement possible with improved control could have been grossly overestimated.

6.B.3b. Variable distribution – The distribution is not assumed to follow any typical distribution, normal, etc. The distribution does not have to be symmetric or continuous. It must only be "representative". Representative data should characterize the behavior of the process prior to any proposed changes by the project.

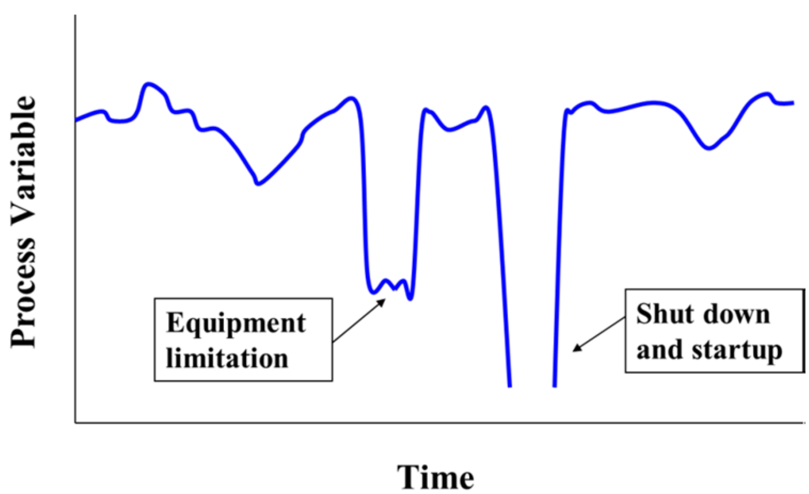


Figure 6.B.7. Historical data with events not affected by control. Marlin et. al. (1987)

Non-symmetric frequency distributions are common when a high penalty is paid for violations of a constraint. When operators are responsible for adjustments, they will make very strong corrections when nearing a constraint. When a control system is making adjustments, the design engineer can introduce a non-linear effect in the controller to take strong corrective actions when the variable approaches the constraint. The method described here can properly represent process performance with non-symmetric distributions.

6.B.3c. Process performance curve – The process performance curve does not have to conform to any specific criteria. It need not be linear, convex, or continuous.

Although we do not want process variables to violate constraints, they will occasionally exceed their limits due to large disturbances. The process performance correlation can represent the penalty, as shown in Figure 6.B.8. As expected, the performance becomes worse during violations, and the performance can suffer a step decrease immediately upon the violation. Naturally, performance curves with other shapes are possible.

6.B.3d. Frequency of variability – The condensation of trend data into a histogram results in the loss of frequency information. Because frequency data is not included in the histogram, the same histogram could summarize a process with either high frequency or low frequency variation. While all data are inherently dynamic, the process performance correlation nearly always applies to steady state. Therefore, the typical assumption made is that the process data variability is of low frequency and the process is operating at quasi-steady state, so that the process correlation can be applied.

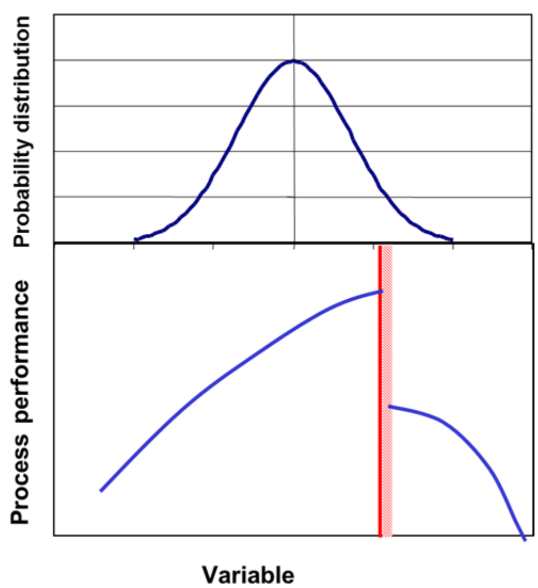


Figure 6.B.8. Typical process performance curve with discontinuity at constraint violation.

6.B.3e. Economic value of change – The engineer must determine the economic value for the change in process operations. This is often not as straightforward as one might initially think, because the economic change must represent all changes to the operation of the process resulting from the control system. Let's consider some issues.

- The net change in profit includes positive and negative effects. If improving product quality requires additional energy, the net improvement is the sum of the positive quality value and the negative value of additional fuel.

- It is not correct to use average values for energy, feeds, products, utilities and so forth. Incremental or marginal values must be used, since they represent the change in profit from a base case, and it is this change that the control system will effect. These incremental values should not include the fixed costs for items not influenced by a small change in the control system. These fixed costs include major capital equipment recovery (equipment not changed by the control design), personnel, laboratories, offices, waste treatment, etc.

- The process change must be related to costs, i.e., to purchases and/or sales. This principle is demonstrated in the distillation example in Example 6.B.3.

- The benefit for control improvement is an element in a project profitability analysis using the standard time-value of money approach, such as net present value or discounted cash flow. Costs include additional installed equipment (sensors, valves, computing equipment, etc.), engineering time for design and technician time for programming.

- The incremental value for a change in operation is not necessary constant. As the change increases in magnitude (or changes sign), the process responses can change. The changes can be dramatic, as when constraints are encountered; for example, Smith and Varbanov (2005) show substantial changes in the cost for steam as letdowns (direct connections between headers) are opened.

Example 6.B.3 Reduced steam for distillation reboiling – A control system has been proposed to reduce the steam use in the reboiler of a distillation tower in Figure 6.B.9. How is the value of the steam reduction determined?

The steam is generated in the plant, so that the effect of a reduction in steam must be traced through the boiler and steam system to determine the reduction in fuel achieved by the reduction of steam in the distillation tower. (If the steam were being purchased at a constant value in $/ton, determining the value would be easy.) As shown in the figure, a reduction of reboiler steam (i) reduces one outflow from the medium pressure steam header, (ii) reduces the steam through the extraction valve from the turbine, (ii) since the turbine is required to provide a fixed amount of power, the steam flow through the condensation path must increase (less that the extraction decreases), (iv) less flow leaves the high pressure header, (v) to balance the flows in the high pressure header, the pressure controller reduces the fuel flow.

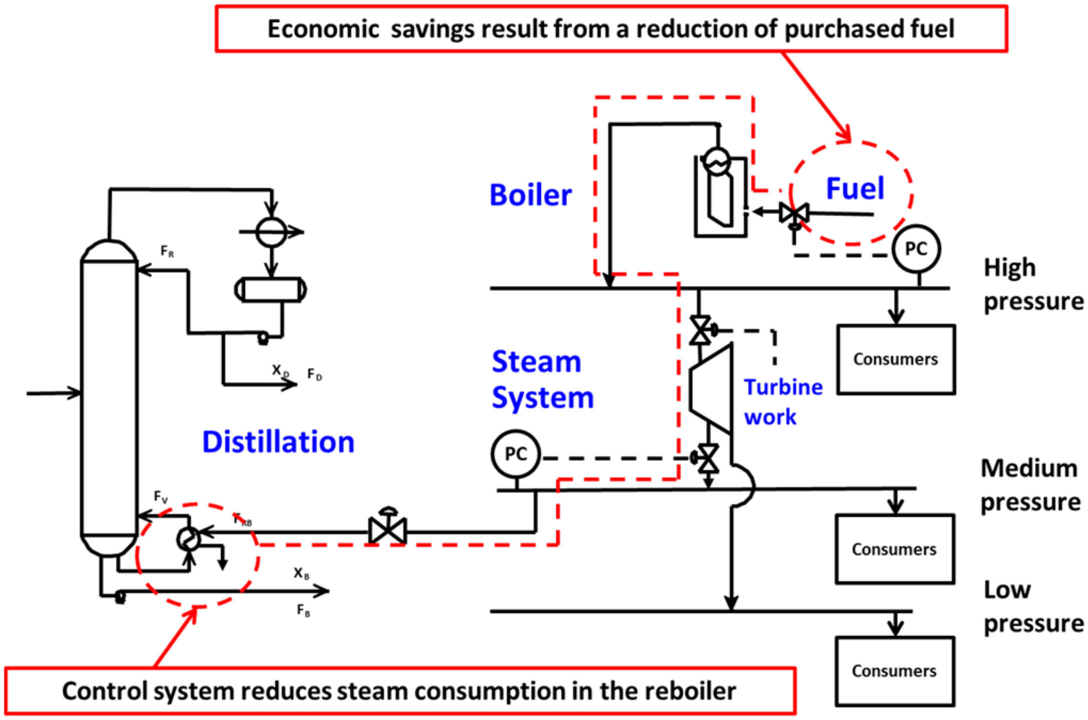


Figure 6.B.9. Tracking the steam change to fuel purchase.

Why not use the cost of high-pressure steam, which is easily determined from the cost of fuel, heat of vaporization and the boiler efficiency? For example, a typical cost of high-pressure steam is 5.32 \$/ton, while the cost of medium-pressure steam is 2.61 \$/ton. (These costs are for a typical steam network and fuel cost of 4.22 \$/GJ or 0.0152 \$/kWh (Smith and Varbanov, 2005).) Using the simplified high-pressure value would have yielded a benefit estimate about double the correct value!

6.B.3f. Operating conditions on benefits – Benefits for a base case operation can be influenced by changes to the operating conditions. Perhaps the strongest effect is the production rate through the unit. Let's consider two examples.

Example 6.B.4 Distillation energy savings – With poor control, distillation towers are over-refluxed to ensure achieving maximum impurity concentrations. When the distillate product quality is controlled in real time, the reflux and energy consumption are reduced. How does the tower feed flow rate affect the energy benefits for control?

With constant tray efficiency, the optimal energy would be proportional to the feed rate. However, this relationship is limited to the region in which the energy input can maintained proportional to feed rate. One limitation results from the minimum reflux ratio that is imposed to maintain the liquid and vapor rates in the region that provides adequate liquid-vapor contact without weeping. When the feed flow rate is very low, the compositions cannot be controlled to set points; the tower must be "over-refluxed", which results in products purer than necessary, but within specifications. This situation occurs when the feed is substantially below the design value, which can occur due to market fluctuations.

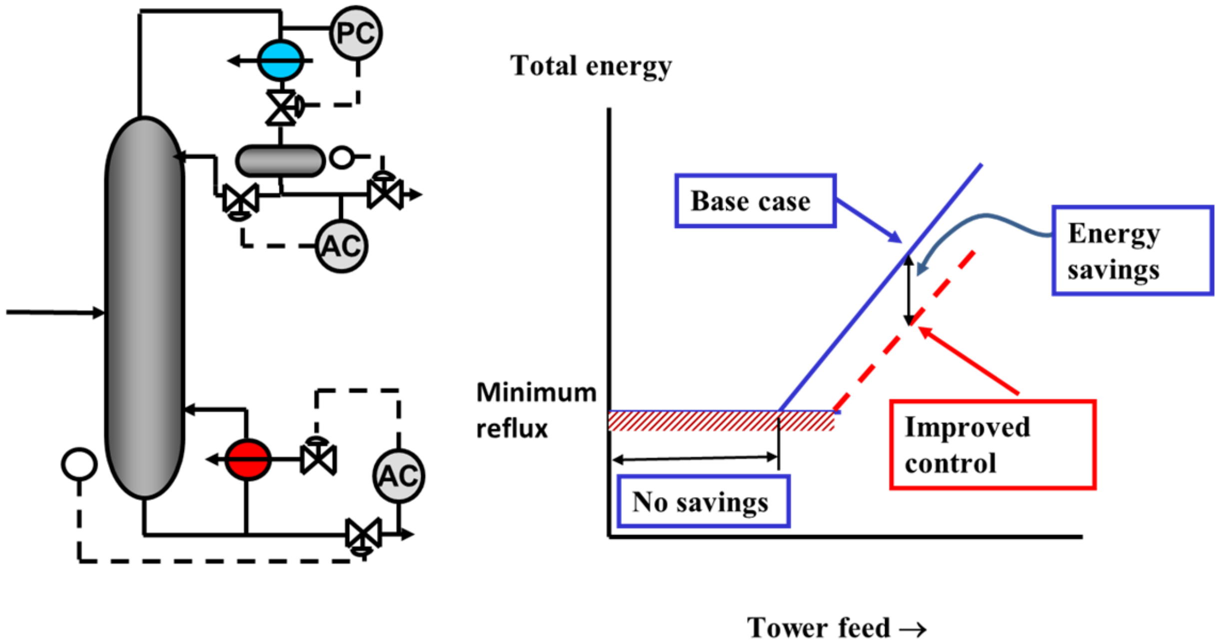


Figure 6.B.10. Distillation energy reduction occurs in a limited range of feed flow rate.

When evaluating the benefits for control, the feed rate has to be predicted over the project length (many years), and the benefits adjusted to account for any time when the feed rate is expected to be in the "no savings" region. Any reduction would appear in the "T" factor in equation (6.B.1).

Example 6.B.5 Compressor energy savings – Rotatory compressors require a minimum flow rate to prevent damage, which is required to prevent "surge" that can reverse flow direction in the high-speed machine. All compressors have at least basic anti-surge control that provides recycle to provide sufficient flow through the compressor when the feed flow is less than required. A basic control is shown in Figure 6.B.11. How do benefits for anti-surge control depend on throughput flow?

The minimum surge flow depends upon operating conditions such as compressor speed and gas molecular weight. Thus, the simple control design in Figure 6.B.11 would have to a very conservative, i.e., high, minimum flow because the recycle flow controller does not account for the other operating conditions. More advanced control designs use additional real-time measurements to provide updated estimates for the minimum flow required (Staroselsky and Laudin, 1979; Smith and Kurz, 2005, Engencyclopedia, 2012); lower surge flow limits result in smaller recycle and less energy consumption by the variable-speed machine powering the compressor. The comparison of base case and advanced anti-surge control is shown in Figure 6.B.11. It is apparent that benefits occur at low feed flow rates and that no benefits are realized at high flow rates. Again, the engineer must estimate the distribution of flow rates that will occur in the future and adjust the energy savings accordingly.

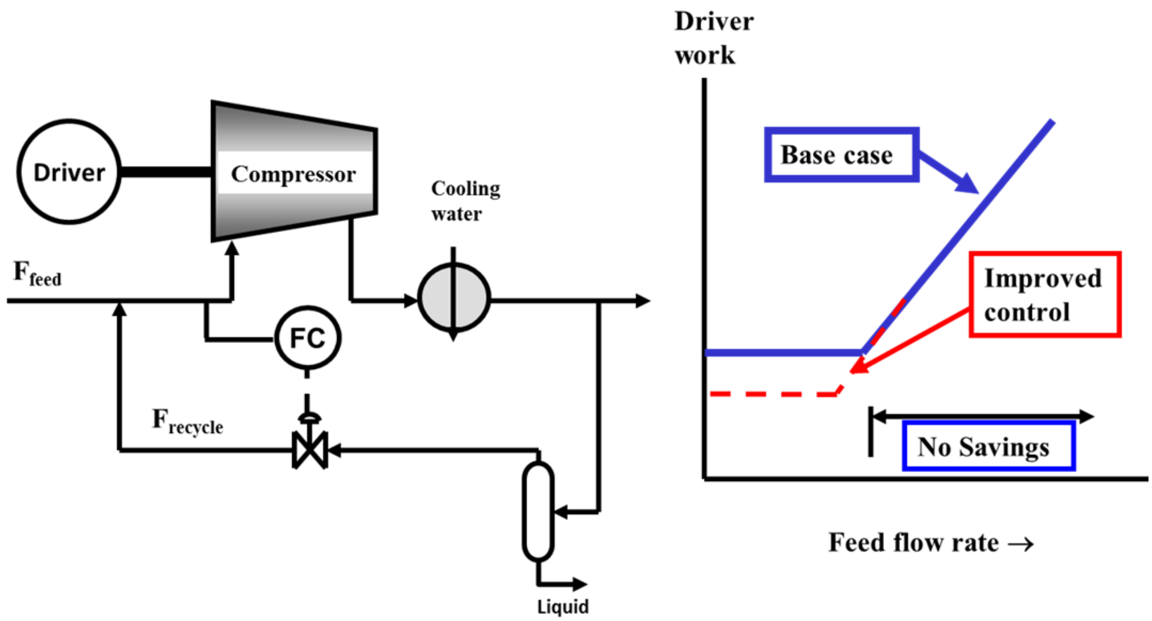


Figure 6.B.11. Compressor anti-surge energy reduction occurs in a limited range of feed flow rate.

6.B.3g. Mixing effects on process performance – Often, mixing can be represented as a linear process. However, complex reactions and other chemical interactions can lead to non-linearities that strongly affect the behavior of material properties that have been stored in inventory. Often, engineers place inventories for streams with key composition specifications. The purpose is to provide mixing so that upstream quality control does not have to be excellent, because the tank outlet composition is easily maintained at the specification. The following example demonstrates a potential negative impact of inventories.

Example 6.B.6 Effects of Non-linear mixing – A crude distillation unit separates crude oil into many streams, each of which is processed in subsequent processes. Often, storage tanks are located between crude distillation and subsequent units so that the production rates in various units can be determined independently, at least over short durations. The crude distillation unit and diesel storage tank are shown in Figure 6.B.12. The engineers operating the unit suggested that good quality control of the stream leaving the distillation tower was not important because mixing in the storage tank attenuated variability. Is this a good strategy?

The contention that the storage tank attenuated composition variability is certainly correct. But, at what cost is the variability reduced? Many petroleum product qualities do not blend linearly. For example, the diesel cloud point specification is non-linear, so that a small amount of "heavier" material in the blend requires a larger amount of "lighter" material; requiring the lighter materials substantially reduces the yield of valuable product. The benefits for improved control of streams into storage have been studied by Marlin et. al. (1987), and typical results are shown in Figure 6.B.12. Improved distillation control could yield an increase in diesel yield of 1.5% of feed (with a concomitant reduction in heavy gas oil) that represents an enormous economic benefit.

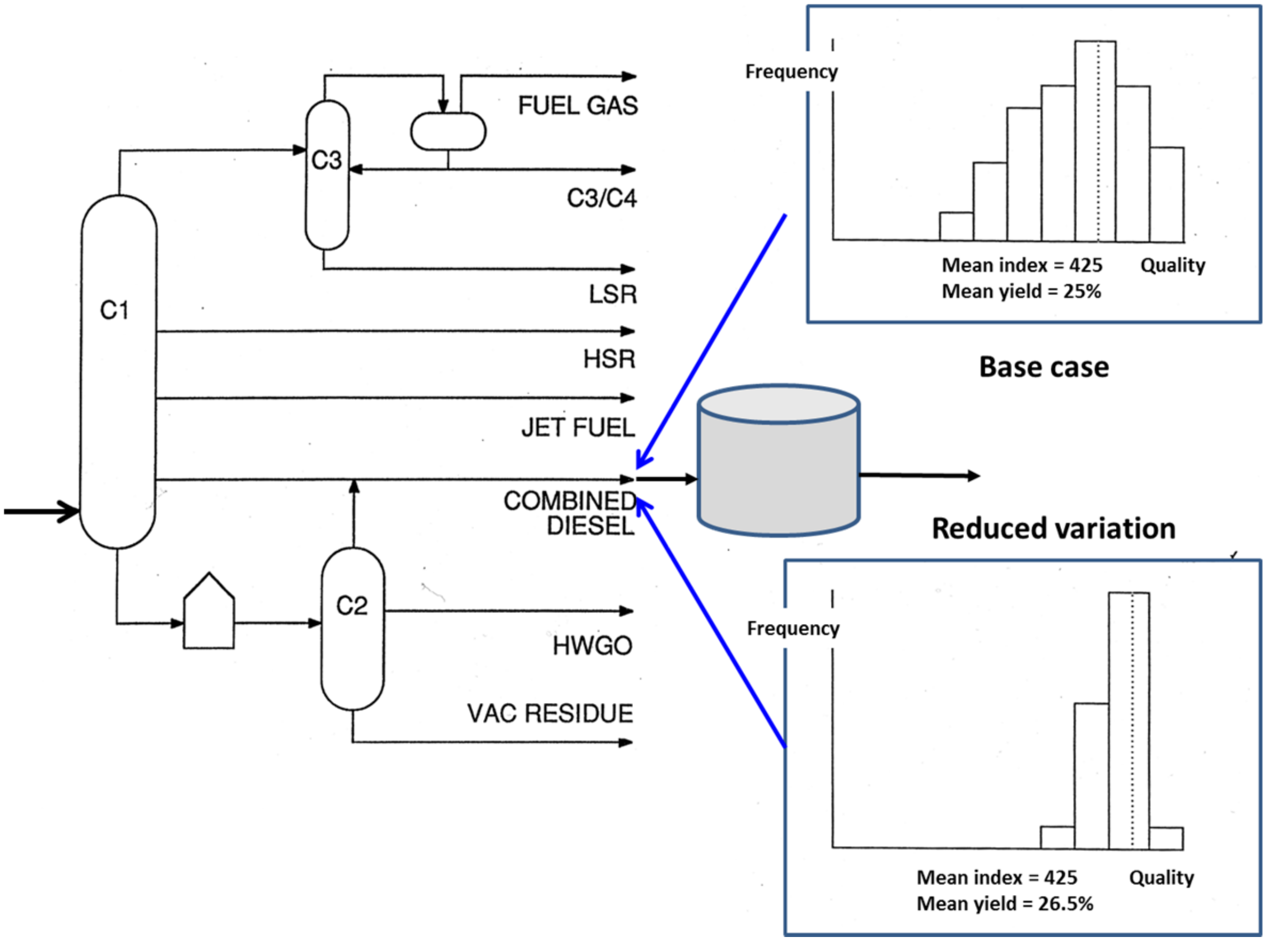


Figure 6.B.12. Crude distillation showing importance of stream composition control before mixing subject to non-linear blending relationship. Newell et. al. (1987)

6.B.3h. Incidence reduction – Reducing the variability of key process variables can prevent infrequent, large deviations that can damage equipment, activate safety instrumented system (SIS) shutdowns, and if not properly moderated, lead to accidents. Typical process examples include compressor anti-surge control, chemical reactor temperature control preventing runaway temperature excursions, and positive displacement pump recycle to prevent excessive pressure. This topic is addressed in Chapter 4 on Reliability. Naturally, the reduction in such incidents can lead to considerable savings and should be included in the economic analysis of process control designs.

In this section, the clear message was delivered, "Understand your data, process principles, equipment performance, operating window limitations, product quality, production rate goals, and economics." With mastery of the situation, the engineer can adapt the benefits calculation to address special features of the problem. However, a major issue remains, i.e., the prediction of future control strategy performance, which is addressed in the next section.

6.B.4 Predicting Future Performance

A pithy quotation is appropriate at this point.

It is difficult to make predictions, especially about the future!

Controversy swirls around the originator of this quote, with attributions to Confucius, Niels Bohr (Danish Nobel Laureate in Physics), Yogi Berra (American baseball player) and many more (Denenberg, 2012). Regardless of who said it first, the quotation is valid and serves as a warning about being overly confident in our predictions. However, we must prepare for the future using predictions, and in engineering, these predictions can be based on solid principles and practices that while not guarantying perfection, often yield results adequate for technical and business decisions.

Here, we are dealing with the challenge of predicting the ability of a control system to reduce the future variability in an operating process. We will begin with some approaches that should not be used. Then, we will consider some approaches that can be used.

Do not use these approaches

- **Do not ignore variability** - Consider a situation in which operating closer to a constraint yields an economic benefit. Sometimes, a person not well versed in plant operation will suggest simply changing the set point without first reducing the process variability with better control. An example of this poor approach is shown for temperature control of steam to a power-generating turbine. High steam temperature improves efficiency of the turbine; however, higher temperatures lead to much lower turbine blade life. High temperature alarms are avoided because they indicate operation that will cause shorter turbine blade life. As demonstrated clearly by data in Figure 6.B.13, raising the temperature controller set point resulted in excessive alarm activation and if continued, would have damaged the turbine (Johnman et. al., 1987). This plant experiment was (appropriately) terminated quickly.

- **Do not assume the answer** - Never simply assume that the base case variability can be reduced by a fixed percentage without analysis of process dynamics. Often, this percentage is 50% of the variance. In reality, the improvement can be much smaller or greater than this arbitrary choice. Assuming the solution is not engineering!

- **Do not base the prediction on un-reviewed historical data** - Do not assume that the best performance in the historical database can be achieved. The best performance can be caused by a combination of conditions that might not occur again. For example, the disturbances might have been temporarily small, the equipment could have been recently refurbished (heat exchanger cleaned, burner adjusted, catalyst regenerated, etc.), and/or the feed material might have the most favorable composition. However, we will reconsider this recommendation with an approach that can be used.

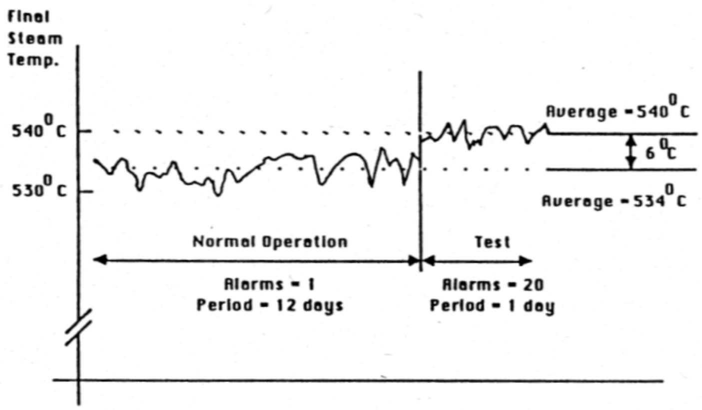


Figure 6.B.13. Changing the set point toward the constraint without improved control leads to excessive constraint violations. Johnman et. al. (1987)

- **Do not assume set points are correct** - When improving the process performance, do not assume that the original (current) controller current set point value is optimal. For example, when investigating the best manner for operating a vinyl chloride process in Figure 6.B.14, engineers recognized that the EDC cracking heater was not being operated at the best temperature (Barton et. al, 1987). The project modified its direction from process control to process optimization, with a real-time optimizer using a process model to predict the best temperature target as operating conditions changed. (For an introduction to real-time process operation, see Marlin and Hrymak, 1986.)

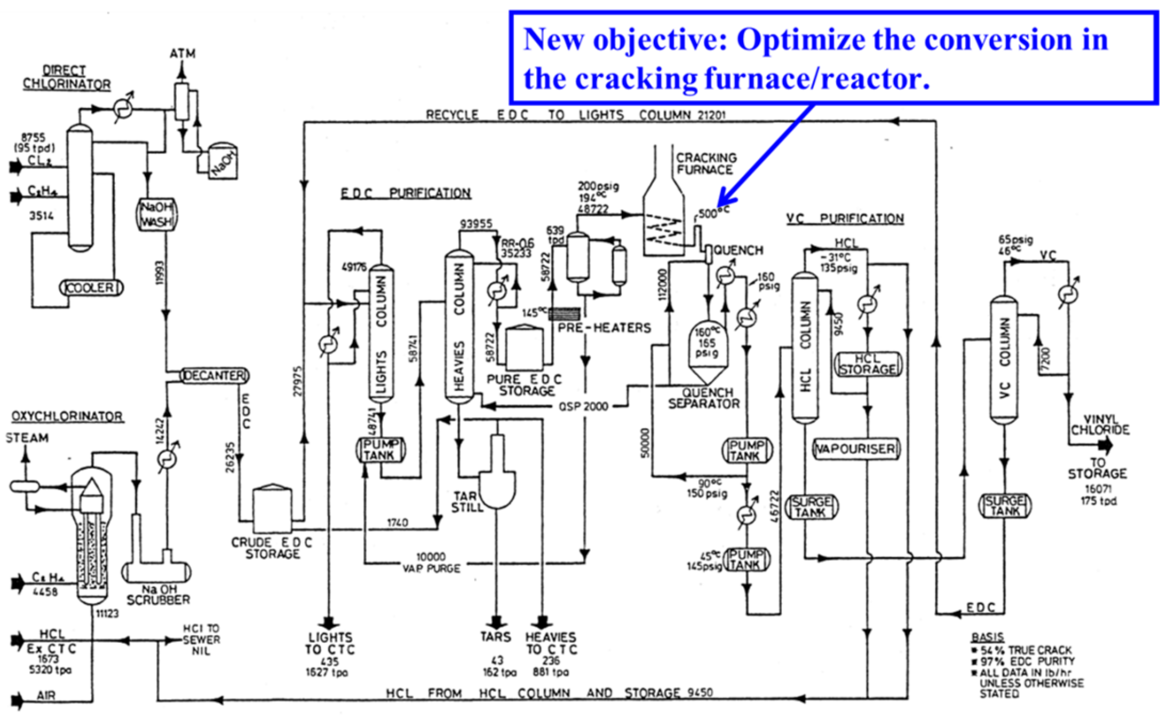


Figure 6.B.14 Greatest benefits realized by optimizing the reactor conversion. Barton et. al. (1987)

- **Do not base business decisions on a literature review** - Finally, do not assume that benefits reported in the literature for similar processes can be achieved in your plant. Control performance depends on factors that could be very different in your plant, such as disturbances (magnitudes and frequencies), equipment capacities, flexibility, energy costs, and product markets. Published reports can be useful in helping an engineer select likely candidates for improvement, but reports on other plants do not replace thorough analysis.

Use these approaches

- **Do use "zero variance" for a quick, limiting estimate** - A simple approach can be used for a quick estimate of whether a significant improvement is possible. The engineer can assume that the future variance of the key variable is zero around its optimal value. We know that this cannot be achieved, but we can estimate the economic benefit with this assumption. If the project is not financially attractive with zero variance, we are sure that it will not be financially attractive with the actual variance achieved by improved control, and the project can be dropped from further consideration. If the project might be attractive, one of the other recommended methods discussed in the following items can be applied to obtain an improved estimate of potential improvement.

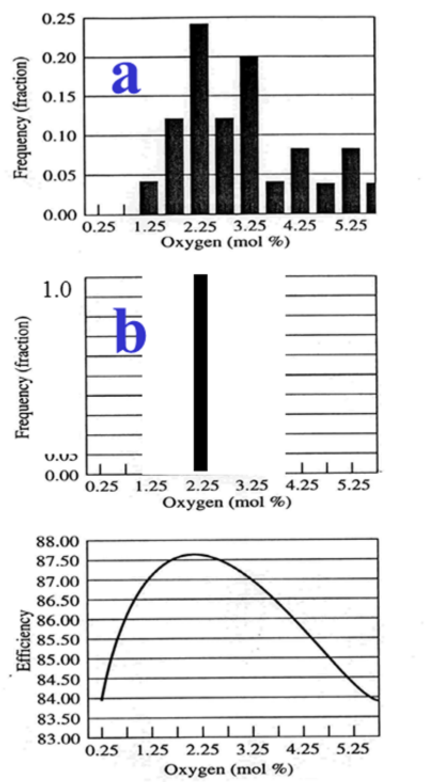


Figure 6.B.17. Example of quick improvement estimate using (a) historical data for the base case and (b) zero variance for the future performance.

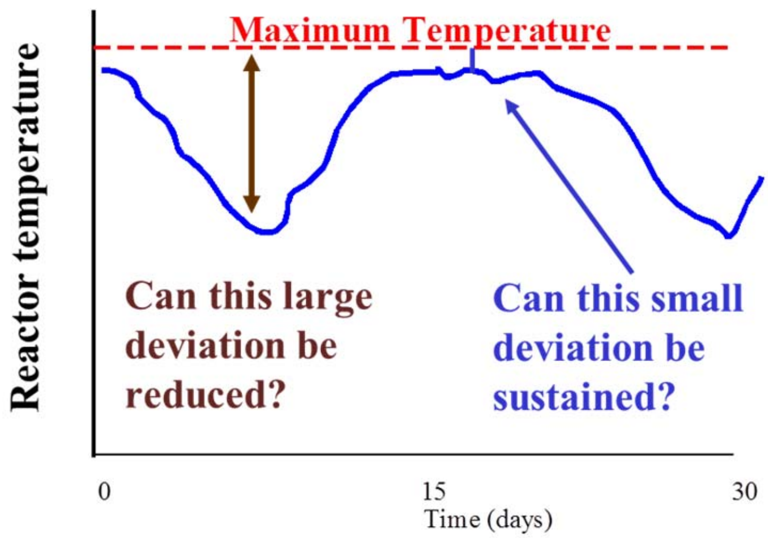


Figure 6.B.15 Historical base case data from a reactor where maximizing temperature had a high economic benefit.

- **Do use best performance after thorough review of scenario** - Let's revisit the approach using past operating performance. The good aspect of this approach is that the data represent actual disturbances, equipment capacity and flexibility. Difficulties with this approach have already been discussed. We can overcome these difficulties by monitoring the process when the data is collected or reviewing detailed operations logbooks, if they exist. The data in Figure 6.B.15 shows base case operation of a reactor where incentive exists to maximize the temperature without exceeding the maximum constraint. The data shows that the temperature was maintained close to the maximum for about seven days. Does this indicate that this good performance can be repeated by improved control for the entire period? The engineer must review the operation to determine if the process was experiencing typical disturbances such as feed tank changes, production rate changes and others that will be typical in the future; if the result is "yes", it can be concluded that this period of good operation can be repeated. Second, there are periods with poor temperature performance. Could control have improved the performance? The engineer must evaluate the equipment performance, capacity and flexibility. If adjustments were possible but not made, improvements could be made. If both answers are yes, it is reasonable to predict that the entire thirty-day period could have been controlled at close approach to the maximum temperature.

- **Do use plant tests** - We could perform plant tests that emulate the proposed control strategy. This approach requires that sensors and final elements exist in the plant and that the process dynamics are slow enough for the control calculation to be performed and implemented periodically by a person. An example is given in Figure 6.B.16, in which crude oil entering the desalter is preheated by vacuum distillation (VDU) pumparound (mid-tower condenser). The goal is to produce a sidestream product in the vacuum tower. In the base case, the heat duty of in the exchanger is too large, resulting in too much condensation below the VDU sidestream, so that no product can be withdrawn in the VDU. Why is the duty high? The operators must ensure that the temperature of the crude oil entering the desalter is above a minimum limit; to be on the safe side, the heat exchanger duty is maintained very high. Plant tests demonstrated that the exchanger duty could be reduced, the side stream product could be withdrawn, and the desalter temperature could be maintained in its desired range. Improvement was possible.

Why was the base case operation poor? The desalter temperature was locally displayed, and the exchanger by-pass valve was manually adjusted. Therefore, adjustments were very time-consuming, so that poor operation was not due to lack of effort by the plant operating personnel. It was a poorly designed plant system; it could not be operated at peak profitability.

- **Do use simulation combined with historical disturbance data** - The previously discussed methods might not be possible, because there is no relevant "best performance data" and emulating the control system is not possible (a sensor or final element is not available). However, if the plant is in operation, data is available for the key controlled variable. Let's look at the situation where (i) trend data is available for the variable without control and (ii) the engineer can obtain an estimate of the feedback dynamics. The dynamics can be estimated using standard experimental methods used in controller design and tuning, like the graphical process reaction curve or statistical methods. The plant data represents the effects of disturbances on the key variable. This plant data can be combined with a linear dynamic simulation of the process and controller to predict the closed-loop behavior of the process. This approach is depicted in Figure 6.B.17; naturally, the process and controller models can be modified to suit the specific system under study.

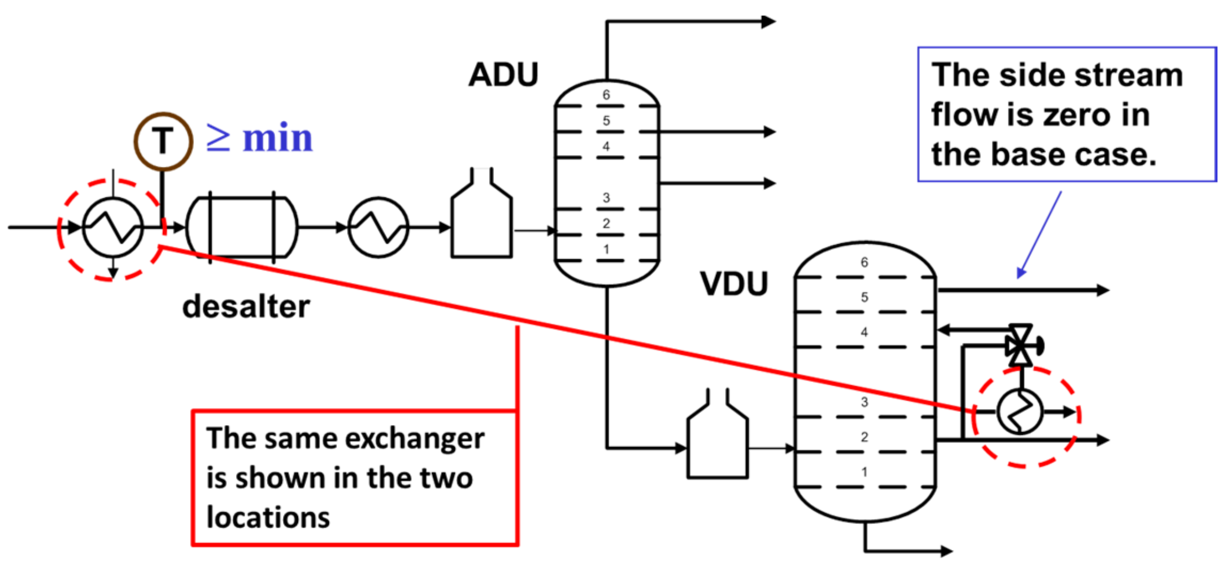


Figure 6.B.16. Process system where crude entering desalter is preheated by vacuum distillation (VDU) pumparound (mid-tower condenser). The goal is to produce a sidestream product in the vacuum tower.

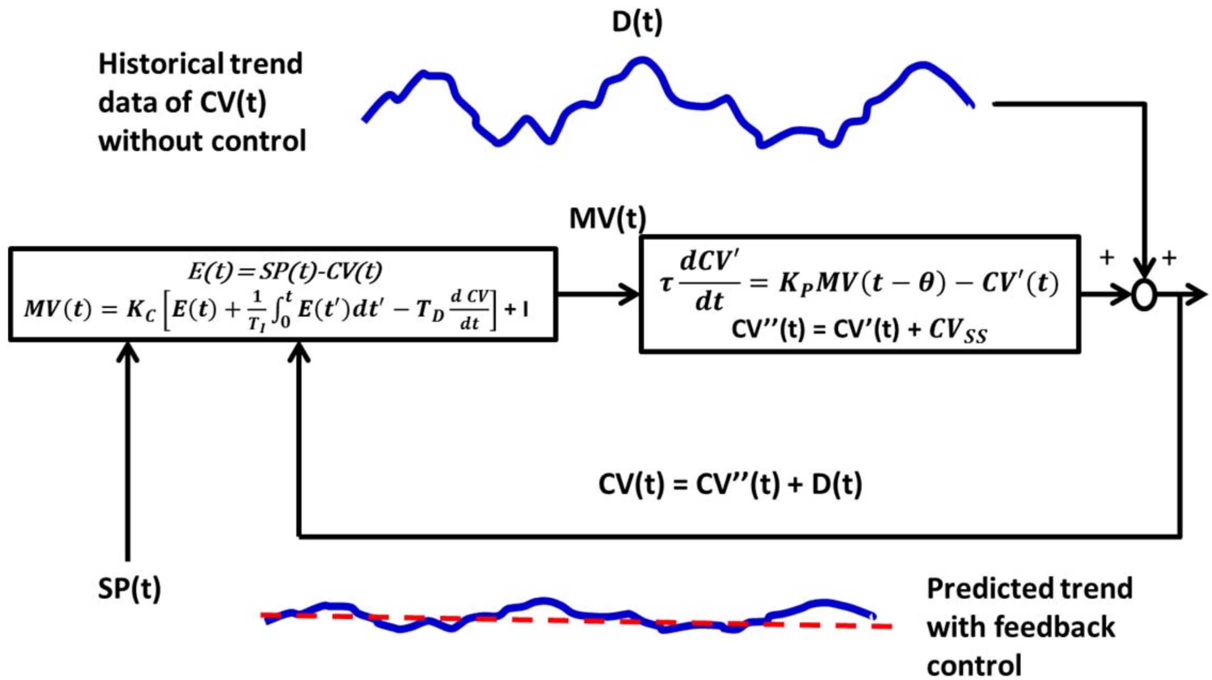


Figure 6.B.17. Schematic of a simulation approach for estimating the control performance under feedback control. Historical data must be without control.

The programming and calculations are simple to prepare and perform. The important limiting assumptions are that (i) the plant data includes no control (including no operator actions for control) and (ii) the process dynamics can be approximated by a linear model over the range of variation experienced in the data and (iii) the historical data are representative of the variability in the plant.

- **Do use advanced statistical method extending the simulation method above** - The requirement that the historical data be free of influence by control is a significant limitation because important variables are typically controlled, even if only periodically by manual adjustments. Fortunately, an advanced approach overcomes this limitation. It performs statistical analysis of the data and predicts the best achievable feedback control performance. The approach can be extended to cascade, feedforward and multivariable control. The topic is more involved than can be presented here; the interested reader can refer to Harris (1987) and Jelali (2006) for further details, and Jelali provides references to software products to perform the calculations and provide helpful visual displays.

- **Do use fundamental models when deviations are large** - In a few cases, the process and control system should be simulated using fundamental, non-linear models. The engineering effort is justified when developing new processes and when analyzing process control for large disturbances and complex physiochemical systems where linearized models do not provide adequate accuracy. An example of dynamic simulation supporting new process development is the famous Tennessee Eastman reactive distillation process (Agreda et. al., 1990). An example of dynamic simulation to evaluate common unit operations connected in a complex system is reported by Harismiadis (2012) for a liquefied natural gas plant design

Developing fundamental dynamic models and simulations could be very time-consuming. Fortunately, software packages are available to assist the engineer; for example, most flowsheeting packages have dynamic modeling capability. These packages provide model libraries, graphical model building, physical property databases and advanced numerical methods.

- **Do consider infrequent, large and costly incidents** - All of the previous methods provide estimates for variance reduction. Process control can also provide substantial benefits by responding properly to very infrequent, large magnitude disturbances that can cause plant shutdowns if not compensated quickly. The report by Zhang and Dudzic (2006) describes a good example of an inferential variable monitoring system that can identify incipient constraint violation and alarm the plant personnel. This application realized good economic benefits and improved the safety of the process by preventing a release of molten steel in areas where people worked. Fundamental simulation studies can also be used to determine the benefits for process equipment and control modifications. Patel et. al. (2012) give a summary of many applications of simulation to predict the dynamic behavior of a complex chemical plant.

6.B.5 Conclusion

Estimating the benefits for process control is an essential skill for engineers. Not only are good estimates developed thorough dynamic analysis, but also information is developed to ensure that the process equipment (capacity and flexibility), instrumentation (sensors and final elements) and the control strategy (feedback, cascade, feedforward, loop pairing, etc.) is designed and implemented to achieve the predictions. In addition, the proposed design must conform to the safety, reliability, and economic goals of the plant.

The topic might more properly be termed “analysis for benefits estimation and preliminary control design”.

The benefits calculation expression is shown schematically in Figure 6.B.18. The reader must keep in mind that the figure is not rigorous, since it implies linear relationships, while the benefits calculation in equation (6.B.1) allows non-linear relationships. However, it serves as a memory aid for the terms in the calculation.

The approaches described in this appendix have been used by practitioners for decades and are well accepted in industry. However, they do not provide a cookbook. They provide a suite of concepts, approaches, and calculations that can be tailored by the engineer to solve many problems in applied process control.

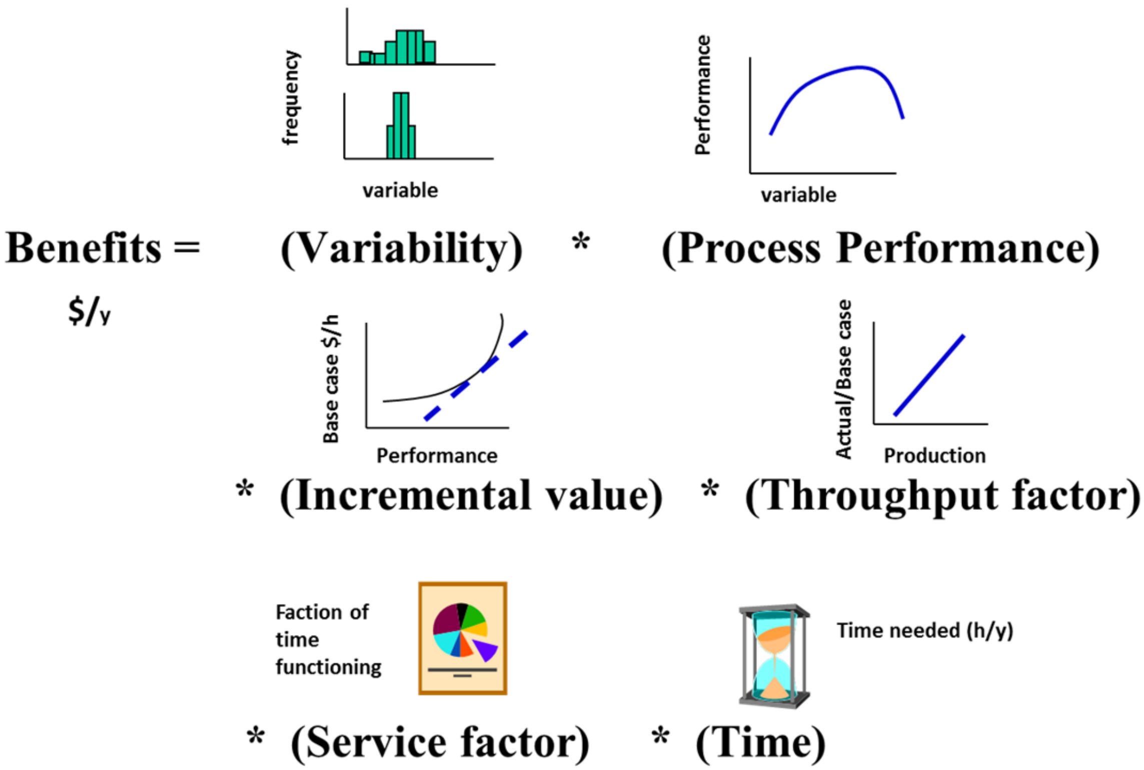


Figure 6.B.18. Schematic of the benefits calculation. (Caution that items might not be separable as shown because of non-linearities.)

The reader should feel comfortable in creatively applying these approaches and where necessary developing innovative new approaches for problems not addressed here.

6.B.6 Additional Learning Resources

This appendix has addressed many important issues in control benefits estimation, and it has provided some practical approaches for engineering practice. Perhaps surprisingly, this topic is not widely addressed in the open literature, although it is of crucial importance to every investment in process control. The references in Table 6.B.1 provide additional details on methods, case studies, and citations for other publications worth reading.

Table 6.B.1 Resources for Further Learning in Process Control Benefits

Citation	Comment on contents
Marlin, Thomas, John Perkins, Geoff Barton, and Mike Brisk (1987) *Advanced Process Control Applications*, ISA, Research Triangle Park. (Book out of print)	This book describes a set of seven industrial studies applying the approach presented in this appendix.
Marlin, Thomas, John Perkins, Geoff Barton, and Mike Brisk (1991) , *Benefits from process control: results of a joint industry-university study*, J. Proc. Cont., 1, 68-83	This journal article summarizes the results of the above studies.
Bauer, M. and I. Craig (2008) Economic assessment of advanced process control – A survey and framework, *Journal of Process Control*, 18, 2–18	This article provides a survey of industrial attitudes and practices regarding process control benefits, and it contains a thorough literature review.
White, D. (2004) Determining the true economic value of improved plant information, *Hydro. Process.*, December 2004, pg. 53–58	This article discusses uncertainty in benefits predictions.
Wagner, S., A. Al-Ghazzawi, R. McLeod (2004) Using Process Models to Justify real-time Optimization, Aspen World, Orlando,	This presentation addresses benefits for Real-Time Optimization.