Solutions for Tutorial 9
The PID Controller Tuning

9.1 The feedback PID controller has been implemented to control the concentration of the reactant in the reactor effluent from a CSTR. The system is shown in Figure 9.1

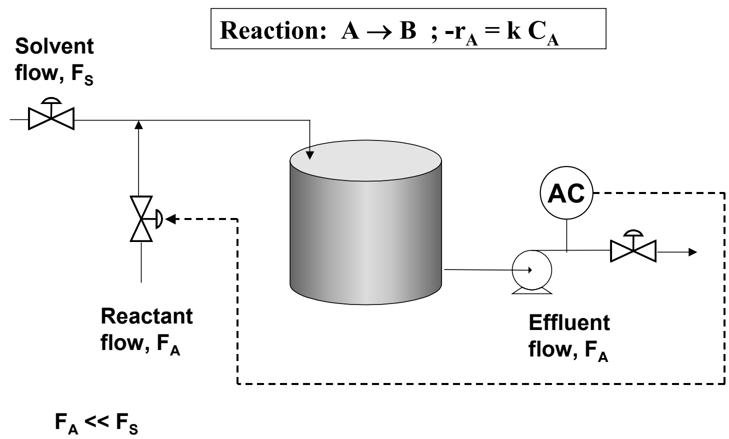


Figure 9.1

a. We have learned that the controller tuning must consider the likely changes in feedback dynamics. Identify several causes for the feedback dynamics to change in this process, and for each cause, explain how the change affects the dynamics.

b. One of the major reasons for feedback control is to compensate for disturbances. Identify several disturbances that would affect the reactant concentration.

a. The dynamic behavior of the model between the pure feed flow rate and the effluent concentration has been derived any times (see textbook Example 3.2 for assumptions and derivation) and is repeated below.

$$V\frac{dC_A}{dt} = F(C_{A0} - C_A) - VkC_A$$

We can determine how changes in operating conditions affect the feedback dynamics, if at all. For example, if we consider just one disturbance (total feed rate) as well as the manipulated variable, we obtain the following models.

$$\tau \frac{dC'_A}{dt} + C'_A = K_F F' + \mathrm{K}_{CA0} C'_{A0} \qquad (3.78)$$

with τ = $V/(F+Vk)$

K_F = $(C_{A0} - C_{As})/(F_s+Vk)$

K_{CA0} = $F/(F+Vk)$

A model for each input can be derived by assuming that the other input is constant (zero deviation) to give the following *two models*, one for each input, in the standard form.

Effect of the disturbance:
$$\tau \frac{dC'_A}{dt} + C'_A = K_{CA0} C'_{A0} \tag{3.79}$$

Effect of the manipulated variable:
$$\tau \frac{dC'_A}{dt} + C'_A = K_F F' \tag{3.80}$$

Clearly, the feedback dynamics depend on

- The total feed rate
- The reactor volume
- The temperature, because of the temperature dependence of the rate constant, k

We can determine the effects from specific changes in sign and magnitude by using the analytical expressions.

b. Many changes will influence the operation of the chemical reactor and affect the effluent concentration. Some examples are given below.

Disturbance	
Feed pressure	A change in pressure changes the flow rate of pure A, even when the valve % open does not change
Solvent pressure	A change in pressure changes the flow rate of solvent, even when the valve % open does not change
Reactor volume	The volume affects the "space time" available for reaction
Feed and solvent temperatures	The reactor temperature affects the rate constant
The solvent valve	A deliberate change in the solvent flow valve opening changes the reactor feed concentration and the total flow rate and "space time"

We must recognize the sources of disturbances so that we can prevent as many as possible and ensure that the feedback control adequately responses to those remaining. For example, we have concluded that we should control the reactor level and temperature. Also, we see the need to control some flow rates to reduce the effects of pressure disturbances. We will use multiple PID controllers to achieve the improvements, so that we must learn the basics of PID control well in Chapters 7-9.

9.2 Let's consider the objectives for the controlled variable, which we must understand to design successful feedback control systems.

a. Several measures of controlled variable "overall" deviation from set point are possible, for example integral of the absolute value of error (IAE) and integral of the error squared (ISE). Compare the two measures.

b. Discuss other measures of controlled variable performance.

a. The two measures are defined in the following equations.

$$IAE = \int_0^\infty | SP - CV | \, dt \qquad\qquad ISE = \int_0^\infty (SP - CV)^2 \, dt$$

Both measures "accumulate" deviations from set point during the transient. Also, they prevent negative and positive values of the errors from canceling each other. They are very useful in summarizing a complete transient response with one number.

- The primary difference is the increased weighting that ISE gives to large errors. Often, large errors (deviations from set point) reduce performance <u>much</u> more than small disturbances; ISE penalizes large disturbances more than small.
- In some cases, the loss of performance is proportional to the deviation from set point; IAE is appropriate for these cases.

The engineer must analyze the process, quality control and economics to select the correct performance measure. Typically, tuning based on IAE or ISE are similar.

b.

<u>Maximum deviation</u>: Perhaps, the most common measure of CV performance, other than IAE or ISE, is the maximum deviation from set point. The maximum deviation must be below a threshold to prevent a hazardous condition (leading to a unit shutdown) or very poor product quality (leading to wasted product).

<u>Rise time</u>: A simple measure of the system's ability to follow a change in command, i.e., set point, is the rise time. In some situations, material produced during a transition between set points cannot be sold; it is waste. In these situations, rise time, and perhaps, settling time, is very important.

<u>Standard deviation</u>: When we consider a long set of data when the plant has been subject to many (nearly random) disturbances, we use the standard deviation of the data from the set point, <u>not</u> from its mean value.

9.3 Let's consider the objectives for the manipulated variable, which we must understand to design successful feedback control systems. Why do we have objectives for the manipulated variables? Give some examples.

The first observation is that we must change the value of the manipulated variable to achieve control. Also, the changes must be rapid enough to return the controlled variable to its set point "quickly". This is required for good CV performance.

However, we should determine limits on the manipulated variable.

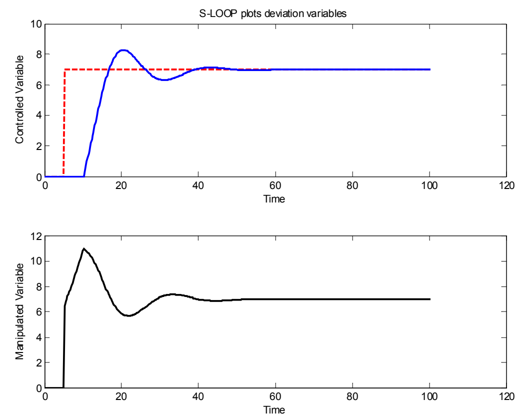
- Very high frequency changes to the manipulated variable will not influence the controlled variable because they will be "filtered" by the process. We should avoid them because they would damage a control valve over a long time.
- Very large, rapid changes are often avoided to prevent damage to equipment. For example, large (fast) changes to a distillation reboiler can cause a high pressure at the bottom of the tower, which can cause a high vapor flow rate and damage to trays.
- A manipulated variable should remain within maximum and minimum values where equipment operates properly. For example, an excessively high fuel rate to a boiler can damage the tubes, and too low a reflux flow rate can lead to poor separation due to dry trays.

9.4 We have collected dynamic data from several different feedback control loops using the PID algorithm. For each, estimate whether the performance is good or not, and when not, diagnose the cause and suggest changes to improve performance. Use the guidelines presented in the textbook for the evaluation; we know that the control performance goals depend on the specific application.

a. The performance appears good.

- The controlled variable achieves zero steady-state offset.
- The dynamic system is stable.
- The process has a dead time of about 5 minutes; therefore, very fast response is not possible.
- The initial change in the MV is nearly equal to the final value, which is good.
- The CV settling time is good.
- The overshoot of the CV past the set point and the MV past its final value are moderate and acceptable.

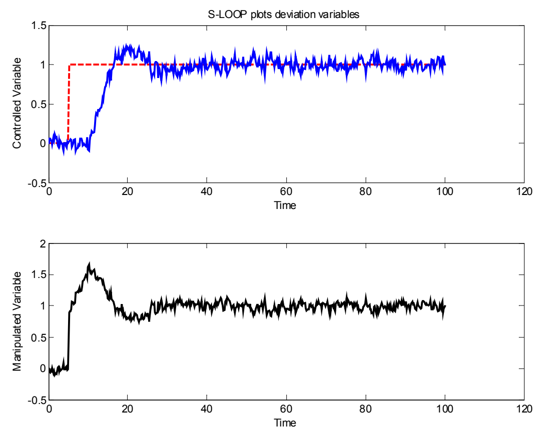
Process:	Controller:
Kp = 1.0 Dead time = 5 Time constant = 5	Kc = 0.90 TI = 7.0



b. The performance appears good.

- The controlled variable achieves zero steady-state offset.
- The dynamic system is stable.
- The process has a dead time of about 5 minutes; therefore, very fast response is not possible.
- The initial change in the MV is nearly equal to the final value, which is good.
- The CV settling time is good.
- The overshoot of the CV past the set point and the MV past its final value are moderate and acceptable.

Process:	Controller:
Kp = 1.0 Dead time = 5 Time constant = 5	Kc = 0.90 TI = 7.0

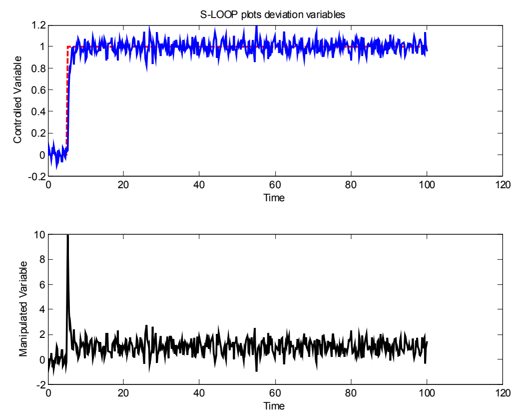


Note that the only difference between cases (a) and (b) is high frequency variation. This could be due to sensor noise or high frequency process disturbances. They are much faster than the feedback dynamics and cannot be controlled.

c. The performance appears <u>questionable</u>.

- The controlled variable achieves zero steady-state offset.
- The dynamic system is stable.
- The process has no dead time; therefore, very fast response is possible.
- The initial change in the MV exceeds its final value by a **factor of about 9**.
- The CV settling time is good.
- The overshoot of the CV past the set point is very small, and the rise time is extremely fast.

Process:	Controller:
Kp = 1.0	Kc = 10.0
Dead time = 0	TI = 7.0
Time constant = 5	

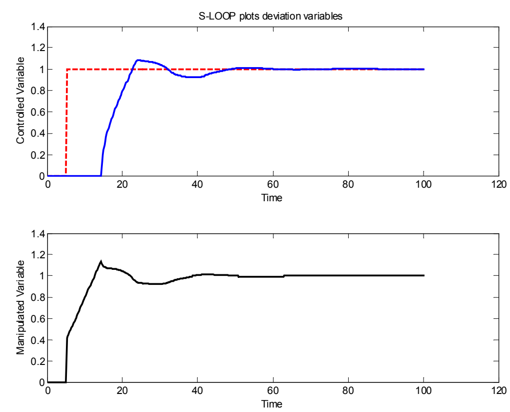


The large overshoot in the manipulated variable would generally not be acceptable. However, if the manipulated variable were cooling water, this might be OK.

d. The performance appears <u>good</u> for this difficult process

- The controlled variable achieves zero steady-state offset.
- The dynamic system is stable.
- The process has 9 minutes of dead time; therefore, very fast response is not possible.
- The initial change in the MV is small, about 40% of its final value, but this is expected because aggressive control of a process with a large fraction dead time is not possible with feedback.
- The CV rise time and settling time are long because of the long process dead time.
- The overshoot of the CV past the set point is very small.

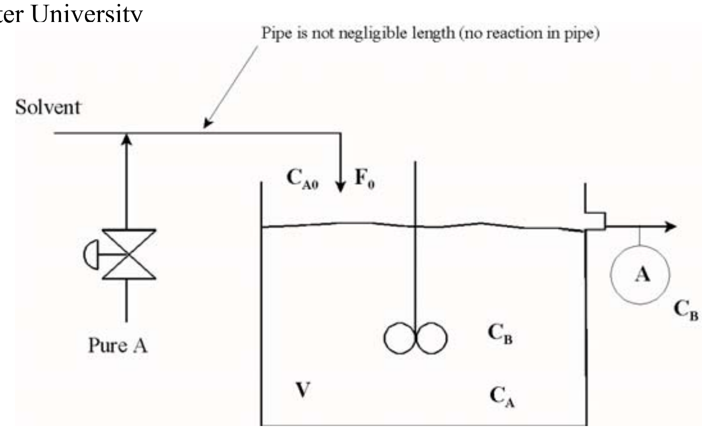
Process:	Controller:
Kp = 1.0	Kc = 0.40
Dead time = 9	TI = 5.0
Time constant = 1	



This process has a long dead time and is difficult to control. While the control performance is much worse the case (a), it is not because of a problem with the controller.

If we want to improve the performance, we should use our engineering skills to shorten the dead time.

Alternatively, we could evaluate the use of new methods (cascade and feedforward) that are introduced later in the course. Something to look forward to!



9.5 Your goal is to control the concentration of B in the reactor effluent by adjusting the pure A control valve.

Determine the tuning for the proposed PID controller based on the data in Figure 9.5, with concentrations in mole/m^3 and time in minutes. Show all calculations and briefly explain decisions you make.

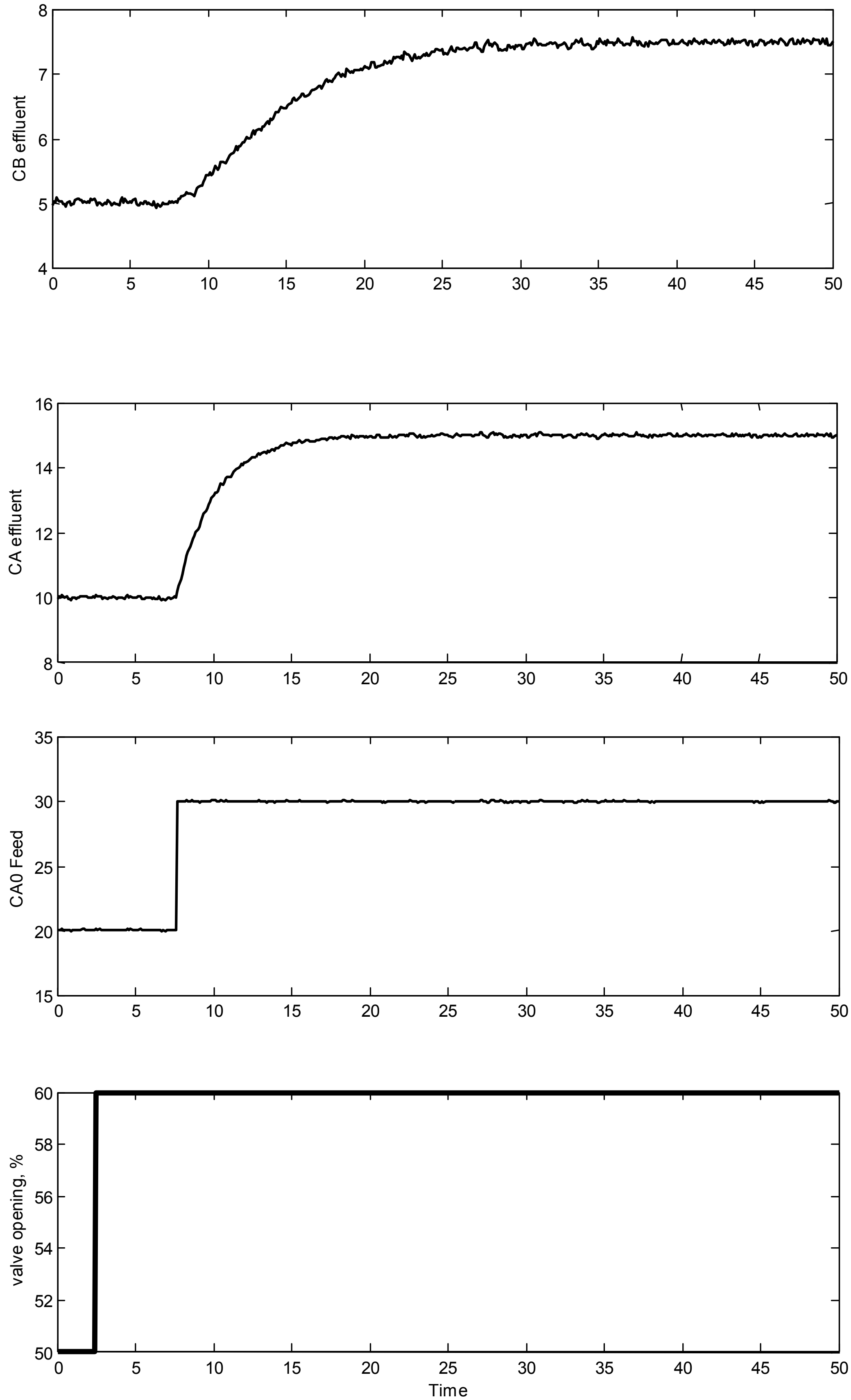
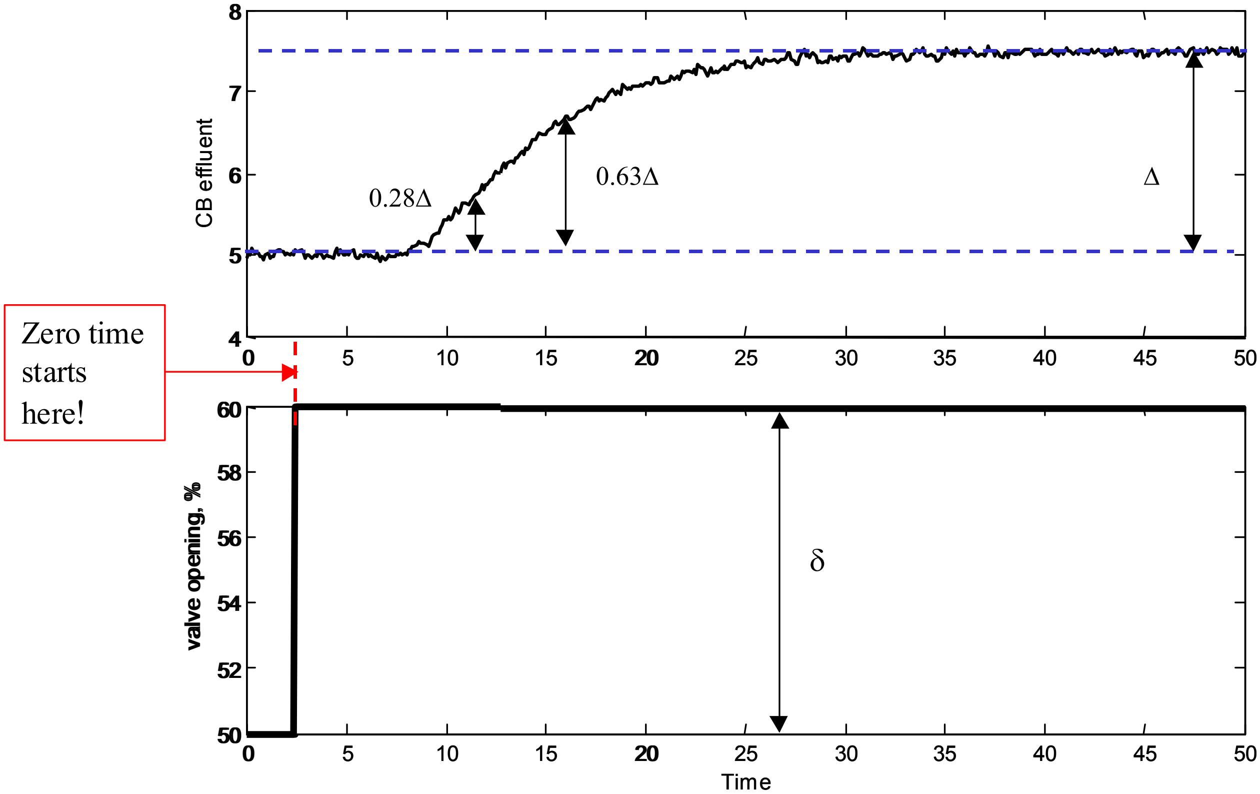


Figure 9.5. Data from process reaction curve experiment.

The procedure is shown on the following graph. Note that we do ***not*** estimate the models for the intermediate variables (CA0 and CA), because we need the dynamics between the final element (valve) and the measured controlled variable (CB).



τ = 1.5 (t63% - t28%) = 1.5 (13.4 – 8.56) = 7.2 minutes
θ = t63% - τ = 13.4 – 7.2 = 6.2 minutes
Kp = Δ/δ = 2.5 mole/m3 / 10% open = 0.25 (mole/m3)/%open

PID tuning from the Charts, Figure 9.5 a-c.

$\theta/(\theta+\tau)$ = 6.2/(13.4) = 0.47

KcKp = 0.9 Kc = 0.9/0.25 = 3.6 %open/ (mole/m3)

TI/($\theta+\tau$) = 0.67 TI = 0.67 (13.4) = 9.0 min

Td/($\theta+\tau$) = 0.06 Td = 0.06 (13.4) = 0.80 min

9.6 We know that a chemical process has many variables to control. How can we achieve good control by using the PID algorithm for feedback, since it is limited to a single measured controlled variable and a single manipulated variable?

It might help if you considered a process example. The CSTR is shown in Figure 9.6. We want to design controls for the four measured variables.

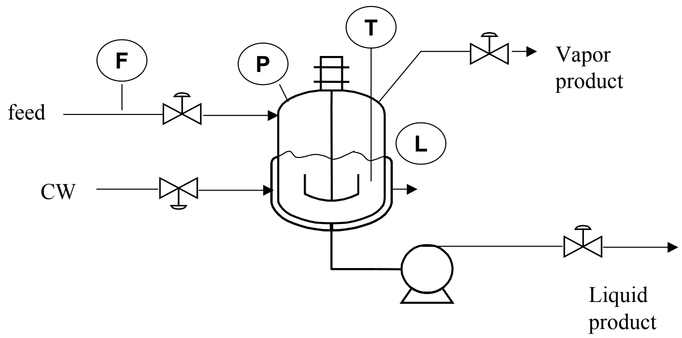
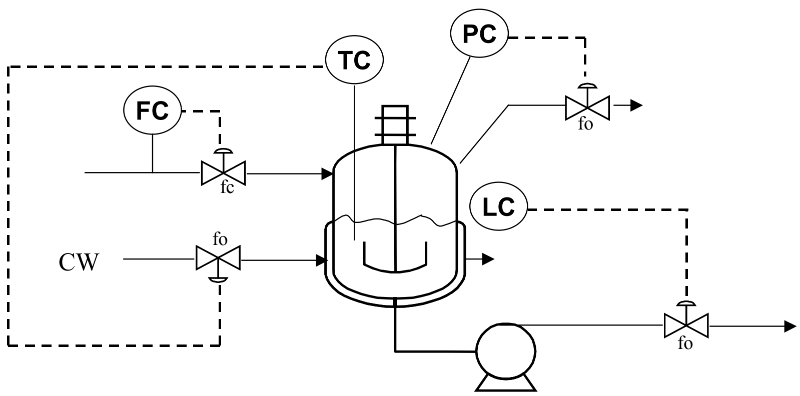


Figure 9.6

The most widely used approach is to control each CV with an individual PID controller, which adjusts an individual manipulated variable, i.e., valve. Thus, each controller has one CV and one MV; we refer to the choice of which MV to adjust to control a CV as loop pairing. We term a design that employs several PID controllers as "multiloop control".

Recall that each controller is completely independent from the others, and no communication is shared among the controllers. We recognize immediately that these controllers will "interact", so the possibility exists for poor (or improved) performance because of the multiple loops. The topic of loop pairing will be covered later in the course. Now, we are concentrating on designing one feedback loop and making it perform well.

A possible multiloop design for the example in this question is shown in the following figure. Each controller (FC, LC, etc.) is an individual PID controller using one measured value and adjusting one valve.



As an exercise, you should discuss this design and determine whether it "makes sense". We will learn a design procedure later.