Traditional Advanced Control Strategies

Traditional Advanced Control Strategies

- High level application of PID controller
    - Time delay Compensator
    - Cascade Control
    - Split range Control
- Feedforward and Ratio Control

    - Selective control
    - Override Control
    - Valve Position Control
    - Variable Structure Control

Cascade Control

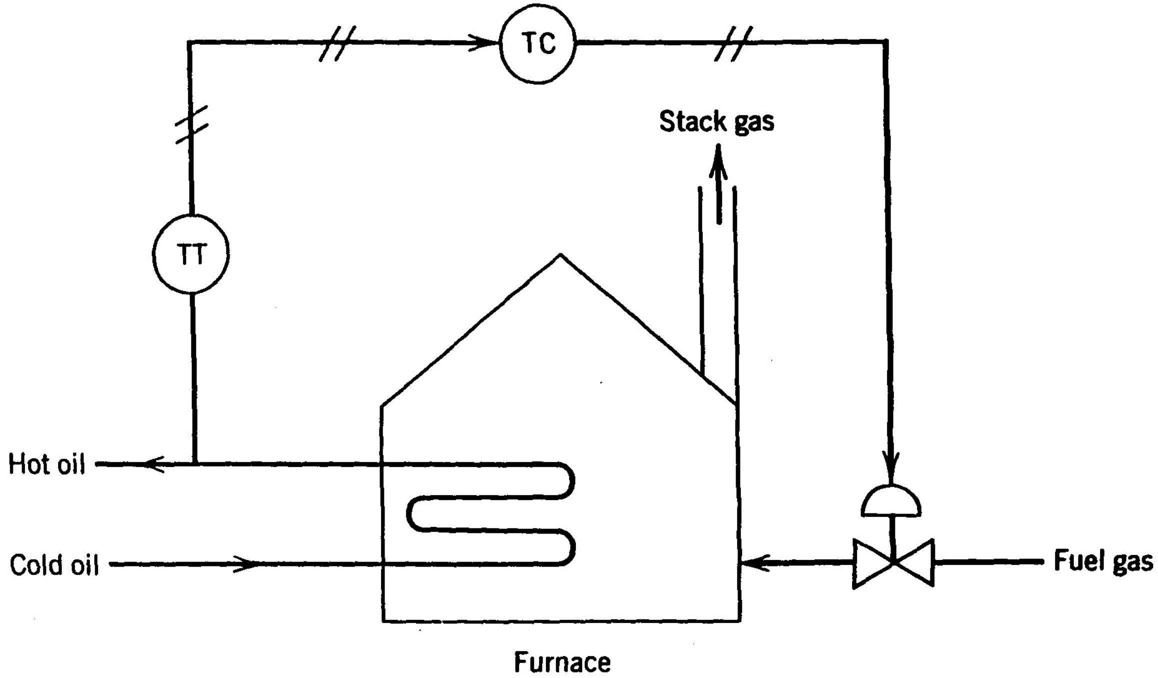


Figure 18.2. A furnace temperature control that uses conventional feedback control.

- Hot oil Temperature is controlled by fuel gas flow rate
- Fuel gas obtained from other process units and so pressure of the fuel gas fluctuates
- Flow rate of fuel gas will be different for the same valve opening

Cascade Control

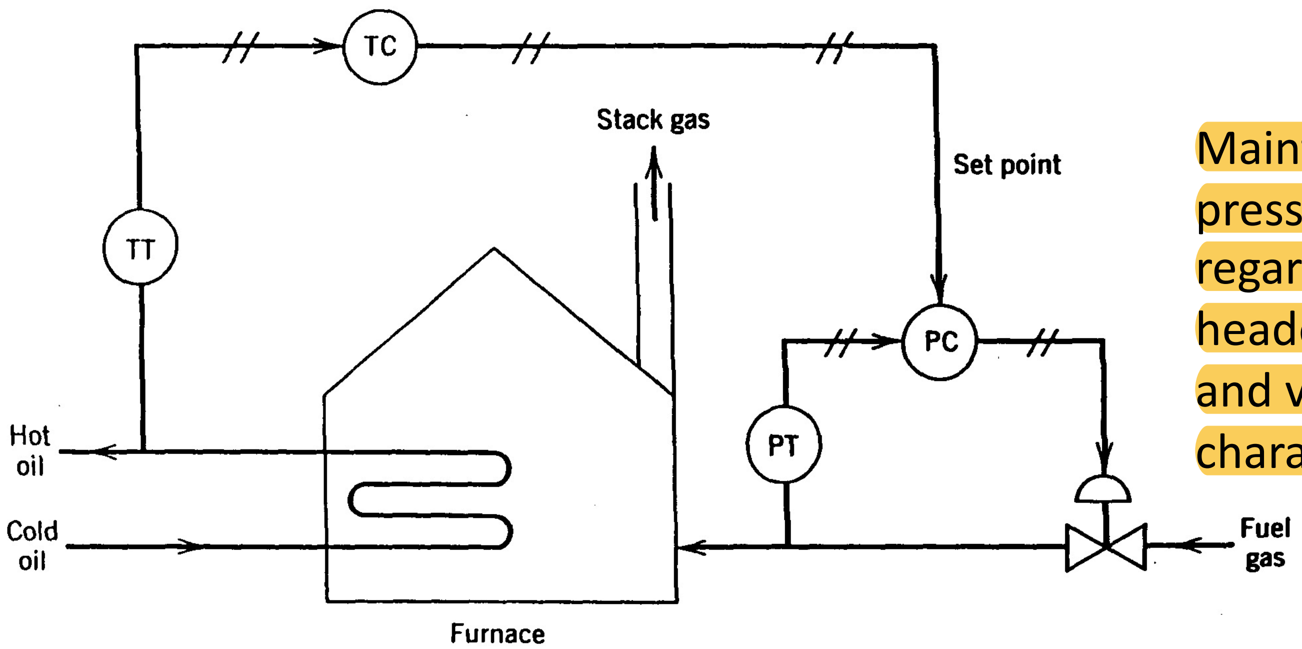


Figure 18.3. A furnace temperature control using cascade control.

Maintain fuel gas pressure / flow regardless of header pressure and valve characteristics

- Two FB controllers but only a single control valve (or other -final control element).
- Output signal of the "master" controller is the set point for “slave" controller.
- Two FB control loops are "nested" with the "slave" (or "secondary") control loop inside the "master" (or "primary") control loop.

Cascade Control

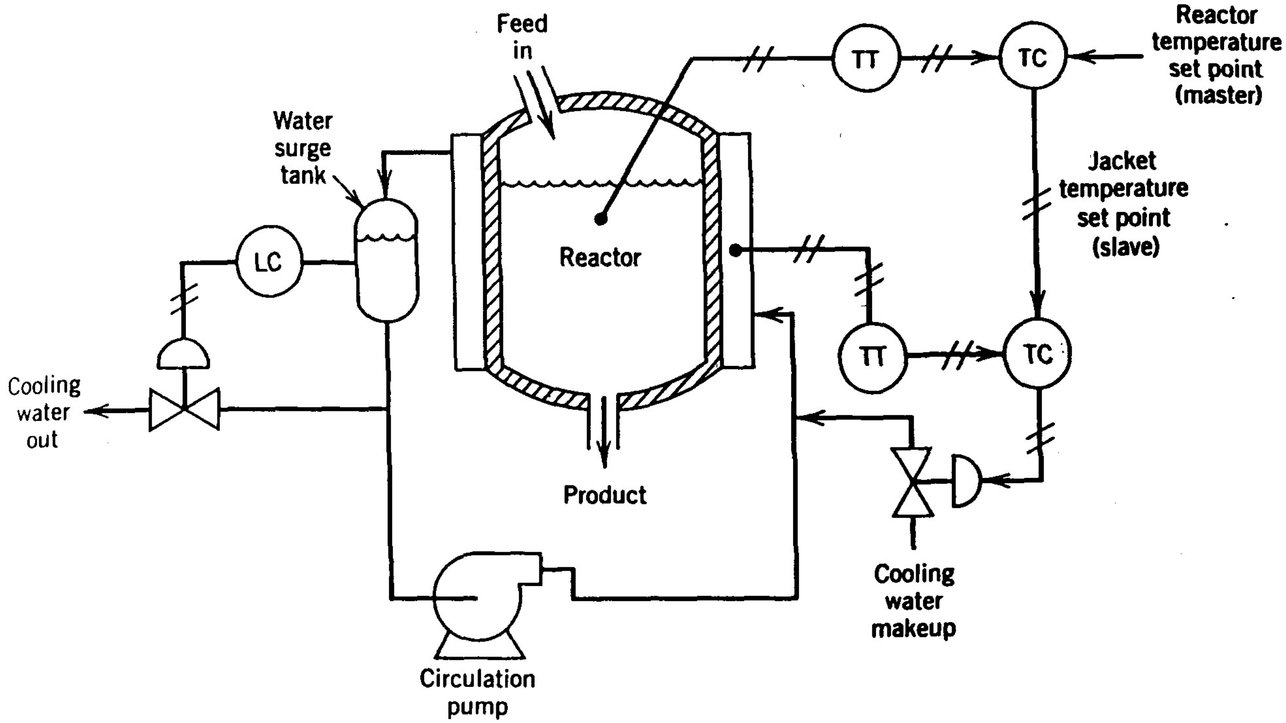


Figure 18.1. Cascade control of an exothermic chemical reactor.

- Jacket Temperature can be controlled by adjusting cooling water flow rate
- If the cooling water inlet temperature changes, the jacket temperature will change with same flow rate
- For rise in Reactor temperature, hot CW will be replaced by cold CW

Cascade Control : Block Diagram

Hot oil Temp y
Cold oil Temp d_1
Fuel gas flow u
Header press d_2
Valve op u_1

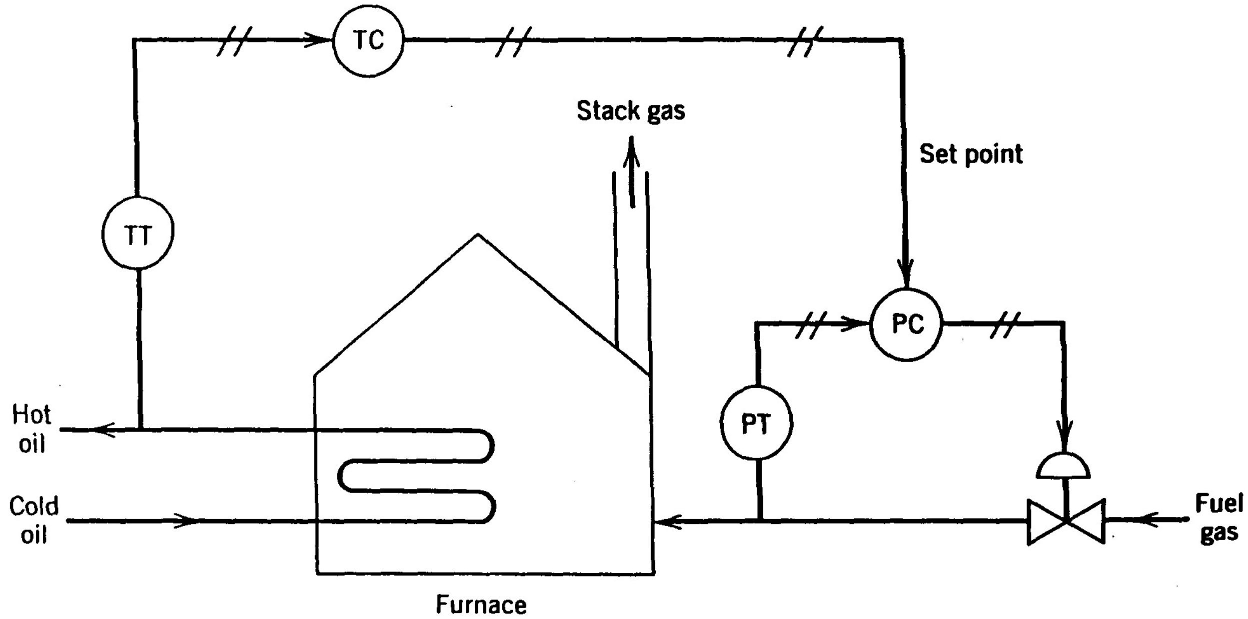
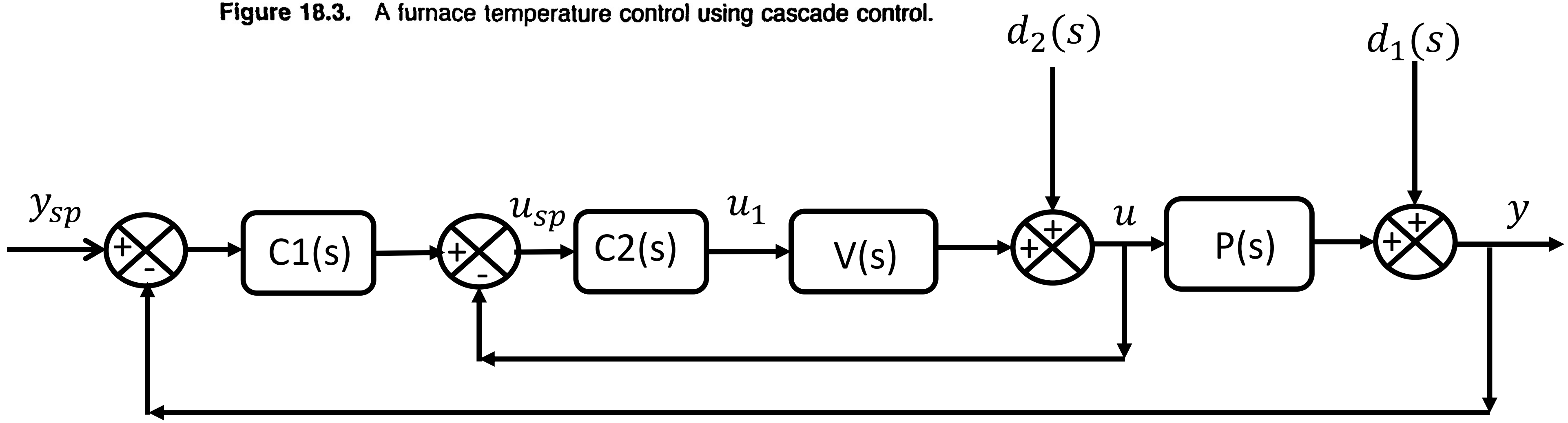
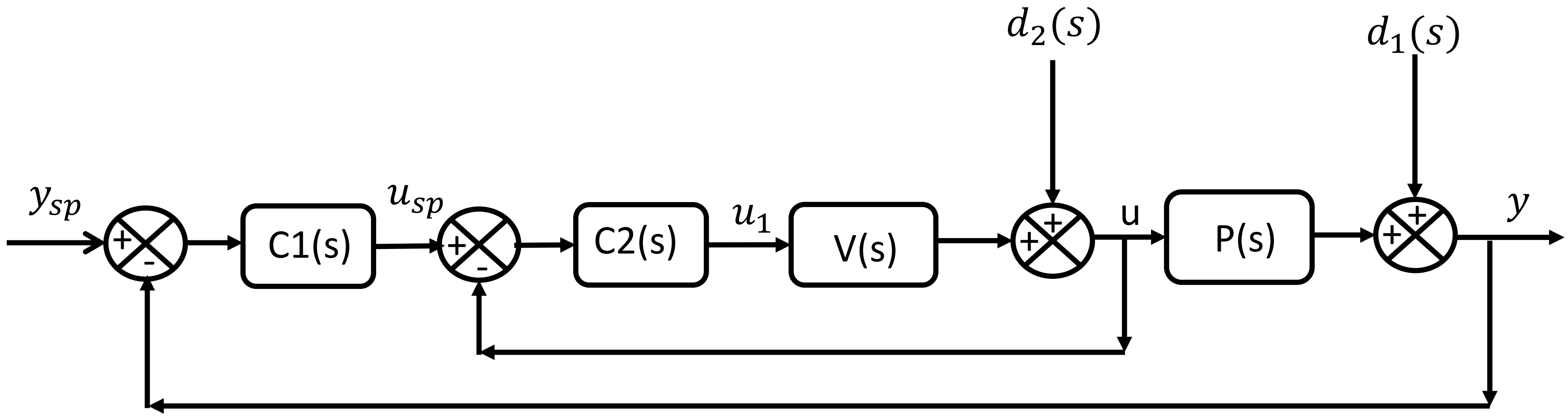


Figure 18.3. A furnace temperature control using cascade control.



Cascade Control : Feedback Equation



$$u(s) = \frac{C_2(s)V(s)}{1+C_2(s)V(s)} u_{sp}(s) + \frac{1}{1+C_2(s)V(s)} d_2(s)$$

$$y(s) = \frac{C_2(s)V(s)P(s)}{1+C_2(s)V(s)} u_{sp}(s) + \frac{P(s)}{1+C_2(s)V(s)} d_2(s) + d_1(s)$$

$$= \frac{C_1(s)C_2(s)V(s)P(s)}{H(s)} y_{sp}(s) + \frac{P(s)}{H(s)} d_2(s) + \frac{1 + C_2(s)V(s)}{H(s)} d_1(s)$$

$$H(s) = 1 + C_2(s)V(s) + C_1(s)C_2(s)V(s)P(s)$$

Cascade Control

Advantages

– Reject the disturbance in the slave loop before it affects the main process variables.

– Linearize the slave process

– Improve the dynamics of the slave process

Implementation

– The slave process is at least 3 times as faster as master loop in terms of response time.

– The slave loop may not need to be controlled exactly at set point. (I- mode is not necessary in many cases.) It needs to be controlled to provide fast action on disturbance and set point change. (P-mode may be suffice in most cases.)

– The slave loop should be tuned first while the master in manual. Then tune the master while the slave is in automatic mode.

– If the slave is retuned for some reasons, the master should be retuned.

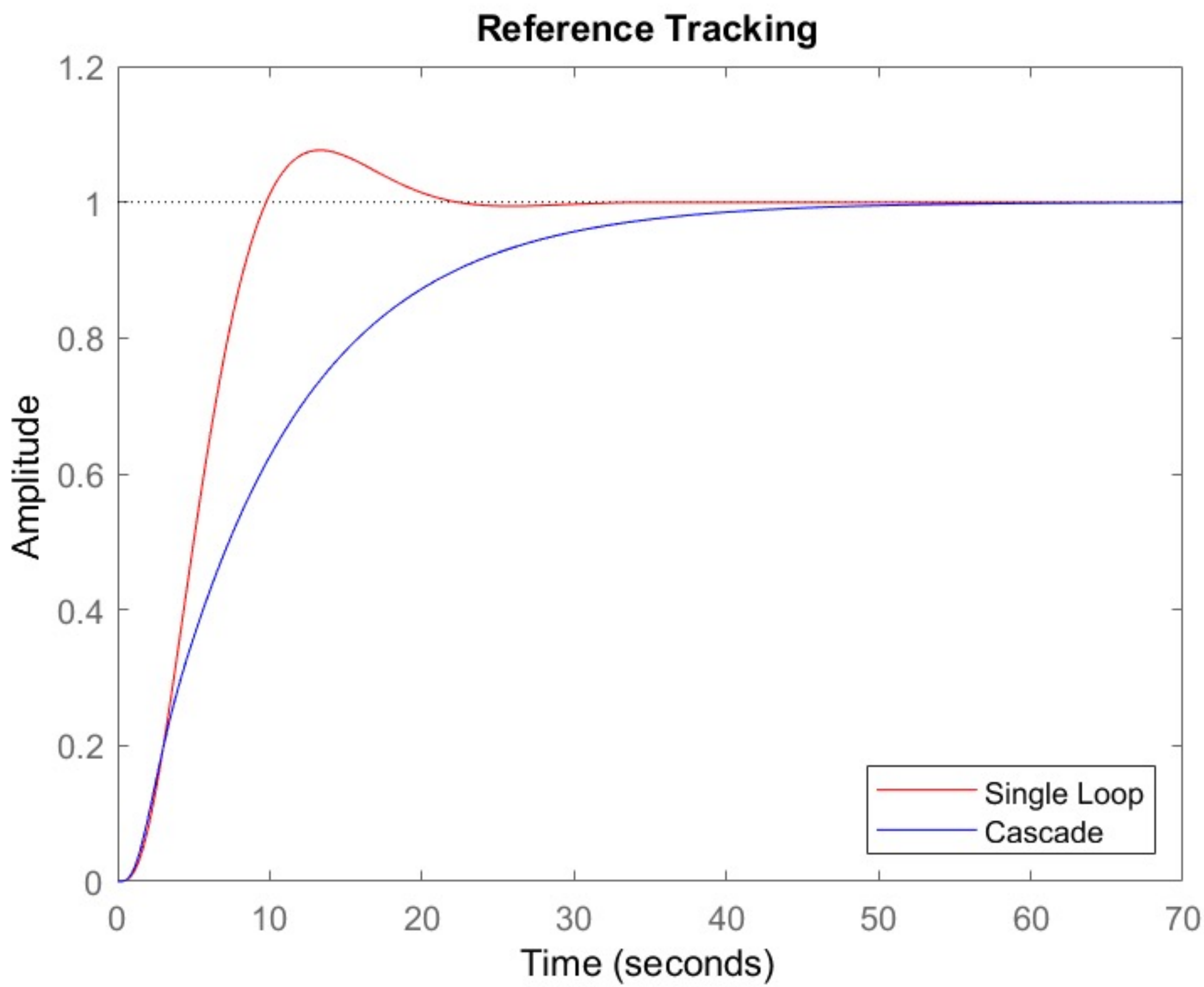
– The master can be transferred to ‘auto’ after the slave becomes ‘auto’.

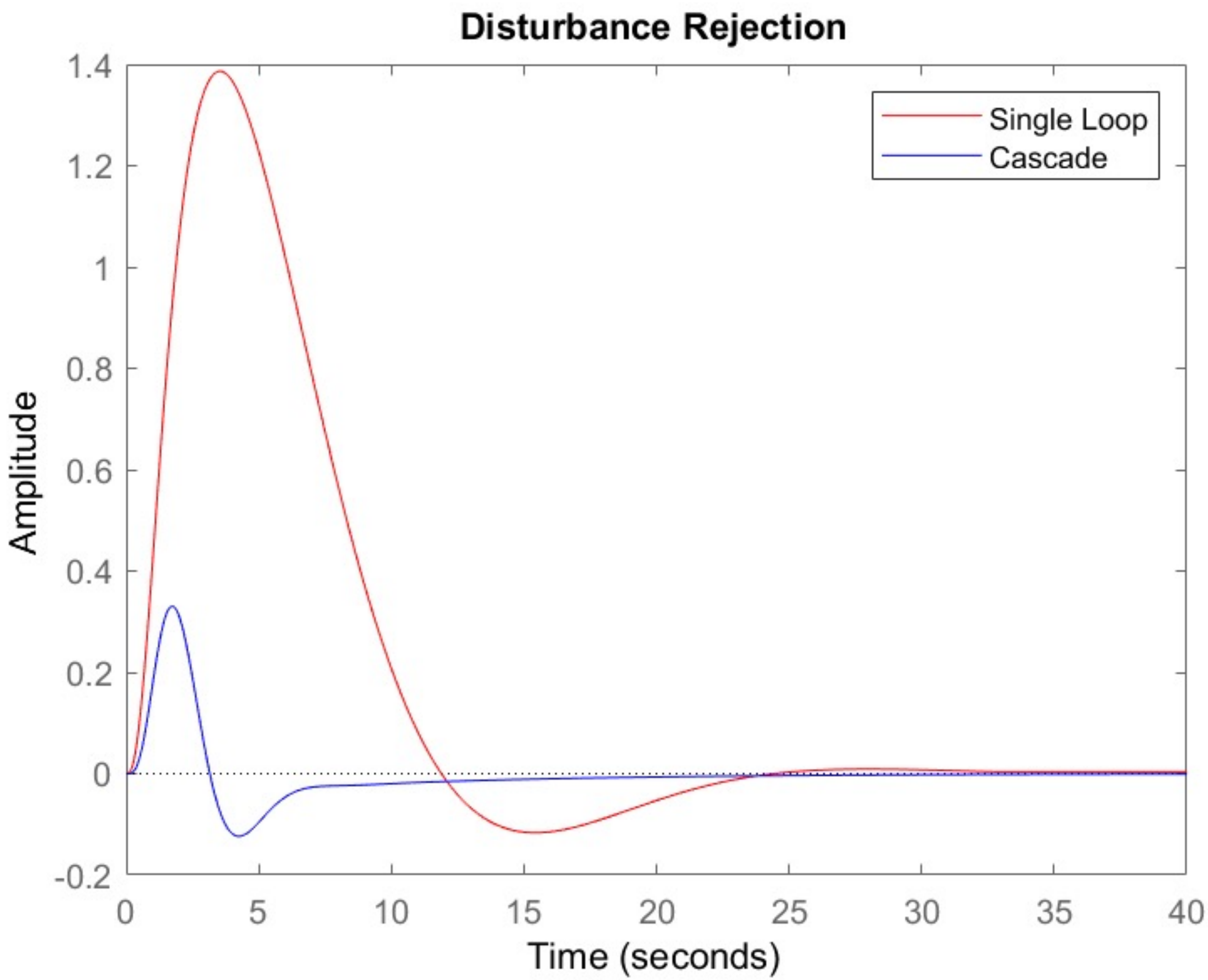
Furnace Control : Matlab Implementation

$$y(s) = \frac{10}{(s+1)(s+2)(s+3)}u(s);$$ y is Temperature, u is flow

$$u(s) = \frac{2.2}{2s+1}u_1(s) + \frac{3}{s+2}d_2(s)$$; u_1 is CV opening and d_2, header pres

```
s=tf('s');
P=10/(s+1)/(s+2)/(s+3);
V=2.2/(2*s+1);
VD=3/(s+2);
% Tune PI controller for target
%bandwidth is 0.2 rad/s and
%phase Margin 60
C = pidtune(P*V,pidstd(1,1),0.2);
sys1 = feedback(P*V*C,1);
sys1.Name = 'Single Loop';
% target bandwidth 2 rad/s and
% phase margin 60 for Inner controller
C2 = pidtune(V,pidstd(1,1),2);
clsys = feedback(V*C2,1);
sys2 = feedback(clsys*P*C,1);
sys2.Name = 'Cascade';
figure;
step(sys1,'r',sys2,'b')
legend('show','location','southeast')
title('Reference Tracking')
% Disturbance rejection
sys1d1 = feedback(P,V*C);
sys1d1.Name = 'Single Loop';
sys2d1 = P*VD/(1+V*C2+V*P*C*C2);
sys2d1.Name = 'Cascade';
figure;
step(sys1d1,'r',sys2d1,'b')
legend('show')
title('Disturbance Rejection')
```

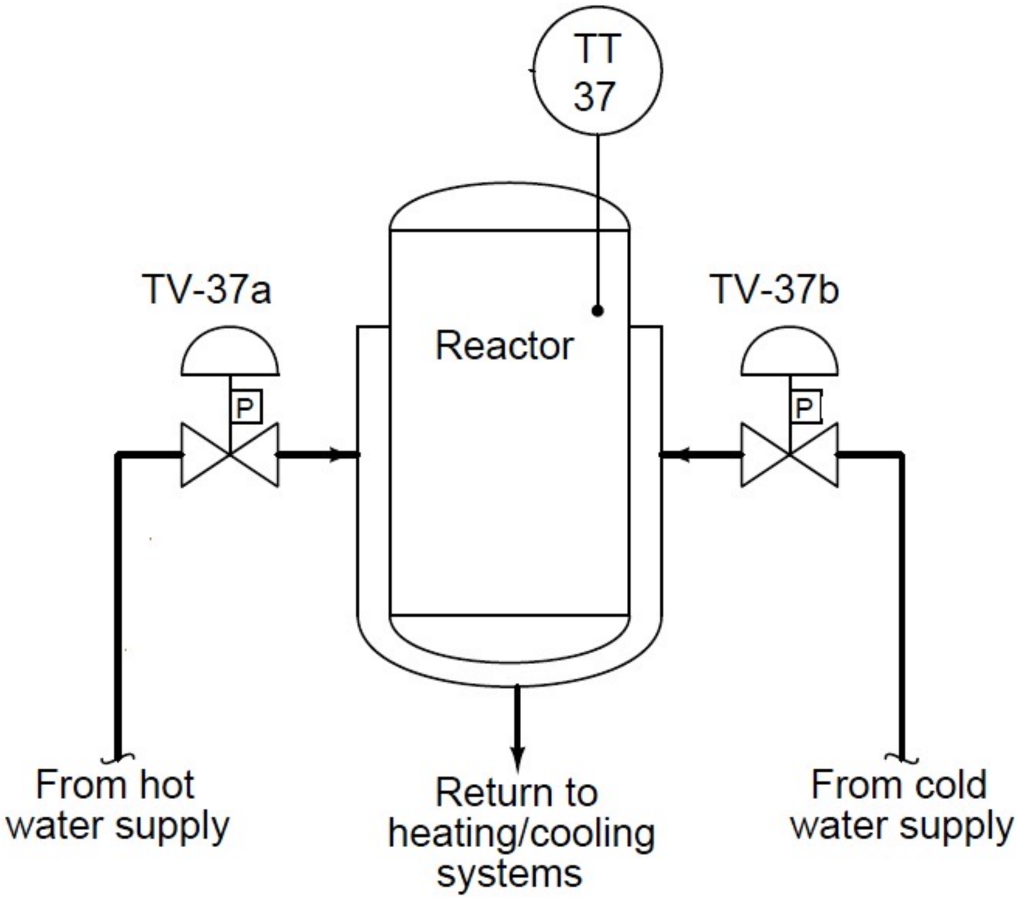

Reference Tracking
Amplitude
Time (seconds)
Single Loop
Cascade
0
0.2
0.4
0.6
0.8
1
1.2
0
10
20
30
40
50
60
70


Disturbance Rejection
Single Loop
Cascade
Amplitude
Time (seconds)
1.4
1.2
1
0.8
0.6
0.4
0.2
0
-0.2
0
5
10
15
20
25
30
35
40

Split Range Control

Control of Batch Reactor

- Ensures maximum production of desired product
- Suppress the production of undesired product
- Desired temperature should Rise at 5^0C per min upto certain time and then decrease at the same rate



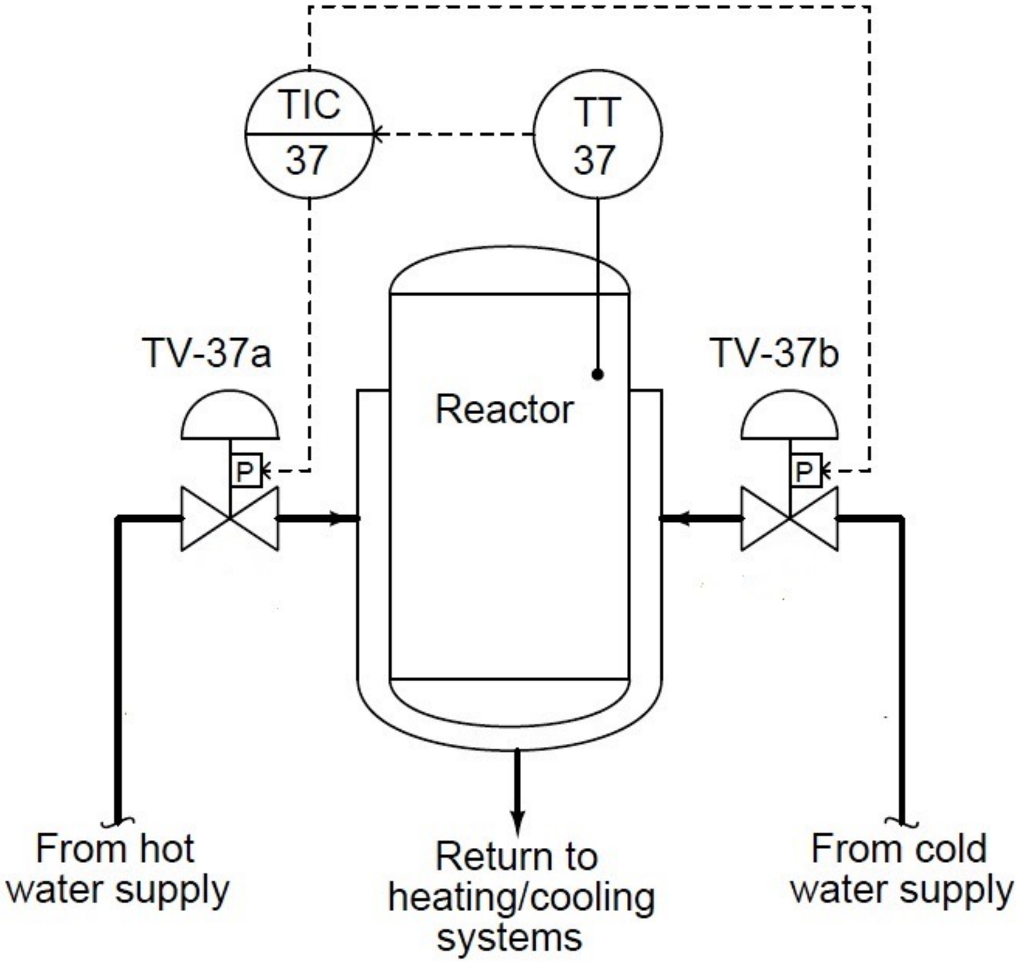
What should be the control strategy ??

Split Range Control

Control of Batch Reactor

- Feedback control of reactor temperature
- Controller output sent to both control valves

How the control valves will operate based on same controller signal?



Split Range Control

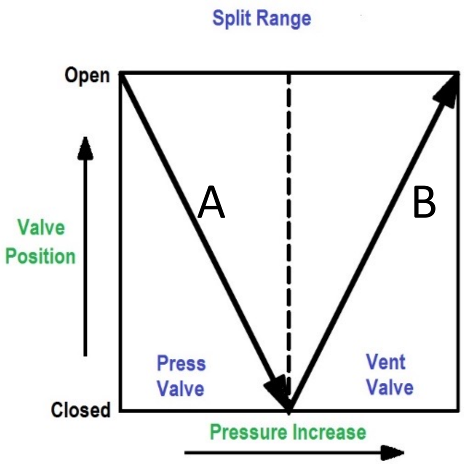
- Split Range Control Loop is used, where there are several manipulated variables, but a single output variable.
- The coordination among different manipulated variables is carried out by using Split Range Control.

The split range control loop can be implemented in three different ways.

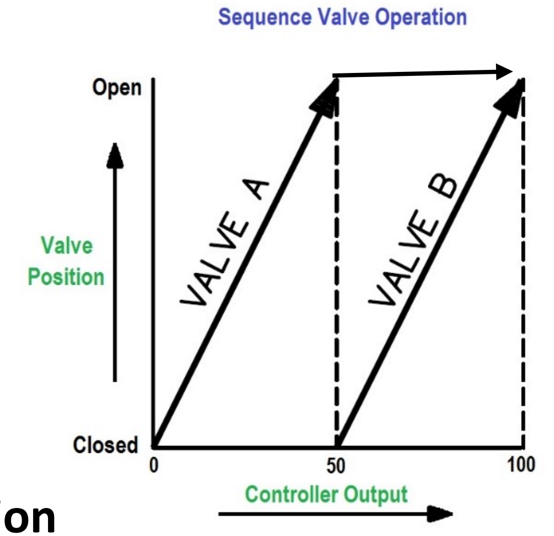
- Split Range Control
- Sequence Control
- Opposite Action Control

Split Range Control

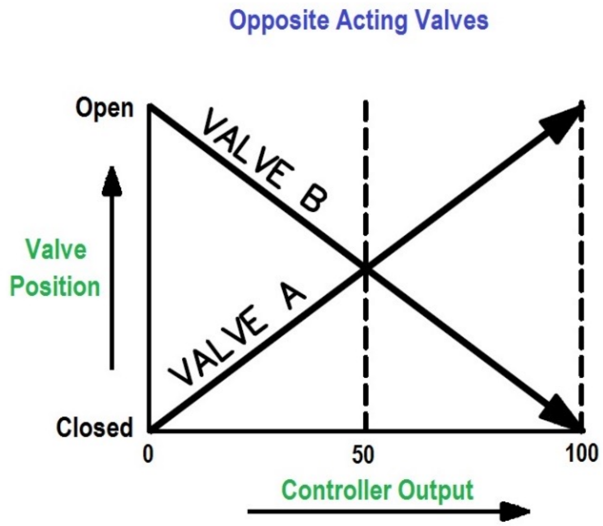
Split Range



Sequence Operation



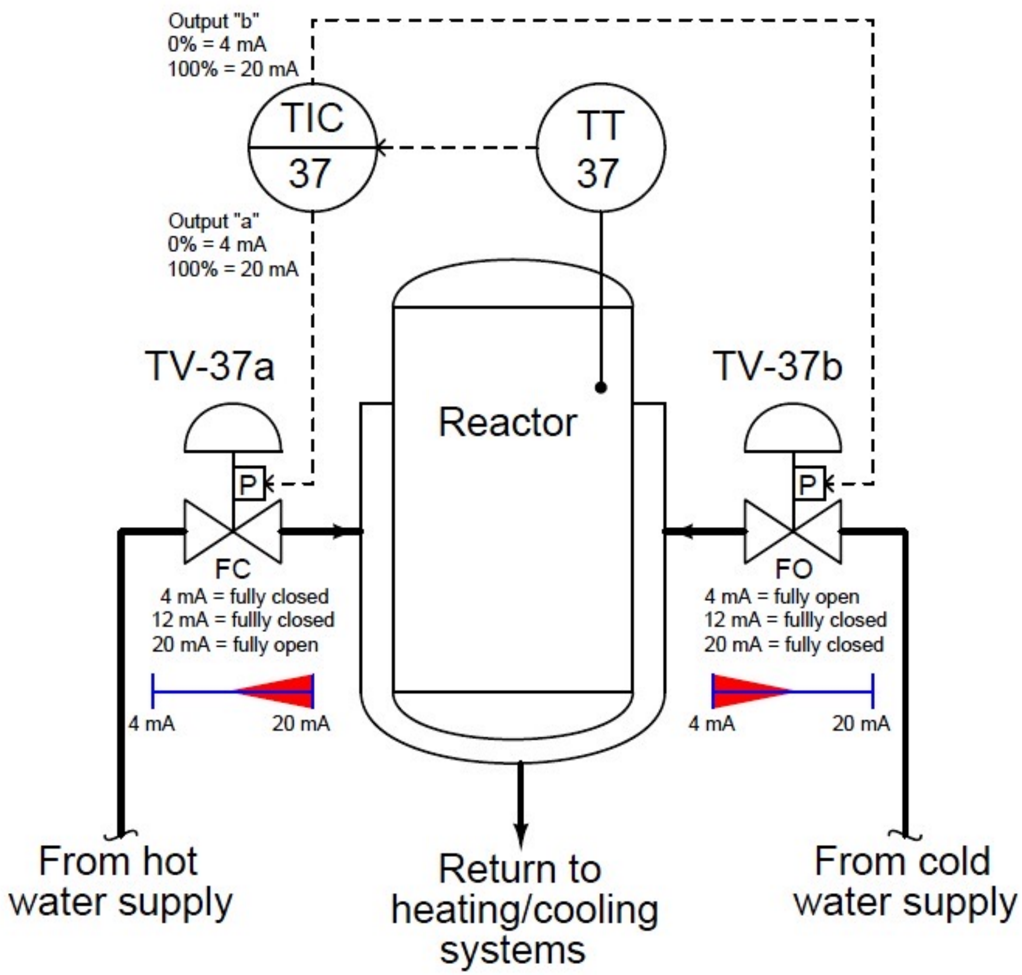
Opposite Action



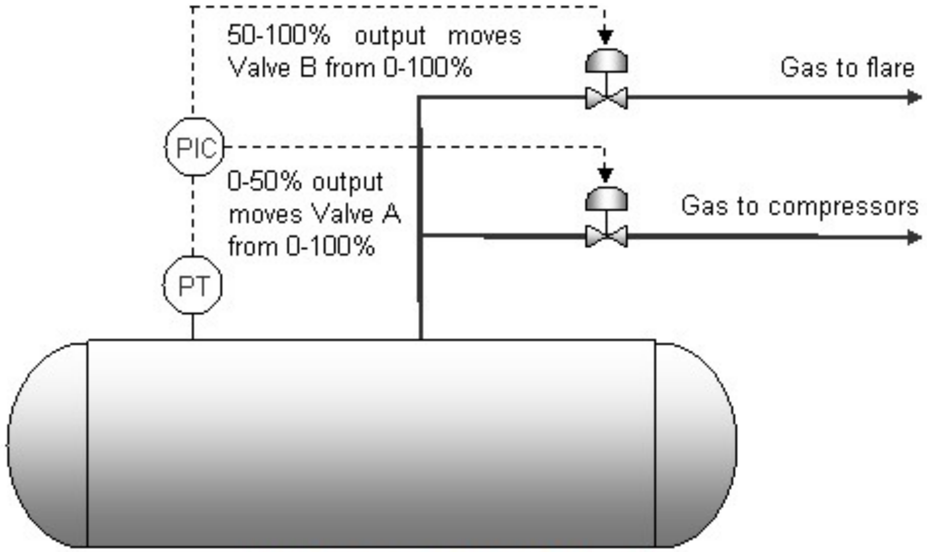
Split Range Control

Control of Batch Reactor

- Controller output (range 4-20 mA) sent to both control valves
- As temperature increases, controller output decreases.
- Cold water supply should increase.
- Hot water supply should decrease
- Cold side CV work between 4 to 12 mA
- Hot side CV work between 12 to 20 mA



Split Range Control

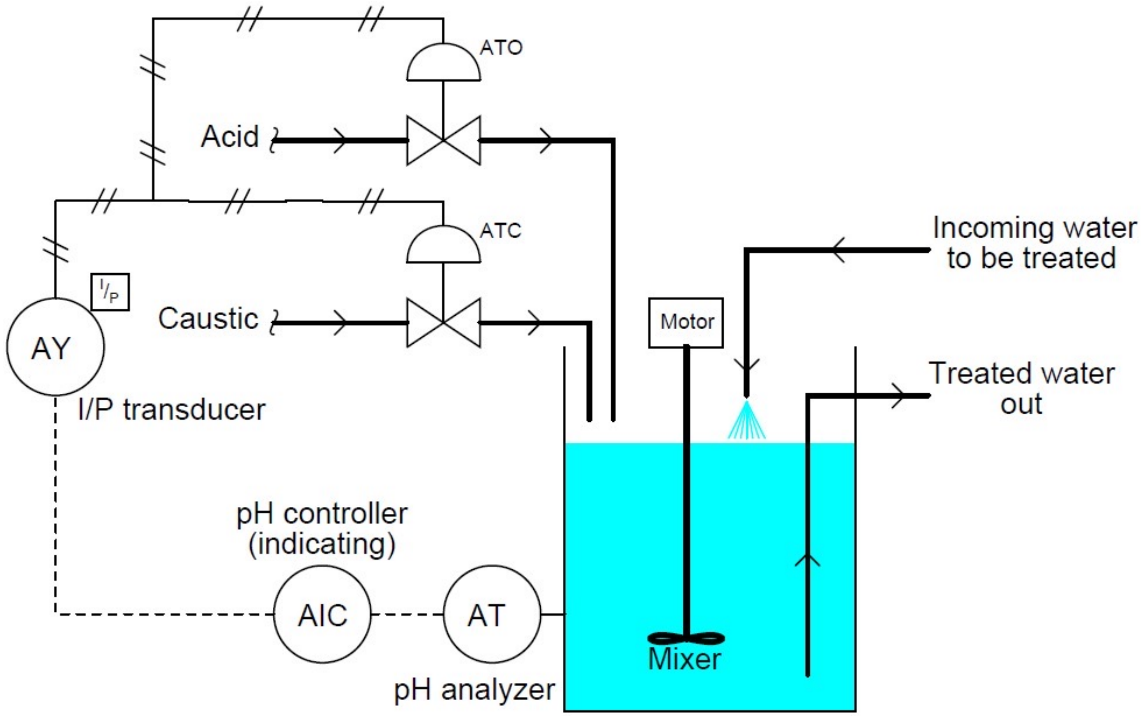


Split range control scheme

Control of overhead drum in a separator

- For normal operation, gas is sent to header
- Gas is sent to flare when drum pressure increases above normal operating range

Split Range Control



Control of pH neutralization reactor

Feedforward & Ratio Control

Feedback Control

Advantages of Feedback Control

1. Corrective action occurs as soon as the controlled variable deviates from the set point, regardless of the source and type of disturbance.

2. Feedback control requires minimal knowledge about the process to be controlled; in particular, a mathematical model of the process is *not* required, although it can be very useful for control system design.

3. PID controller is both versatile and robust. If process conditions change, retuning the controller usually produces satisfactory control.

Feedback Control

Disadvantages of Feedback Control

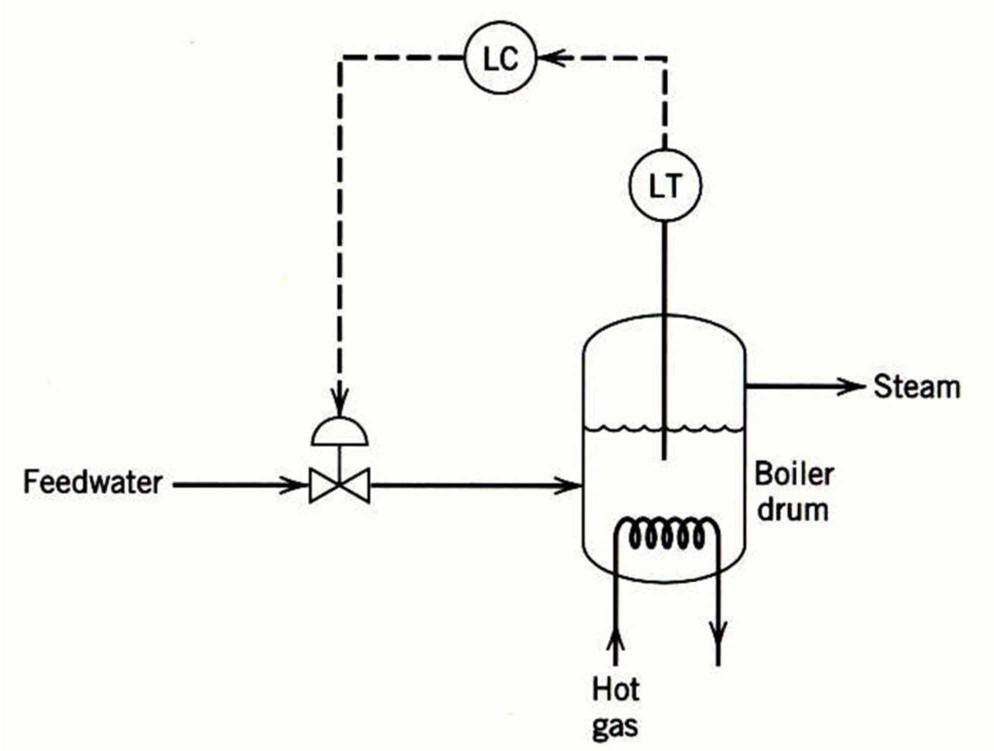
1. Corrective action is taken only after a deviation in the controlled variable occurs. Thus, *perfect control*, where the controlled variable does not deviate from the set point during disturbance or set-point changes, is theoretically impossible.
2. Feedback control does not provide predictive control action to compensate for the effects of known or measurable disturbances.
3. It may not be satisfactory for processes with large time constants and/or long time delays. If large and frequent disturbances occur, the process may operate continuously in a transient state and never attain the desired steady state.
4. In some situations, the controlled variable cannot be measured on-line, and, consequently, feedback control is not feasible.

FeedForward Control

The basic concept of feedforward control is to measure important disturbance variables and take corrective action before they upset the process.

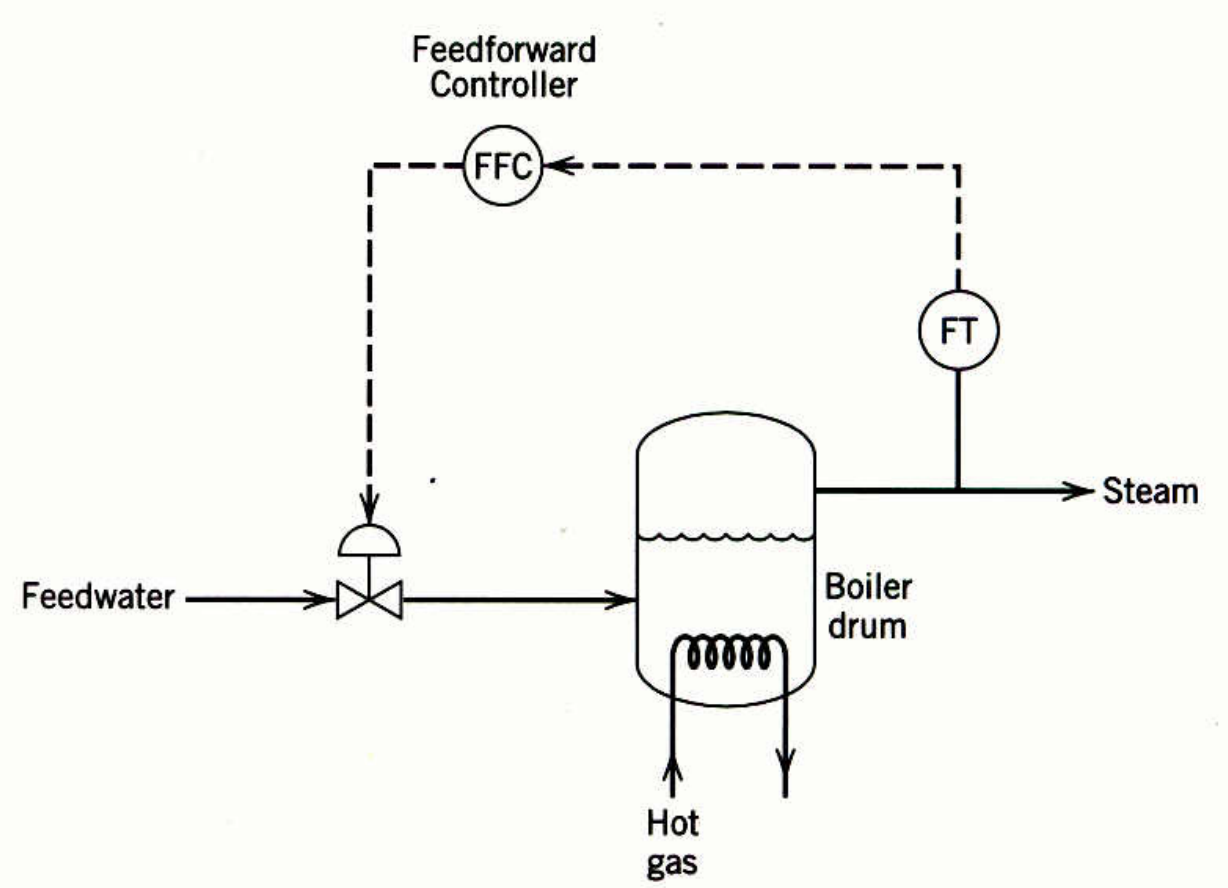
Rapid disturbance changes can occur as a result of steam demands made by downstream processing units.

This control system tends to be quite sensitive to rapid changes in the disturbance variable, steam flow rate, as a result of the small liquid capacity of the boiler drum.



Feedforward Control

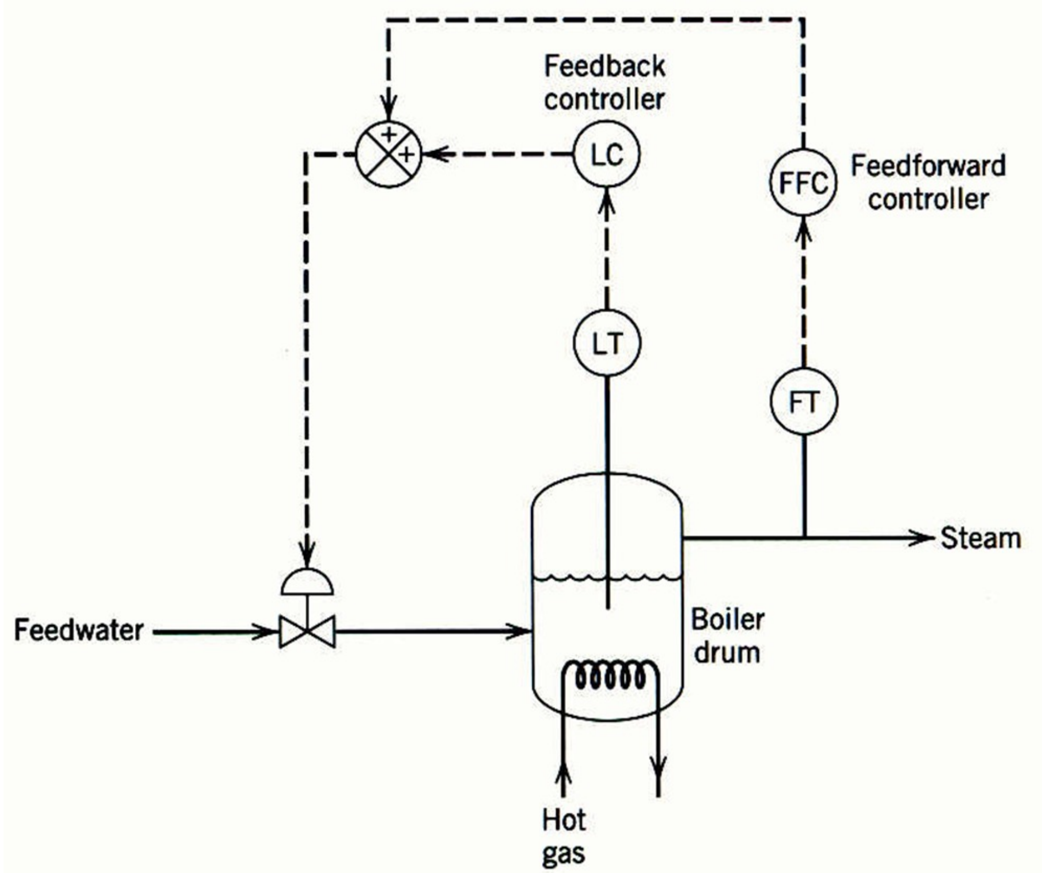
The feedforward control scheme can provide better control of the liquid level. Here the steam flow rate is measured, and the feedforward controller adjusts the feed water flow rate.



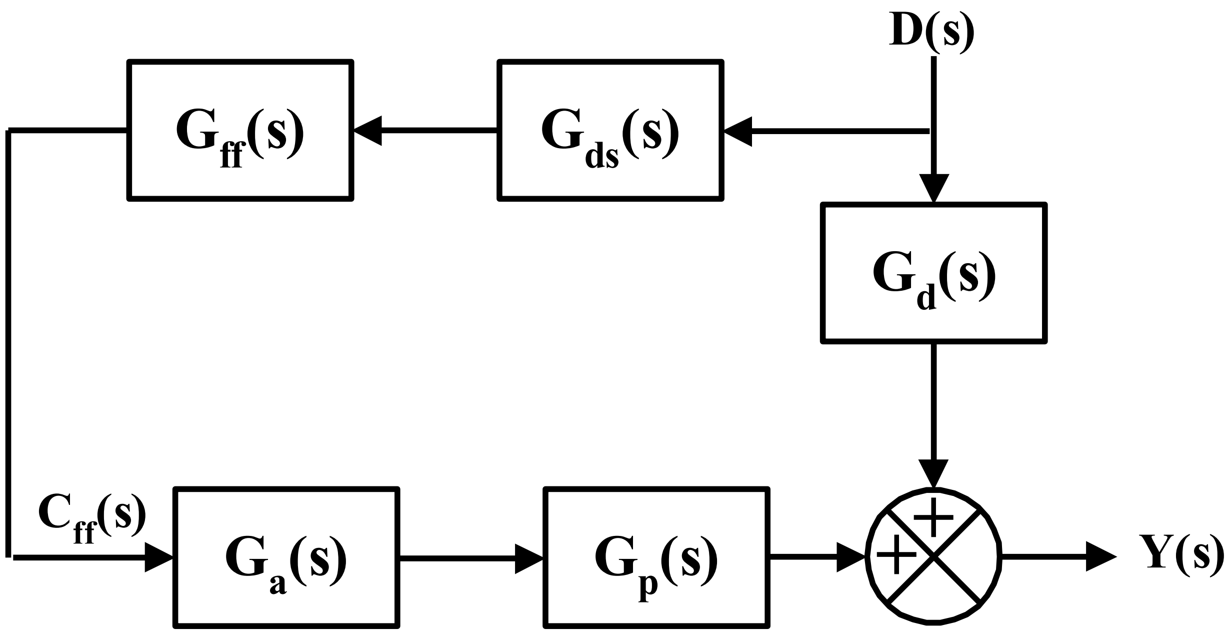
Feedforward-Feedback Control

In practical applications, feedforward control is normally used in combination with feedback control.

Feedforward control is used to reduce the effects of measurable disturbances, while *feedback trim* compensates for inaccuracies in the process model, measurement error, and unmeasured disturbances.



Derivation of FF Controller



$$Y(s) = D(s)G_{ds}(s)G_{ff}(s)G_a(s)G_p(s) + D(s)G_d(s) = 0$$

Solving for $G_{ff}(s)$

$$G_{ff}(s) = \frac{-G_d(s)}{G_{ds}(s)G_a(s)G_p(s)}$$

Lead/Lag Element for Implementing FF Control

$$G_{ds}(s)\,G_a(s)\,G_p(s) = \frac{K_p\, e^{-\theta_p s}}{\tau_p s + 1}$$

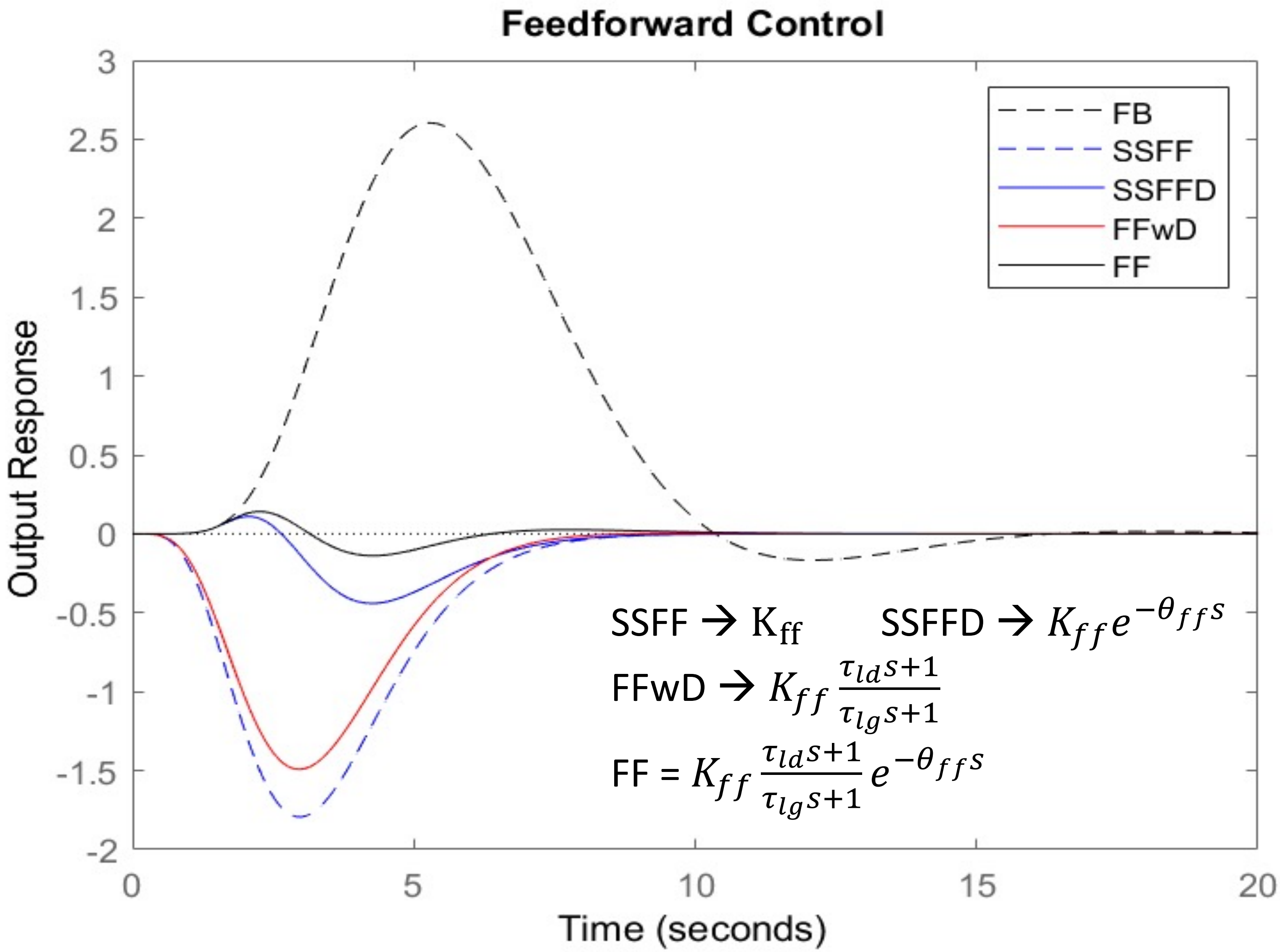
$$G_d(s) = \frac{K_d\, e^{-\theta_d s}}{\tau_d s + 1}$$

$$G_{ff}(s) = -\frac{K_d\,(\tau_p s+1)\, e^{-\theta_d s}}{K_p\,(\tau_d s+1)\, e^{-\theta_p s}} = \frac{K_{ff}\,(\tau_{ld} s+1)\, e^{-\theta_{ff} s}}{(\tau_{\mathrm{lg}} s+1)}$$

Lead/Lag time constants : $\tau_{ld}\,, \tau_{\mathrm{lg}}$

Feedforward Controller parameters : $K_{ff}\,, \tau_{ld}\,, \tau_{\mathrm{lg}}\,, \theta_{ff}$

Feedforward Control



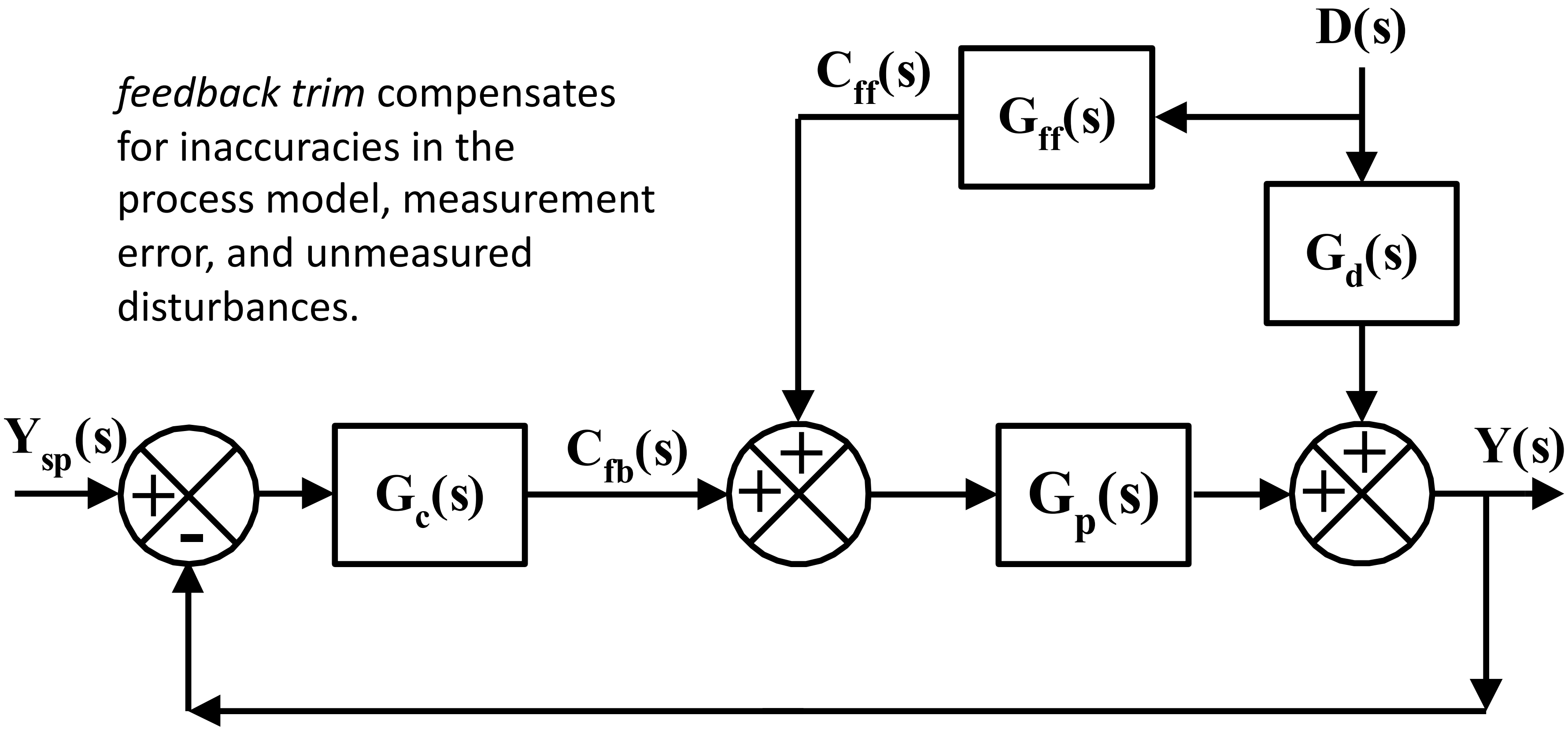
Feedforward Control

Points to consider:

- If $\theta_d < \theta_p$, then $e^{-(\theta_d-\theta_p)s} = e^{+\theta_{ff}s}$
- So the feedforward control element is not practically realizable because of predictive element.
- However, we can approximate it by omitting the $e^{+\theta_{ff}s}$ term and increasing the value of the lead time constant from τ_{ld} to $\tau_{ld} + \theta_{ff}$.
- If $G_{ff}(s) = K_{ff}\frac{(\tau_{p1}s+1)(\tau_{p2}s+1)}{\tau_d s+1}$ then , it is physically unrealizable. Here, also we can approximate as
- $G_{ff}(s) \cong K_{ff}\frac{(\tau_{p1}+\tau_{p2})s+1}{\tau_d s+1}$

Combined FF and FB Control

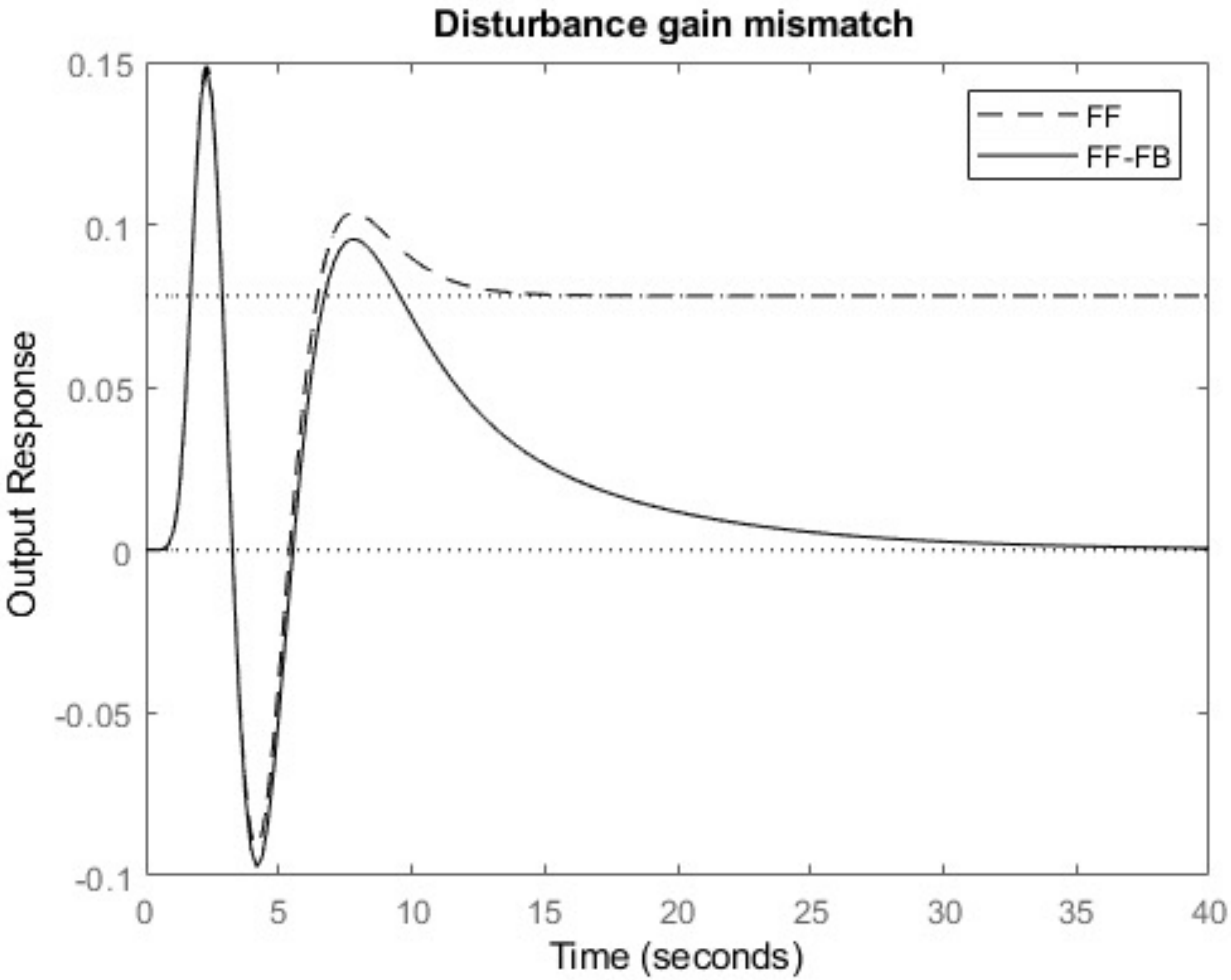
feedback trim compensates for inaccuracies in the process model, measurement error, and unmeasured disturbances.



$$Y(s) = \frac{G_C(s)G_p(s)}{1+G_C(s)G_p(s)}Y_{sp}(s) + \frac{G_{ff}(s)\ G_p(s) + G_d(s)}{1+G_C(s)G_p(s)}D(s) = 0$$

Stability of the feedback loop is not altered by the Feedforward Controller

Combined FF and FB Control



Ratio Control

- Ratio control is a special type of feedforward control that has had widespread application in the process industries.
- The objective is to maintain the ratio of two process variables as a specified value.
- The two variables are usually flow rates, a manipulated variable *u*, and a disturbance variable *d*.
- Thus, the ratio $R = \frac{u}{d}$ is controlled rather than the individual variables.

Typical applications of ratio control

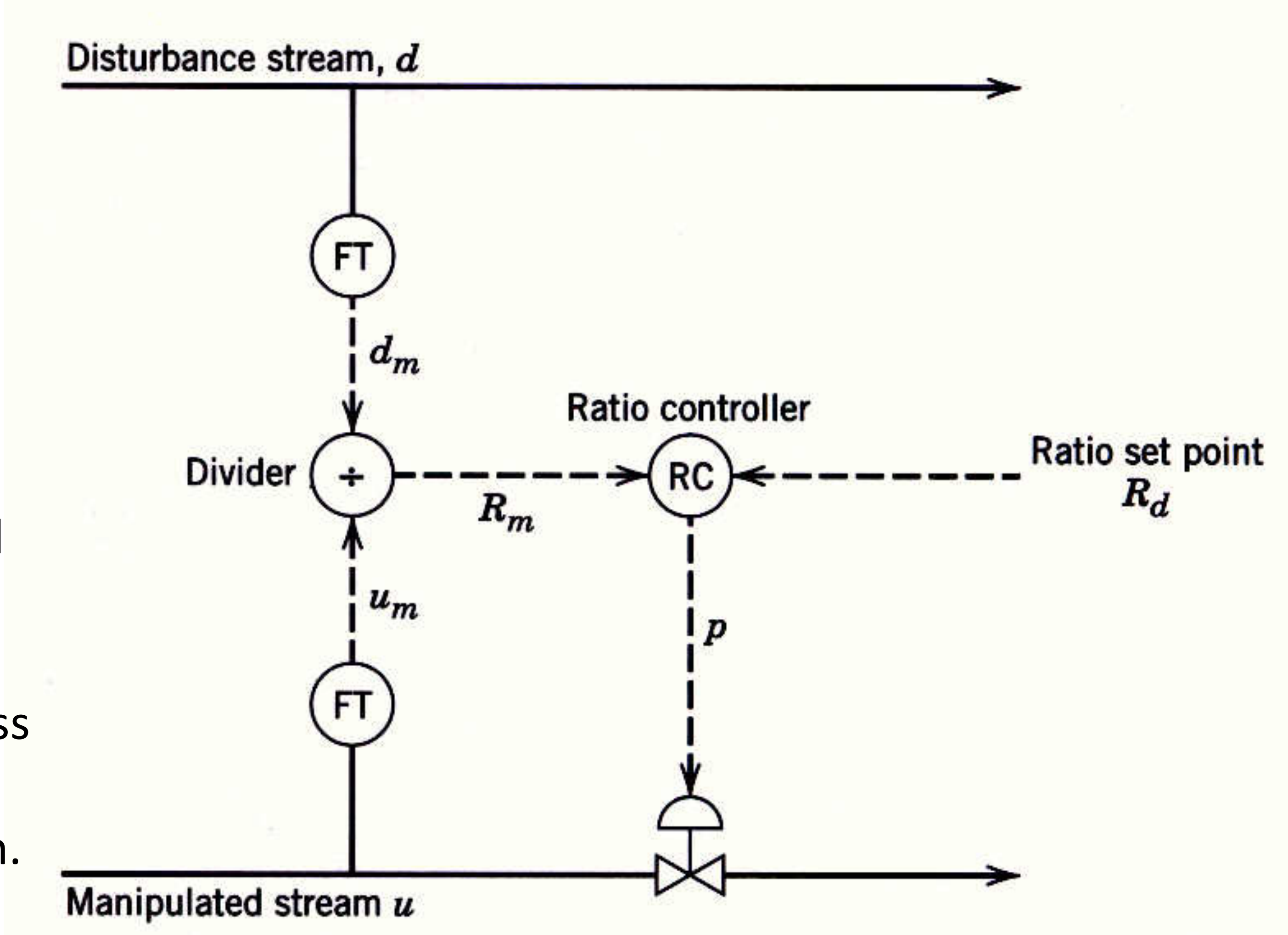
1. Setting the relative amounts of components in blending operations

Advantage

Actual ratio *R* is calculated.

Disadvantage

Divider element must be included in the loop, and this element makes the process gain vary in a nonlinear fashion.



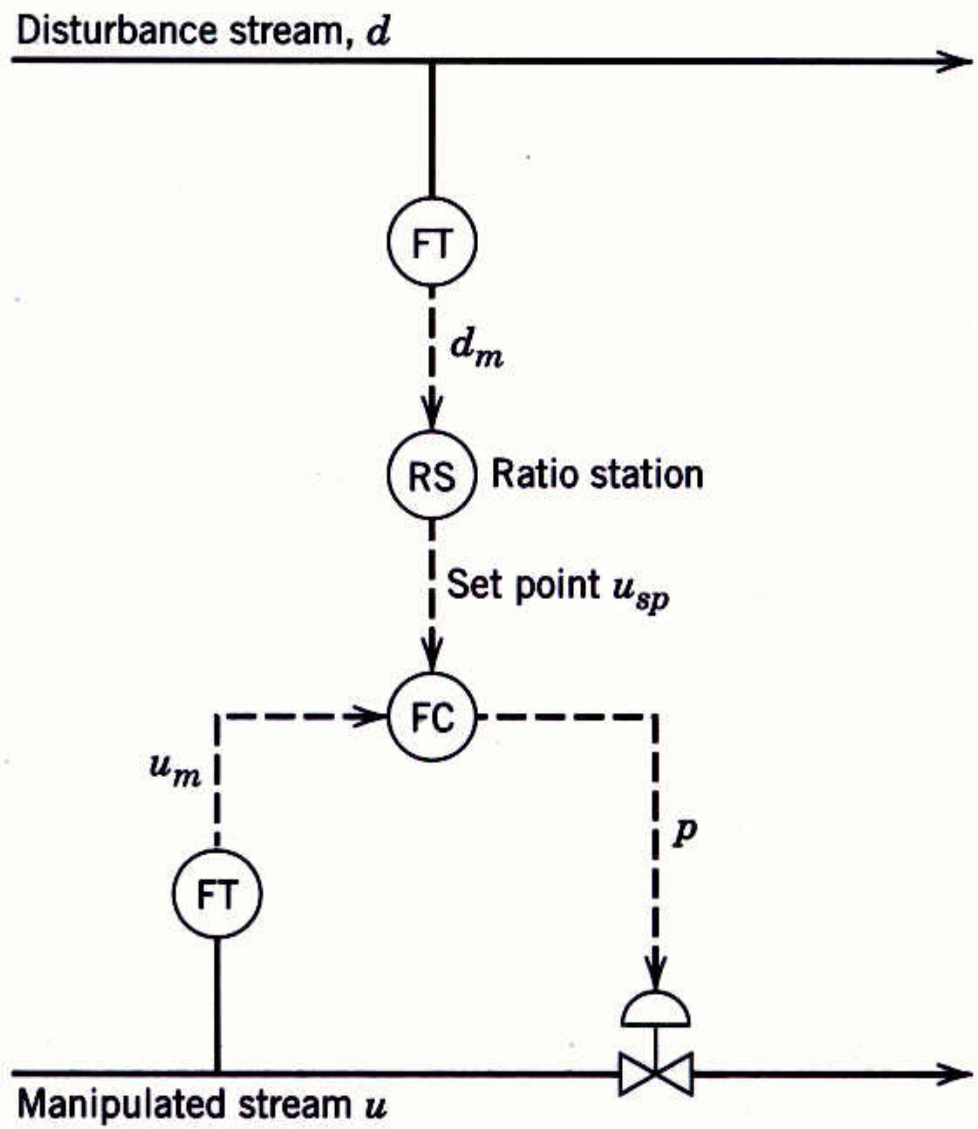
Typical applications of ratio control

Preferred Scheme of Ratio Control.

• Regardless of how ratio control is implemented, the process variables must be scaled appropriately.

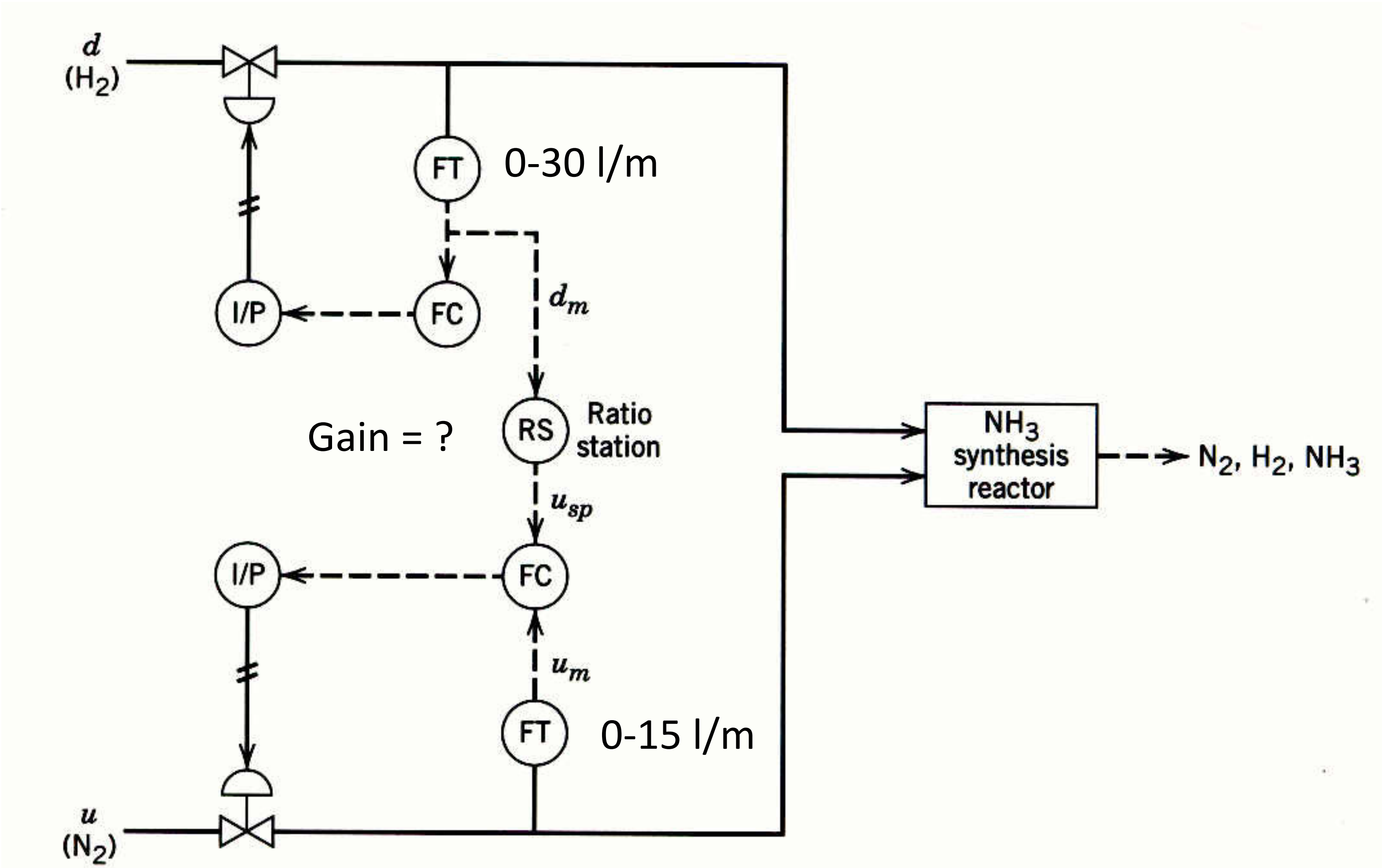
• Gain setting for the ratio station K_r must take into account the spans of the two flow transmitters.

• $K_r = R_d \frac{S_d}{S_u}$

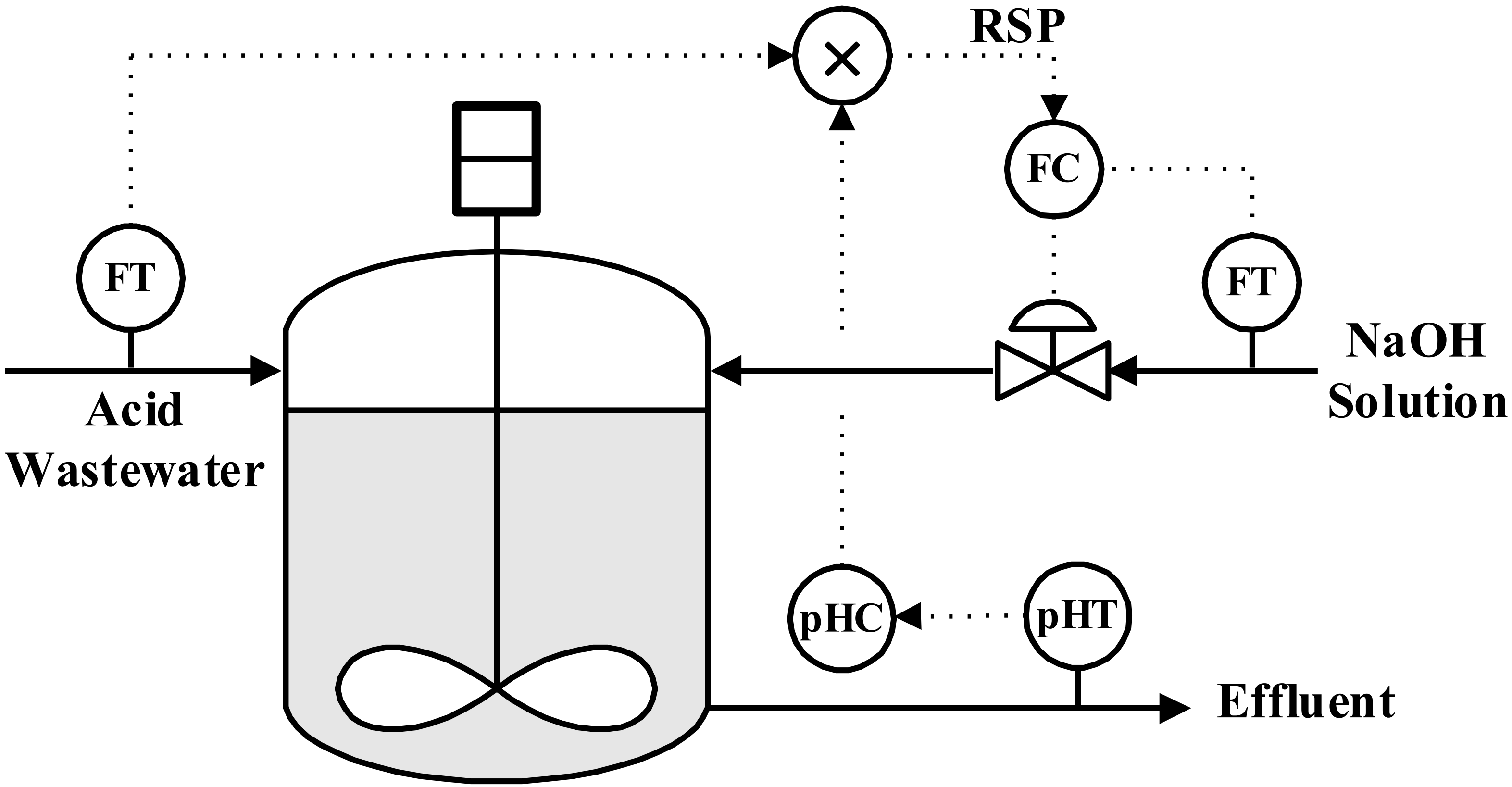


Typical applications of ratio control

- Maintaining a stoichiometric ratio of reactants to a reactor

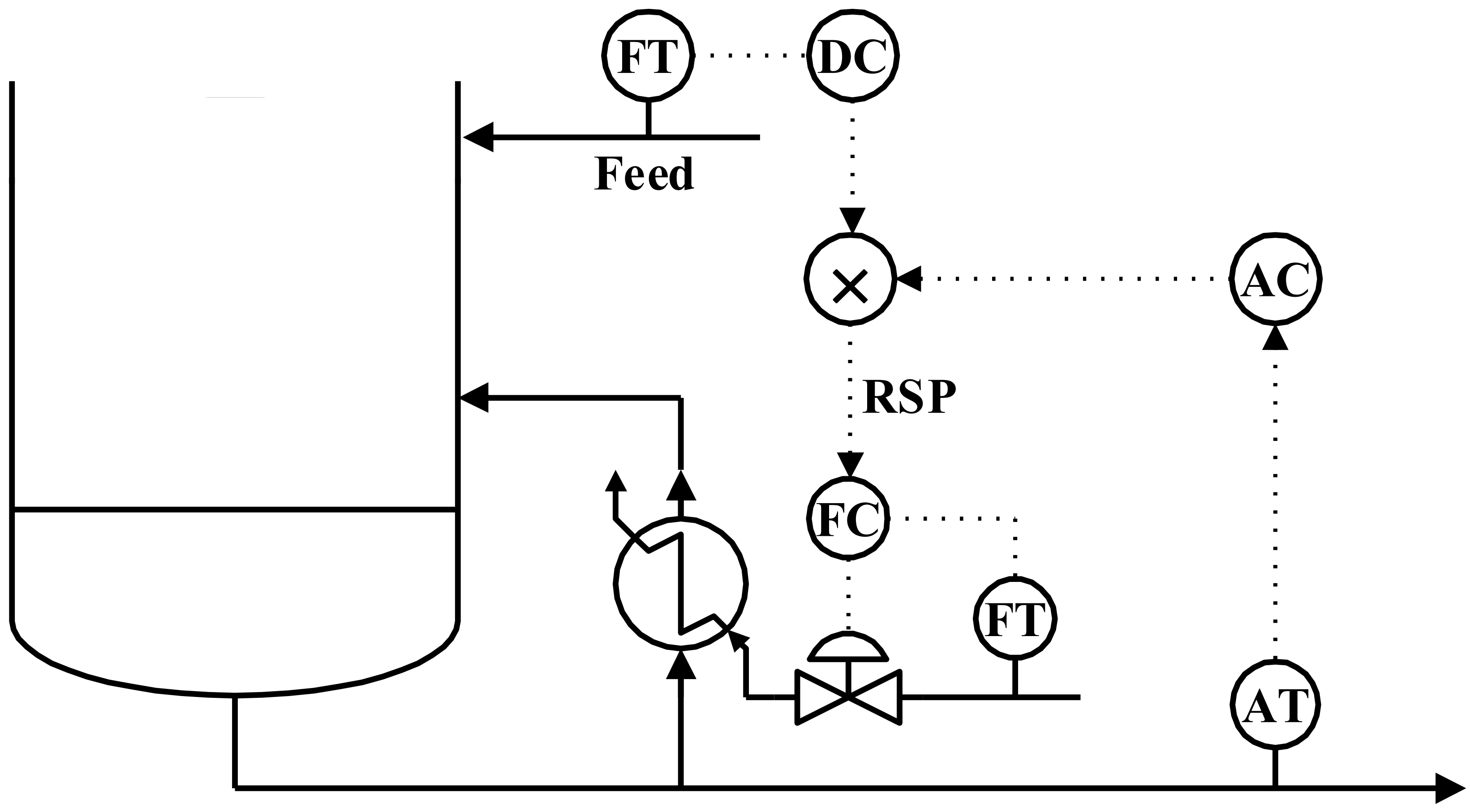


Ratio Control for Wastewater Neutralization



The output of the pH controller is the ratio of NaOH flow rate to acid wastewater flow rate; therefore, the product of the controller output and the measured acid wastewater flow rate become the setpoint for the flow controller of the NaOH addition.

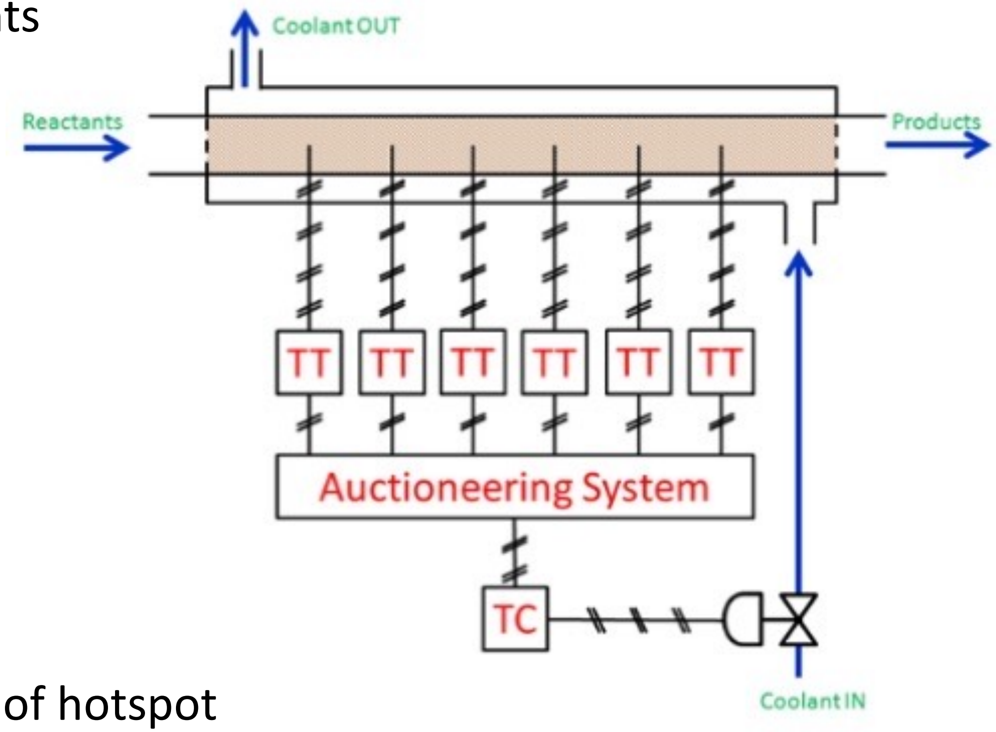
Ratio Control Requiring Dynamic Compensation



Auctioneering Control

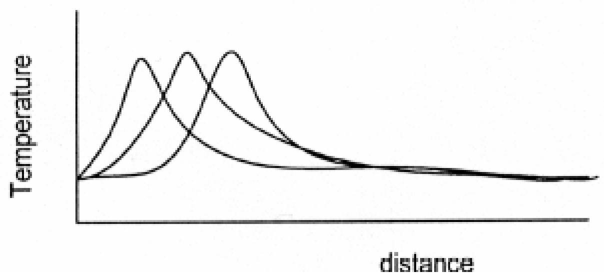
Control of Tubular reactor

- More than one measurements
- One manipulated variable
- Selection of appropriate output from multiple measurements
- Low Selector (LS), High Selector(HS or Median Selector(MS) is used



Hotspot Temperature control

- Location of hotspot travels
- Difficult to pinpoint location of hotspot
- Use multiple temperature sensors
- HS selects the highest temperature



Override/Select Controls

- Process are many times operated at the safety or equipment limits in order to maximize process throughput.
- During upset periods, it is essential that safety limits are enforced.

Objective:

- Regulate level and exit flow rate of sand water slurry
- Slurry velocity in the exit line must be above lower limit

Normal : LC is operating

Too Low Flow: FC overtakes

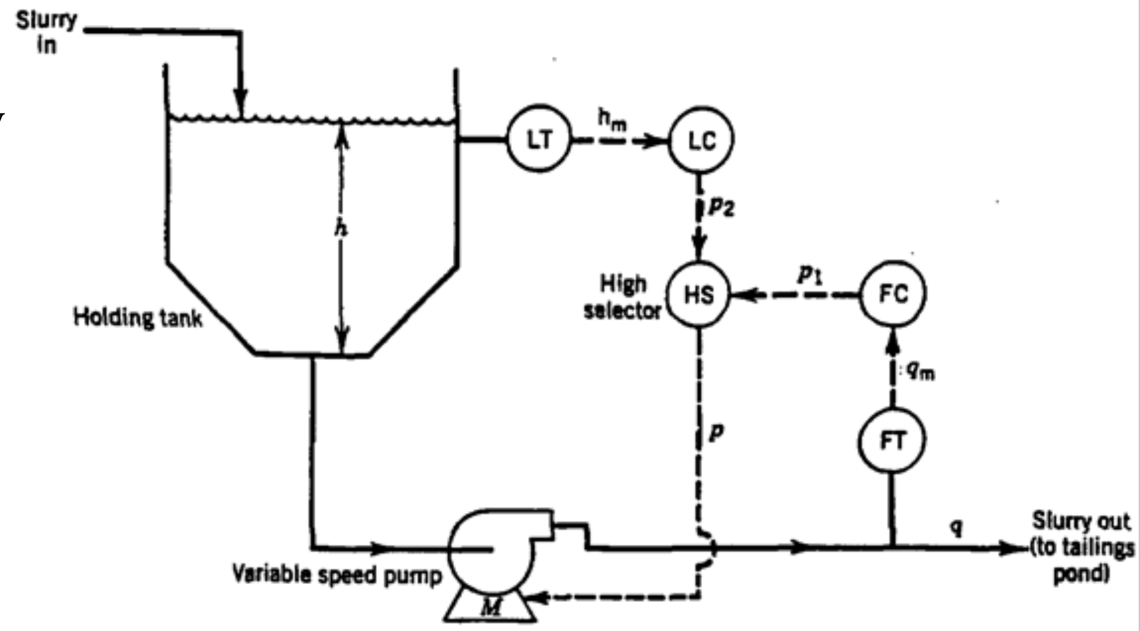
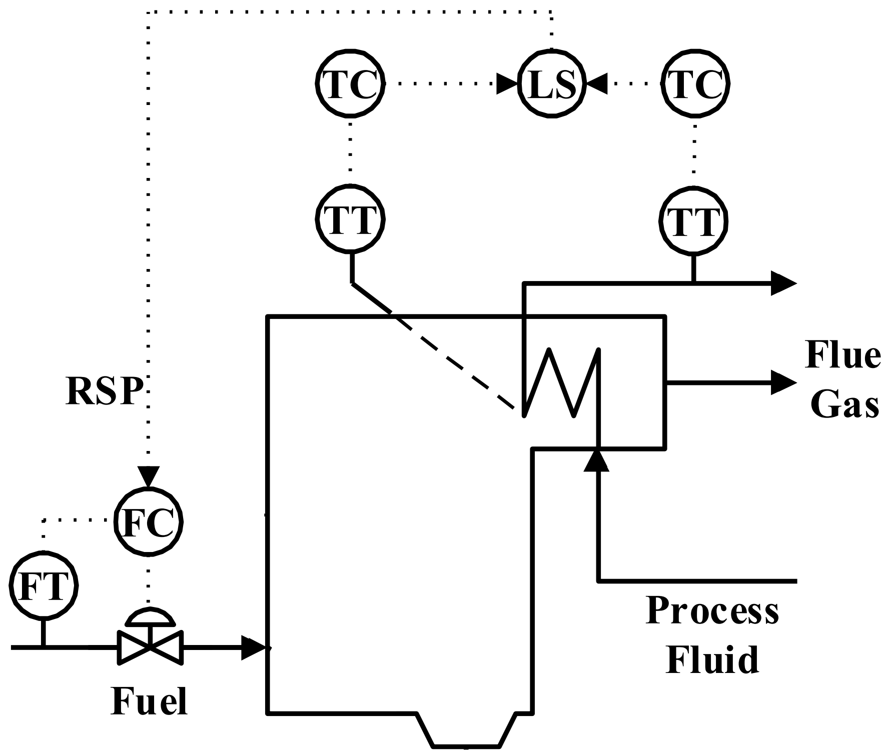


Figure 18.14. A selective control system to handle a sand/water slurry.

Furnace Tube Temperature Constraint Control

- Under normal operation, the controller adjusts the furnace firing rate to maintain process stream at the setpoint temperature.
- At higher fuel rates, excessive tube temperatures can result greatly reducing the useful life of the furnace tubes.
- The LS switch reduces the firing rate to ensure that the furnace tubes are not damaged.



Column Flooding Constraint Control

Objectives:

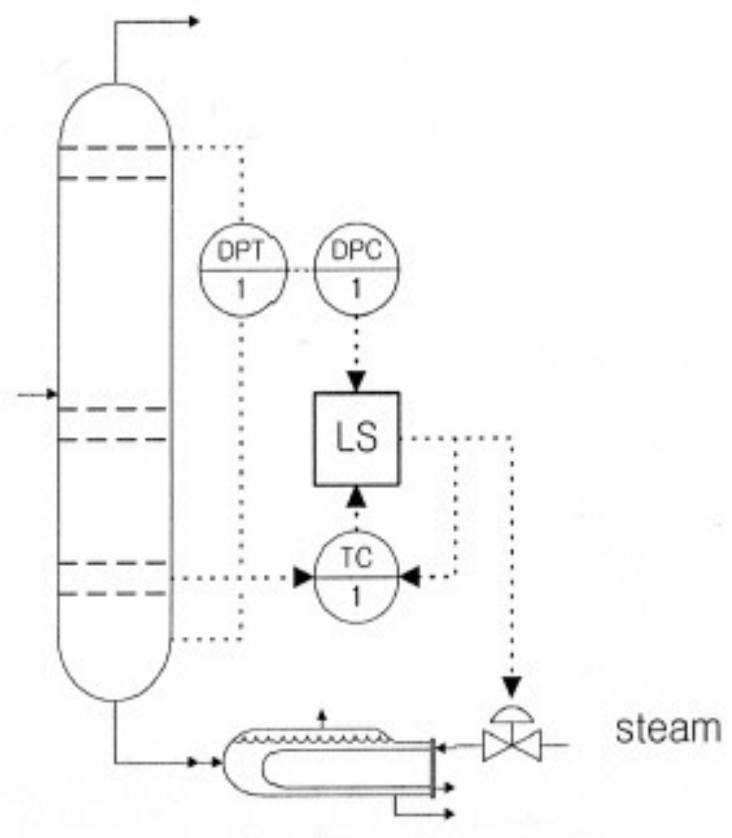
- Maintain the temperature at set point (composition control)
- Maintain $\Delta P < \Delta P_{max}$ (preventing flooding)

Normal: temperature controller is working

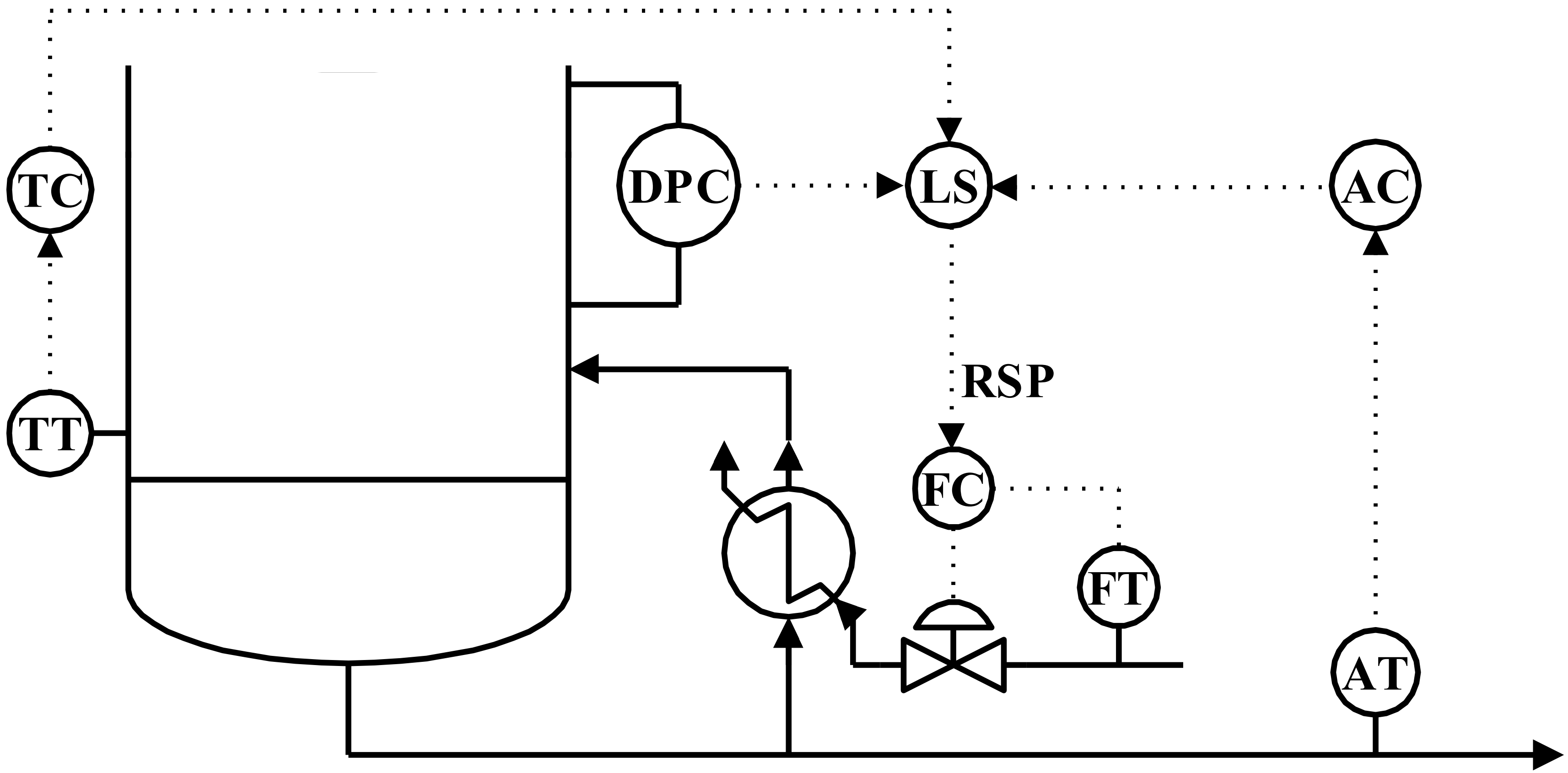
If ΔP is too high, the differential pressure (DP) controller takes over and reduce the steam flow set point.

Temp. controller: PI controller

DP controller: not tightly tuned
P control with flooding limit of DP

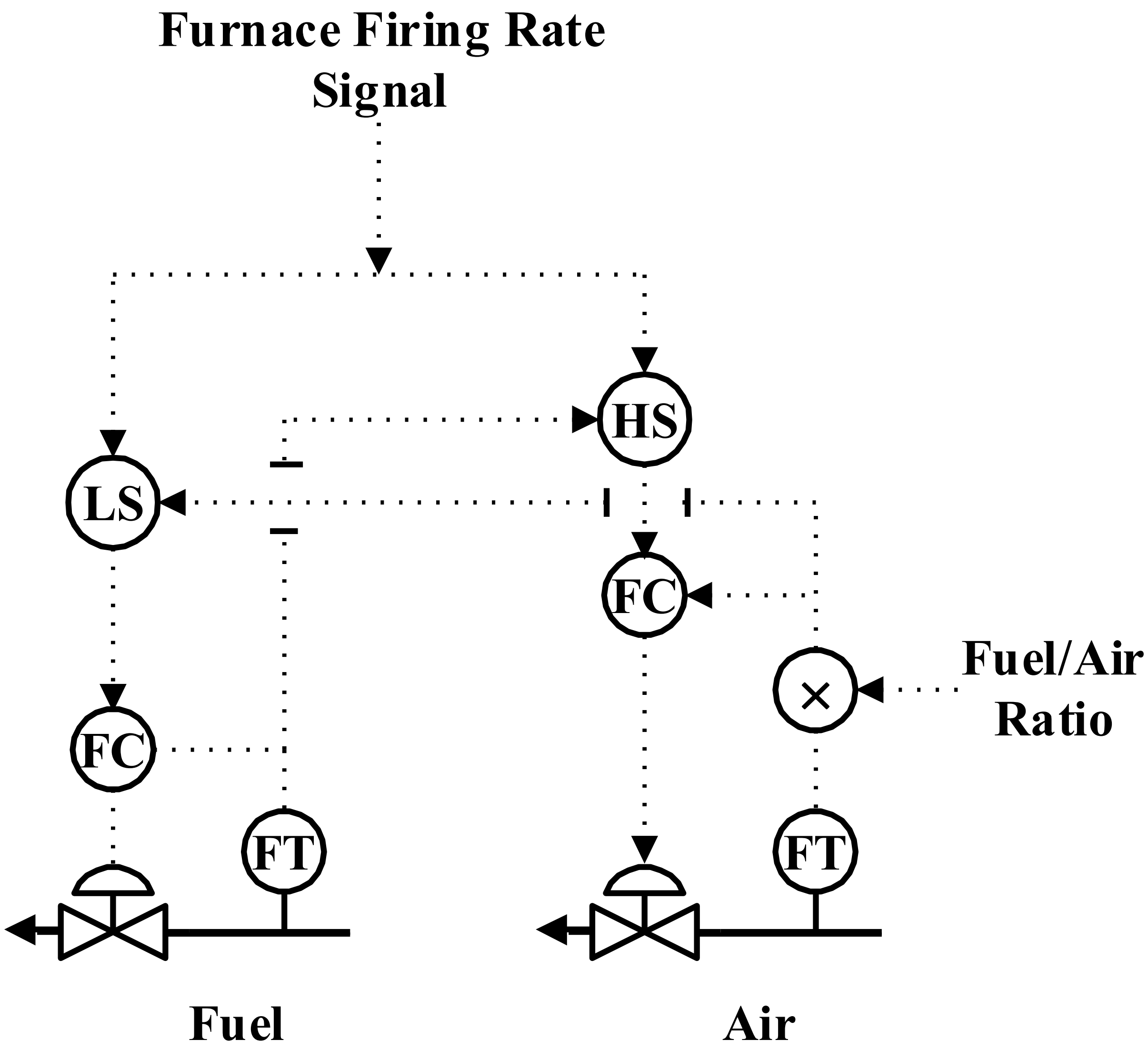


Controlling Multiple Constraints



Cross-Limiting Firing Controls

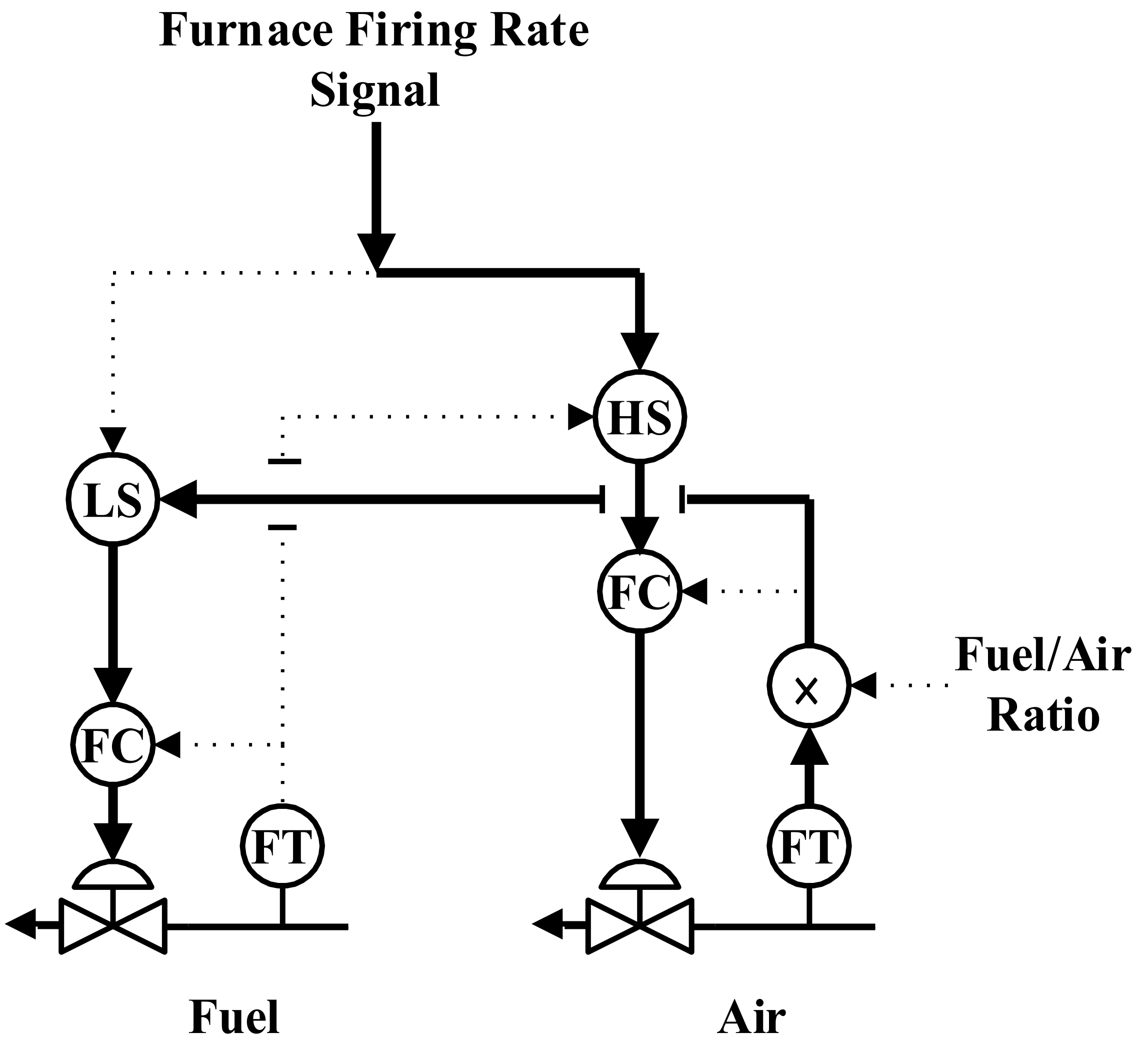
It is critical that excess oxygen is maintained during firing rate increases or decreases or CO will form.



Cross-Limiting Firing Controls

It is critical that excess oxygen is maintained during firing rate increases or decreases or CO will form.

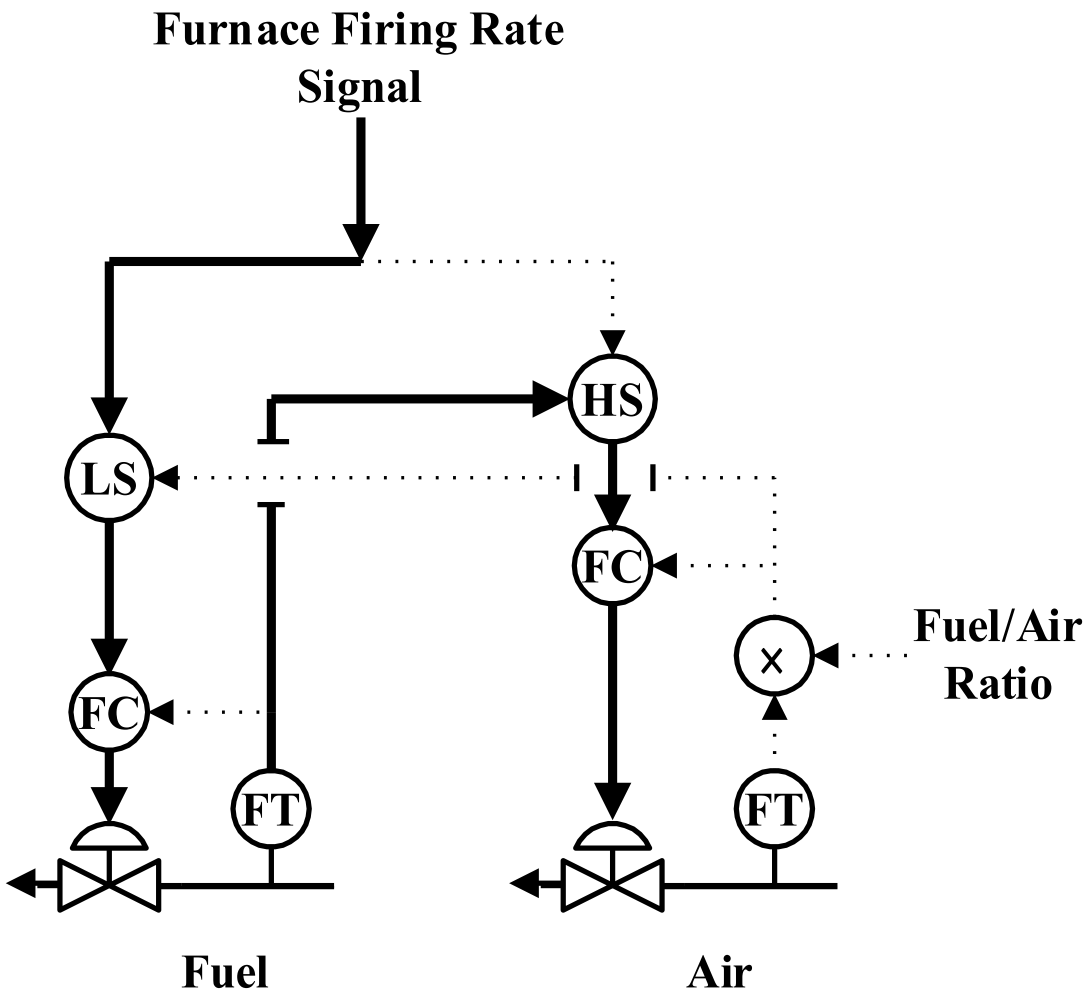
When the firing rate is increased, the air flow rate will lead the fuel flow rate.



Cross-Limiting Firing Controls

It is critical that excess oxygen is maintained during firing rate increases or decreases or CO will form.

When the firing rate is decreased, the fuel flow rate will lead the air flow rate.

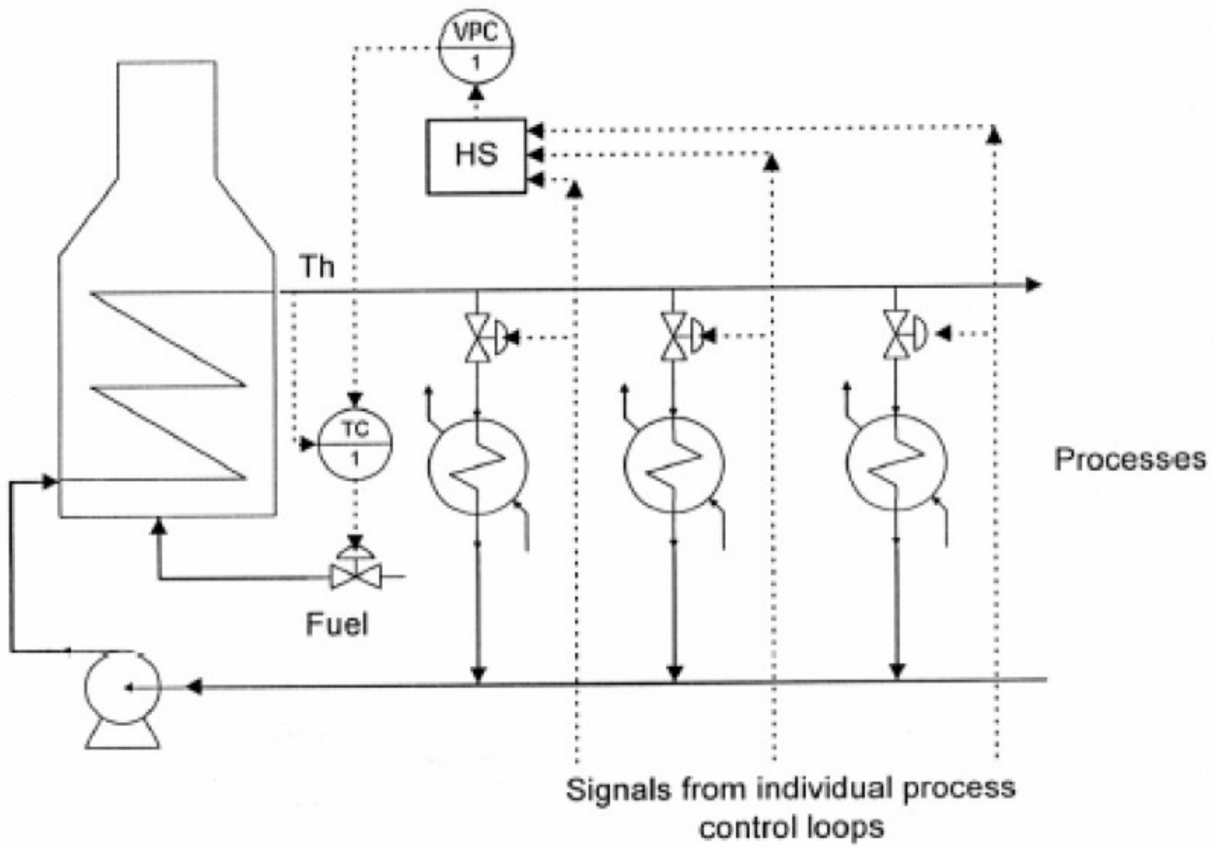


Valve Position Control

Objectives:
- Supply energy to downstream processes as they require
- Minimize the energy consumption of heater

Heater outlet controller: maintain the temp of the heater as it directed for enough supply of energy to down stream processes

Valve position controller:
- Adjust the TC set point so that the valve opening of at least one of the valves is near "fully open". (slow control)
- I-control can be used. (slow but no offset)
- The SP should be around 90-95% (fully open).

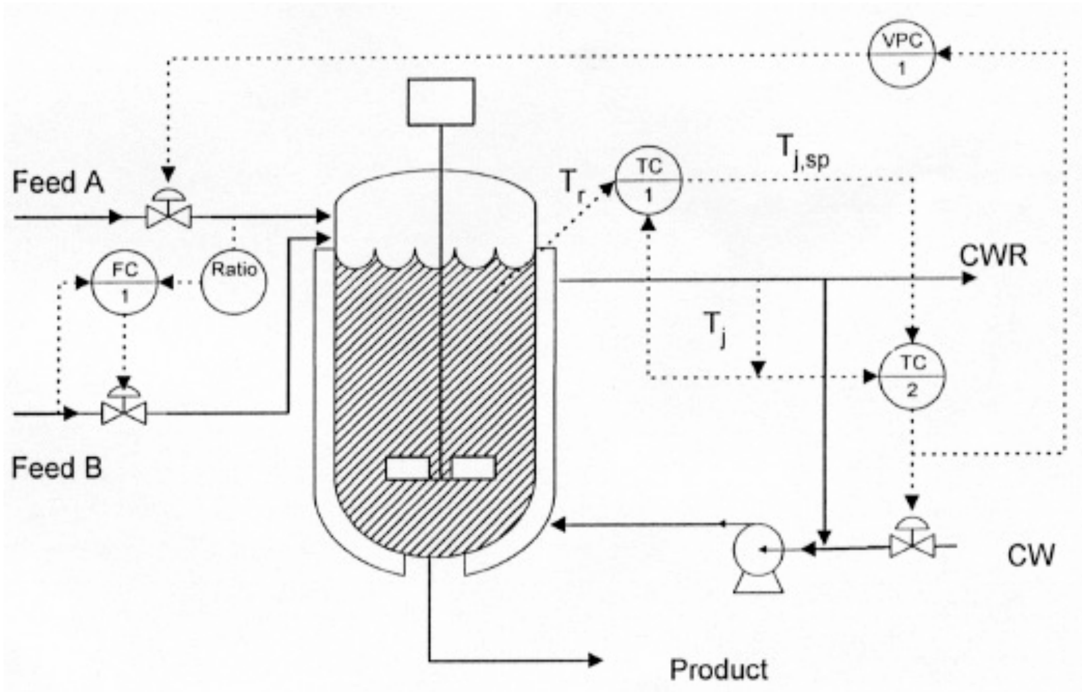


Mixing Tank Cooling Control

Objectives:
- Remove the heat of mixing and maintain mixture temperature
- Maximize the production of mixed feed

Cascaded jacketed tank temp controller: maintain the temp of the mixture by removal of heat of mixing using cooling water

Valve position controller:
- Increase the feed flow of A while cooling capacity allows. (slow control)
- Mix ratio is maintained by ratio control scheme
- The SP should be around 90-95% (fully open).



Floating pressure of distillation column

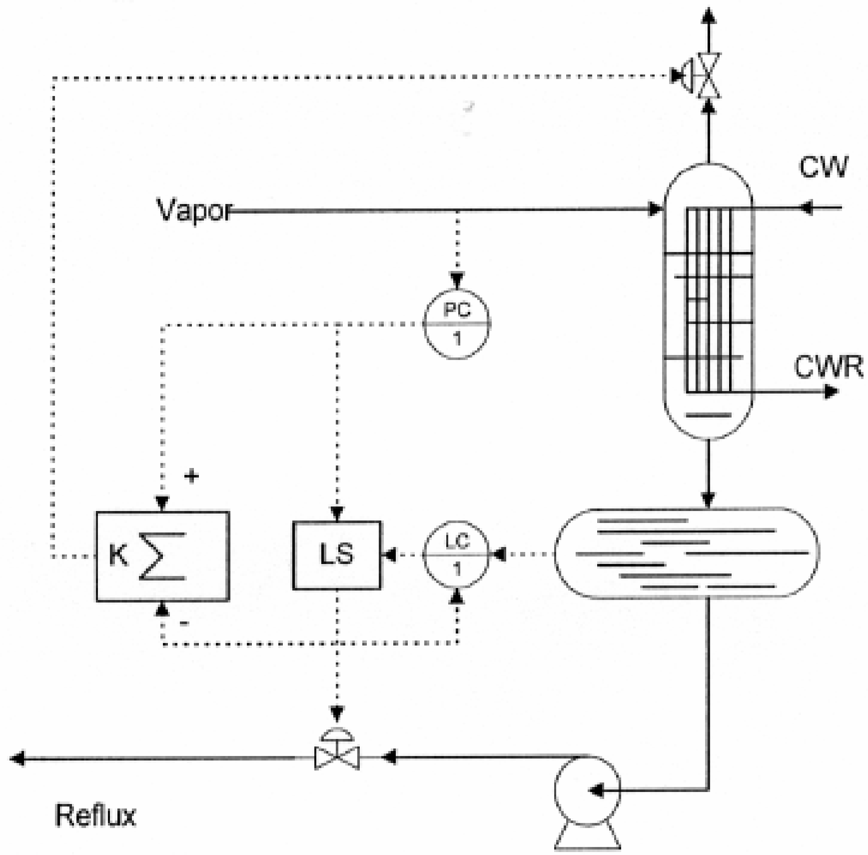
Objectives:

- Maintain the column pressure by adjusting reflux flow rate
- If the level is too low (below the heat transfer area), the vent must be open to relieve the column pressure.

Pressure Controller: control the P by adjusting reflux flow rate.

Level controller: takes over if level is too low. (L<Lmin)

- If LC takes over, there is a discrepancy, and it opens the vent valve.
- The K in summing station decides the vent opening.



Blending Process Control

Objectives:
- Maintain product composition
- Maximize production
- The additive flow rates are limited (constrained)

Control scheme:
- Override control for max production
- Ratio control for composition
- Valve position control for additive flow constraint