Chapter 16

Enhanced Single-Loop Control Strategies

1. Cascade control
2. Time-delay compensation
3. Inferential control
4. Selective and override control
5. Nonlinear control
6. Adaptive control

Example: Cascade Control

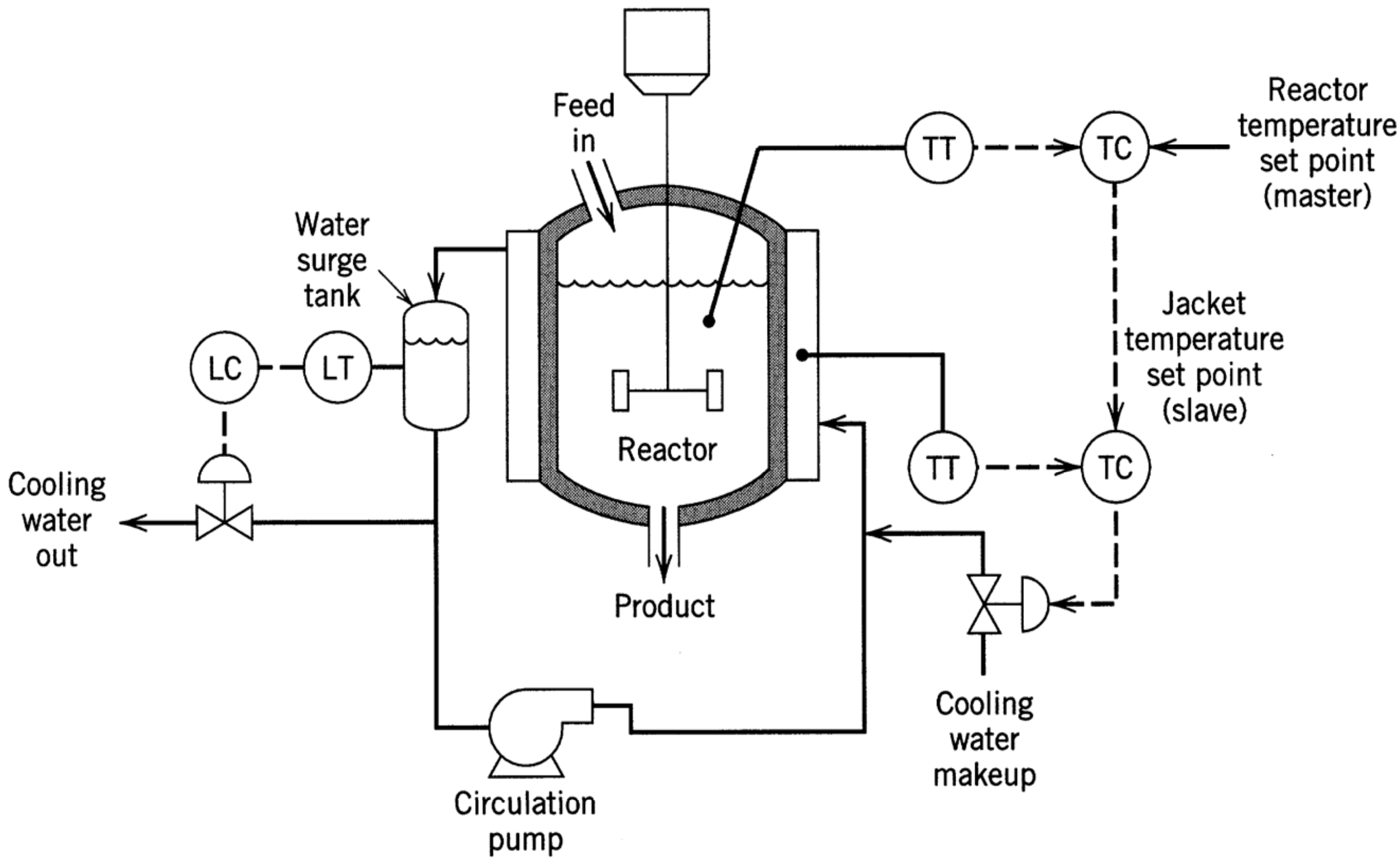

Figure 16.3 Cascade control of an exothermic chemical reactor.

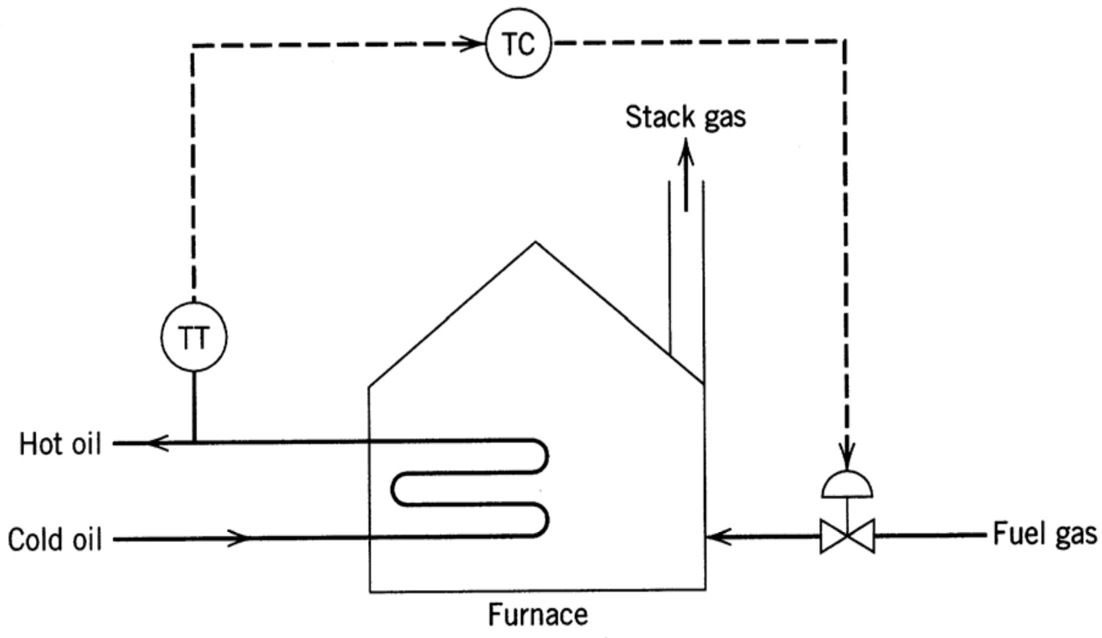

Figure 16.1 A furnace temperature control scheme that uses conventional feedback control.

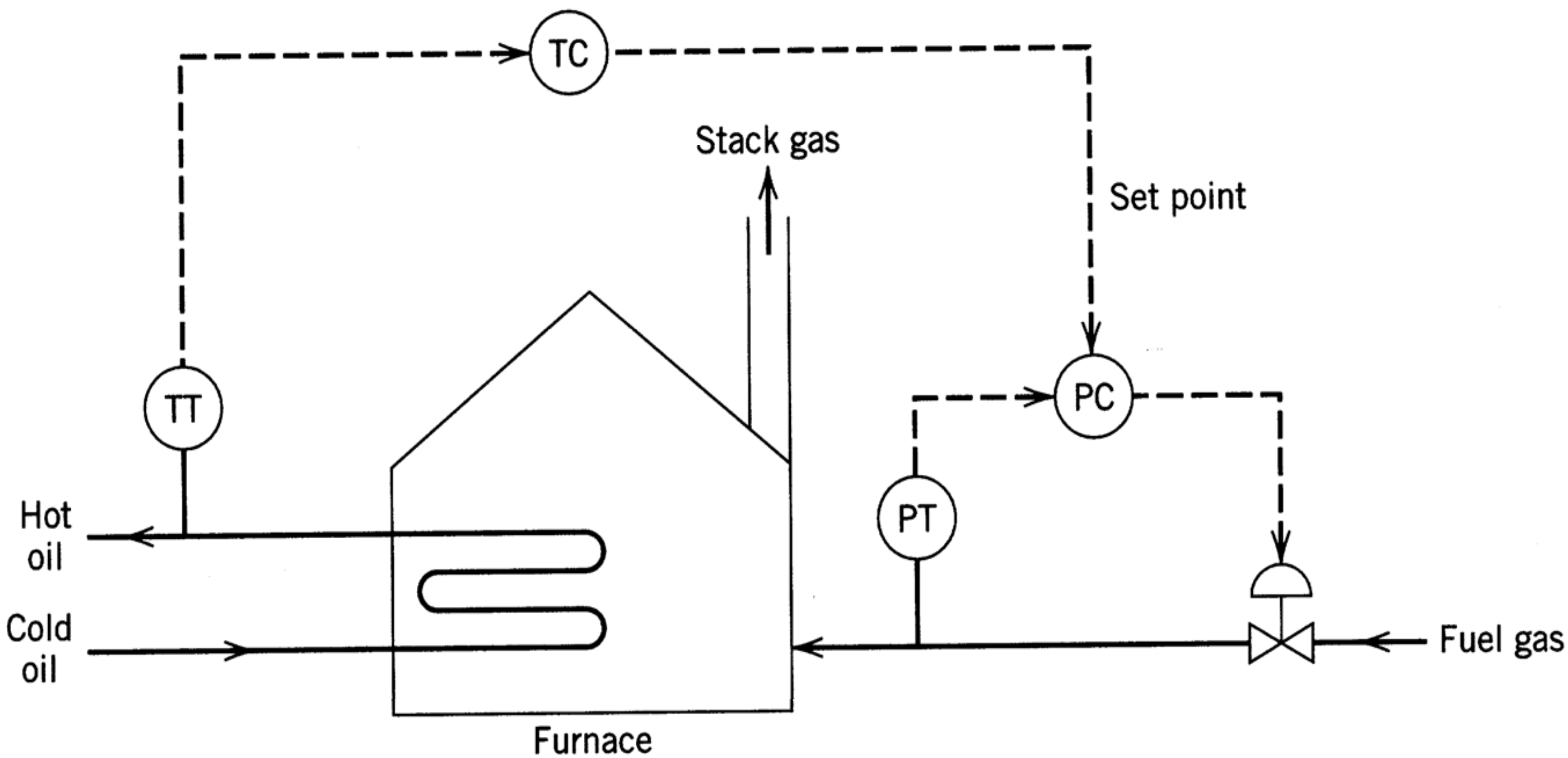

Figure 16.2 A furnace temperature control scheme using cascade control.

Cascade Control

- **Distinguishing features:**

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

- **Terminology:**

slave vs. master
secondary vs. primary
inner vs. outer

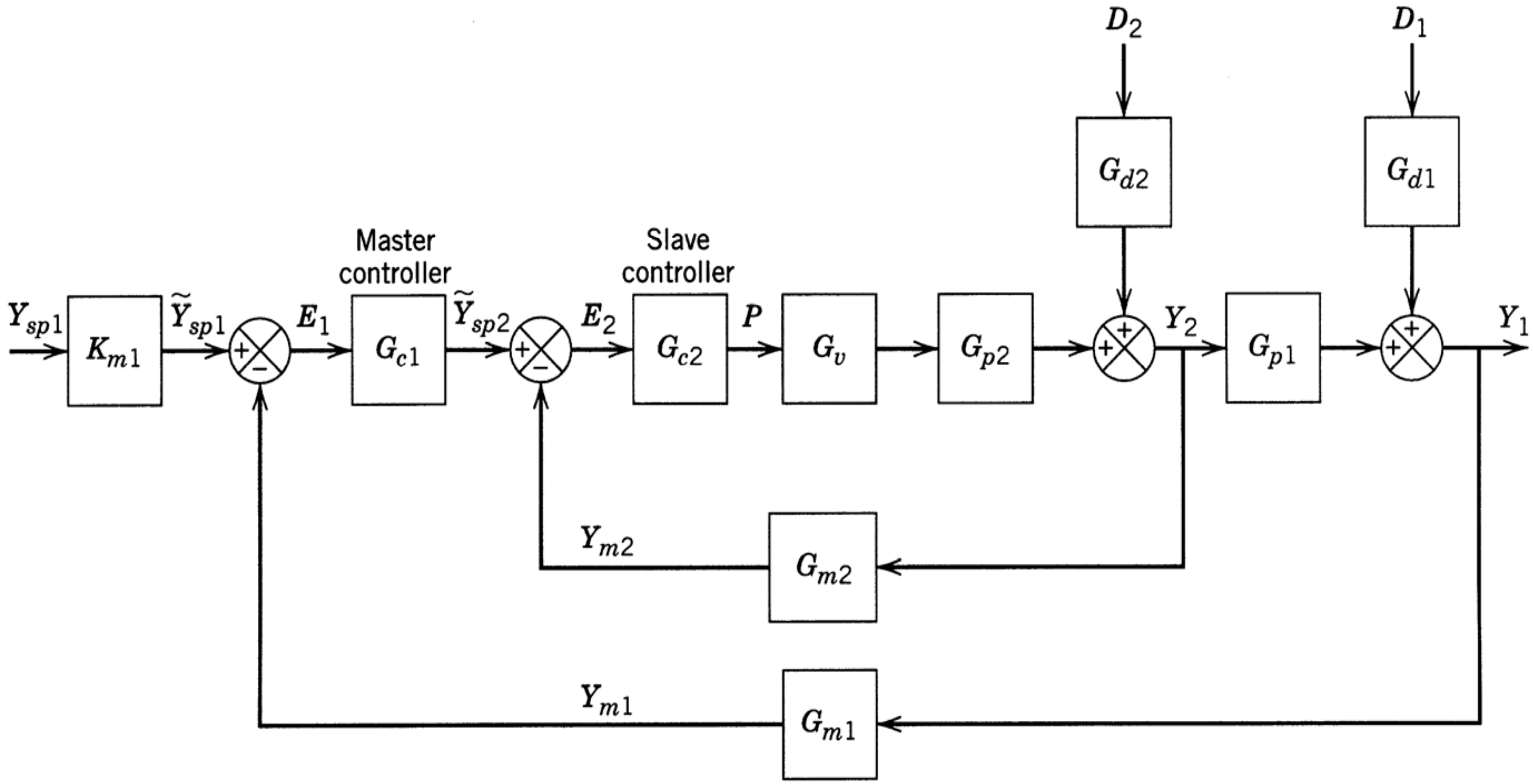

Figure 16.4 Block diagram of the cascade control system.

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$$\frac{Y_1}{D_2} = \frac{G_{P1}G_{d2}}{1+G_{c2}G_vG_{p2}G_{m2}+G_{c1}G_{c2}G_vG_{p2}G_{p1}G_{m1}} \qquad (16-5)$$

Y_1 **= hot oil temperature**

Y_2 **= fuel gas pressure**

D_1 **= cold oil temperature (or cold oil flow rate)**

D_2 **= supply pressure of gas fuel**

Y_{m1} **= measured value of hot oil temperature**

Y_{m2} **= measured value of fuel gas temperature**

Y_{sp1} **= set point for** Y_1

$\tilde{Y}_{sp2}$ **= set point for** Y_2

Example 16.1

Consider the block diagram in Fig. 16.4 with the following transfer functions:

$$G_v = \frac{5}{s+1} \qquad G_{p1} = \frac{4}{(4s+1)(2s+1)} \qquad G_{p2} = 1$$

$$G_{d2} = 1 \qquad G_{m1} = 0.05 \qquad G_{m2} = 0.2 \qquad G_{d1} = \frac{1}{3s+1}$$

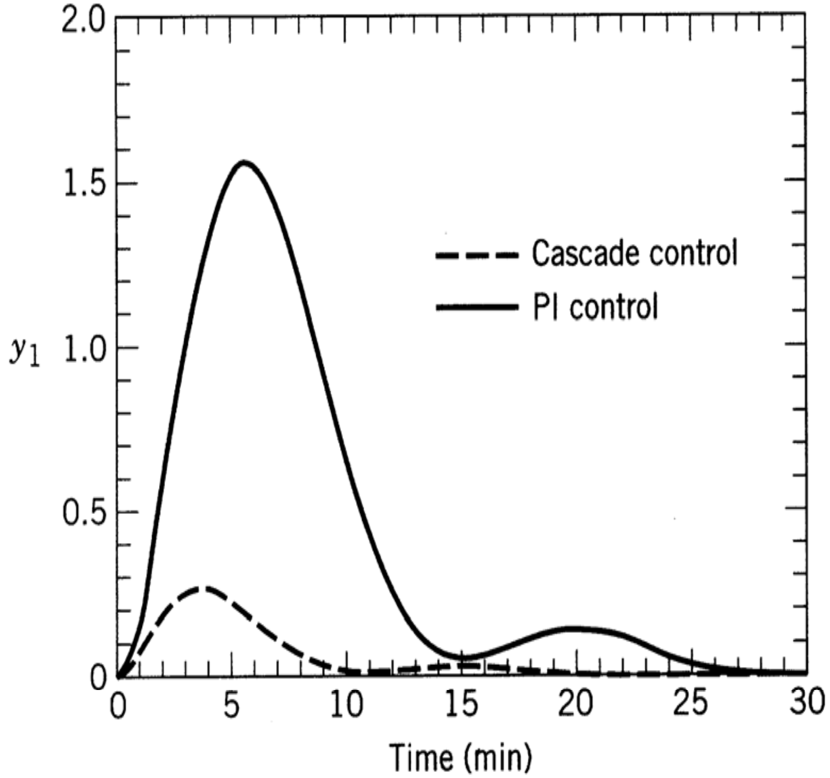

Figure 16.5 A comparison of D_2 unit step responses with and without cascade control.

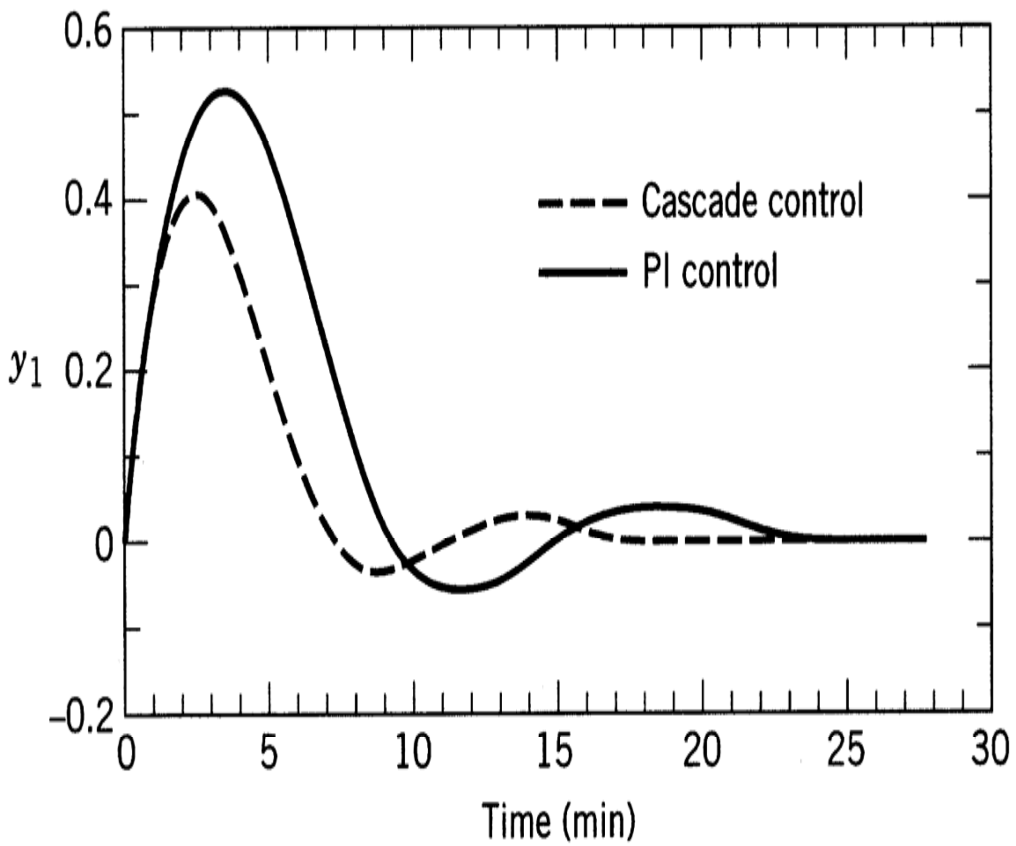

Figure 16.6 A comparison of D_1 step responses.

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Example 16.2

Compare the set-point responses for a second-order process with a time delay (min) and without the delay. The transfer function is

$$G_p(s) = \frac{e^{-\theta s}}{(5s+1)(3s+1)} \qquad (16-18)$$

Assume $G_m = G_v = 1$ and time constants in minutes. Use the following PI controllers. For $\theta = 0, K_c = 3.02 \text{ and } \tau_1 = 6.5$ min, while for $\theta = 2$ min the controller gain must be reduced to meet stability requirements $(K_c = 1.23, \tau_1 = 7.0 \text{ min})$.

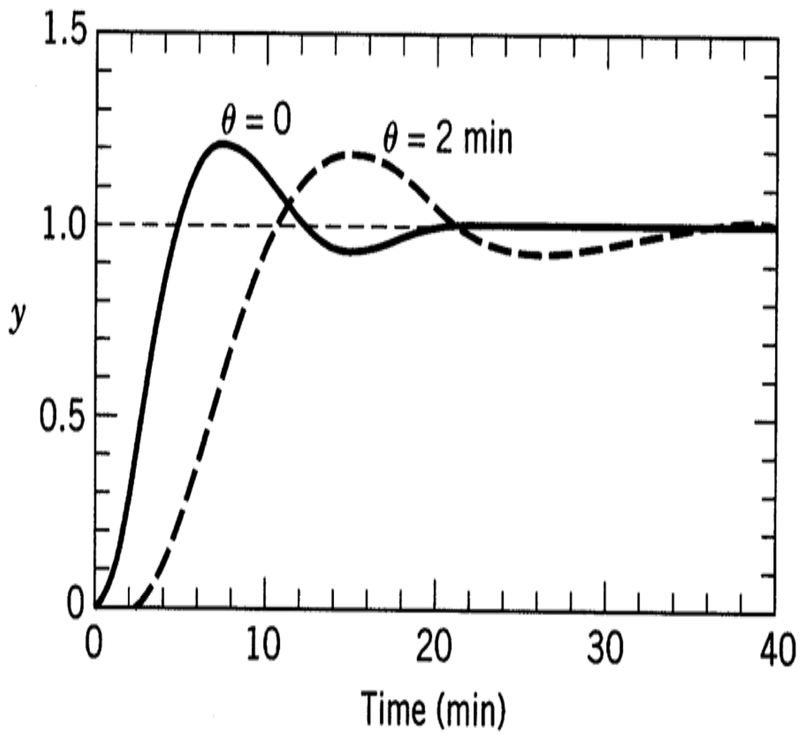


Figure 16.7 A comparison of closed-loop set-point changes.

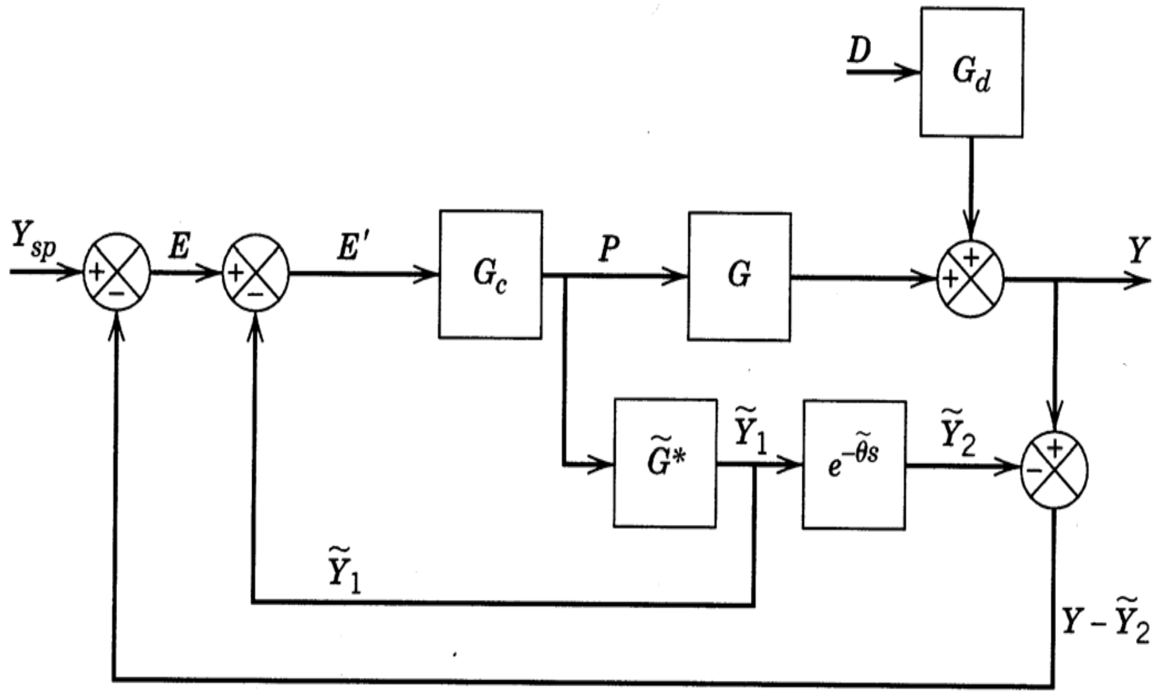


Figure 16.8 Block diagram of the Smith predictor.

$$E' = E - \tilde{Y}_1 = Y_{sp} - \tilde{Y}_1 - \left(Y - \tilde{Y}_2\right) \qquad (16-19)$$

If the process model is perfect and the disturbance is zero, then $\tilde{Y}_2 = Y$ and

$$E' = Y_{sp} - \tilde{Y}_1 \qquad (16-20)$$

For this ideal case the controller responds to the error signal that would occur if not time were present. Assuming there is not model error $\left(\tilde{G} = G\right)$, the inner loop has the effective transfer function

$$G' = \frac{P}{E} = \frac{G_c}{1 + G_c G^* \left(1 - e^{-\theta s}\right)} \qquad (16-21)$$

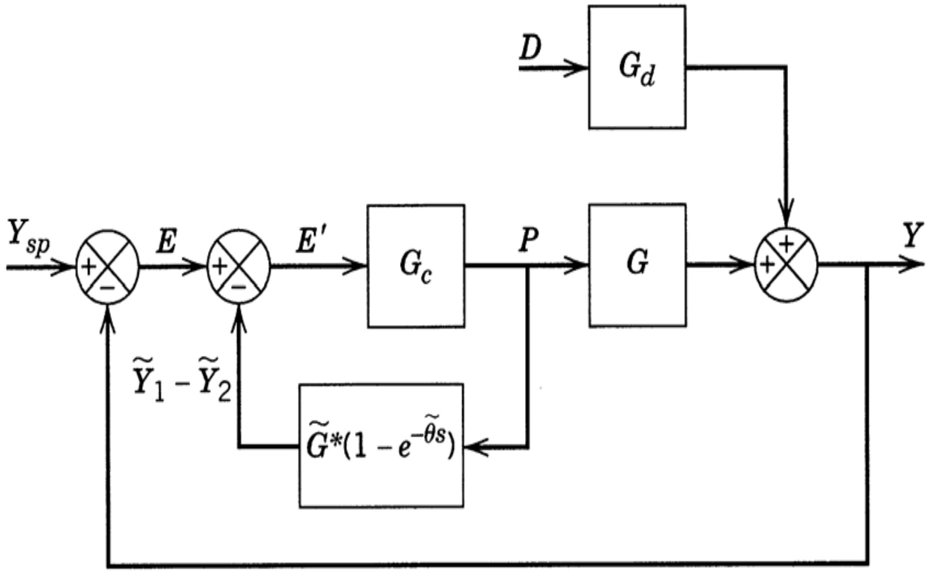


Figure 16.9 An alternative block diagram of a Smith predictor.

For no model error: $\tilde{G} = G = G^* e^{-\theta s}$

$$G_c' = \frac{G_c}{1 + G_c G^* \left(1 - e^{-\theta s}\right)}$$

$$\frac{Y}{Y_{sp}} = \frac{G_c' G^* e^{-\theta s}}{1 + G_c' G^* e^{-\theta s}} = \frac{G_c G}{1 + G_c G^*}$$

By contrast, for conventional feedback control

$$\frac{Y}{Y_{sp}} = \frac{G_c G^* e^{-\theta s}}{1 + G_c G^* e^{-\theta s}} \qquad (16-23)$$

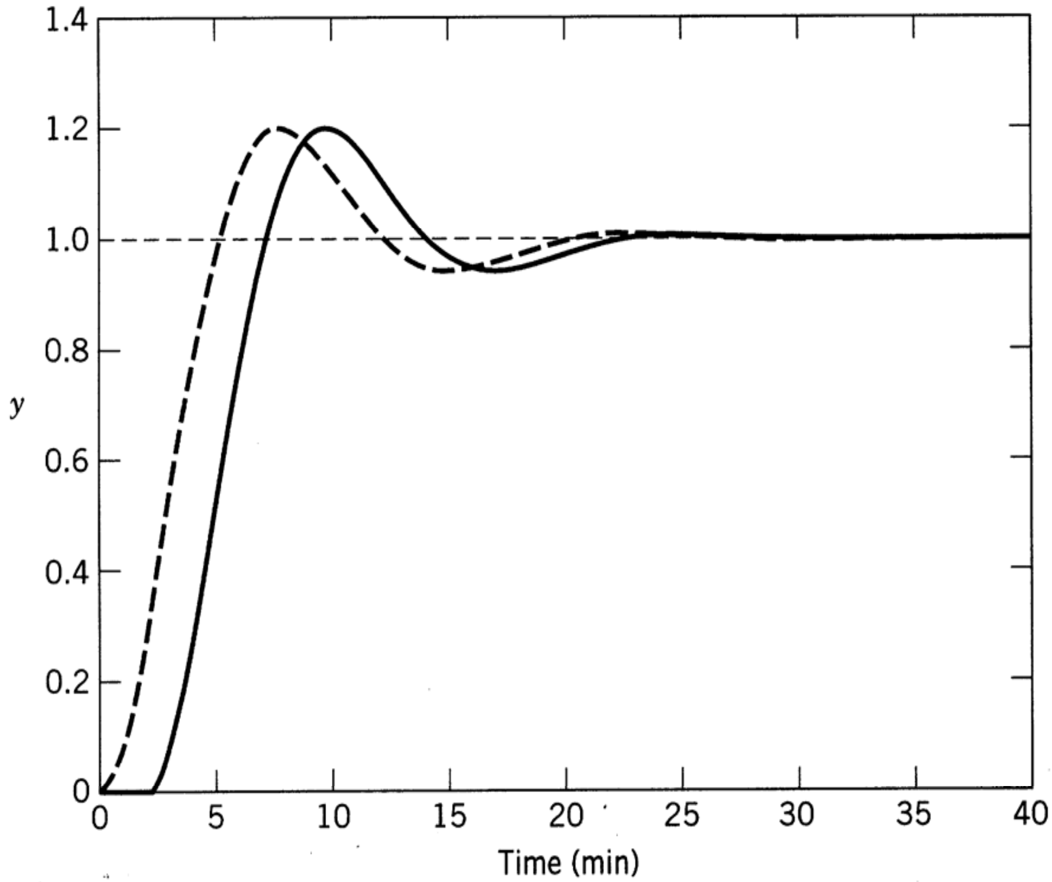


Figure 16.10 Closed-loop set-point change (solid line) for Smith predictor with θ = 2. The dashed line is the response for θ = 0 from Fig. 16.7.

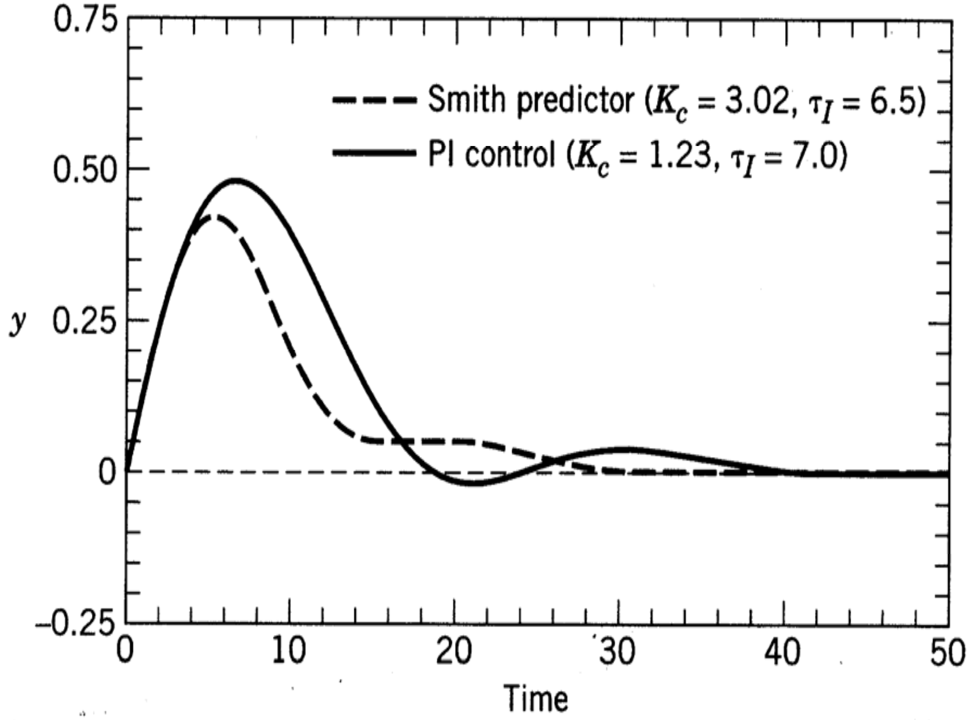


Figure 16.11 A comparison of disturbance changes for the Smith predictor and a conventional PI controller.

Inferential Control

- Problem: Controlled variable cannot be measured or has large sampling period.
- Possible solutions:
  1. Control a related variable (e.g., temperature instead of composition).
  2. Inferential control: Control is based on an estimate of the controlled variable.
     - The estimate is based on available measurements.
       - **Examples**: empirical relation, Kalman filter
     - Modern term: *soft sensor*

Inferential Control with Fast and Slow Measured Variables

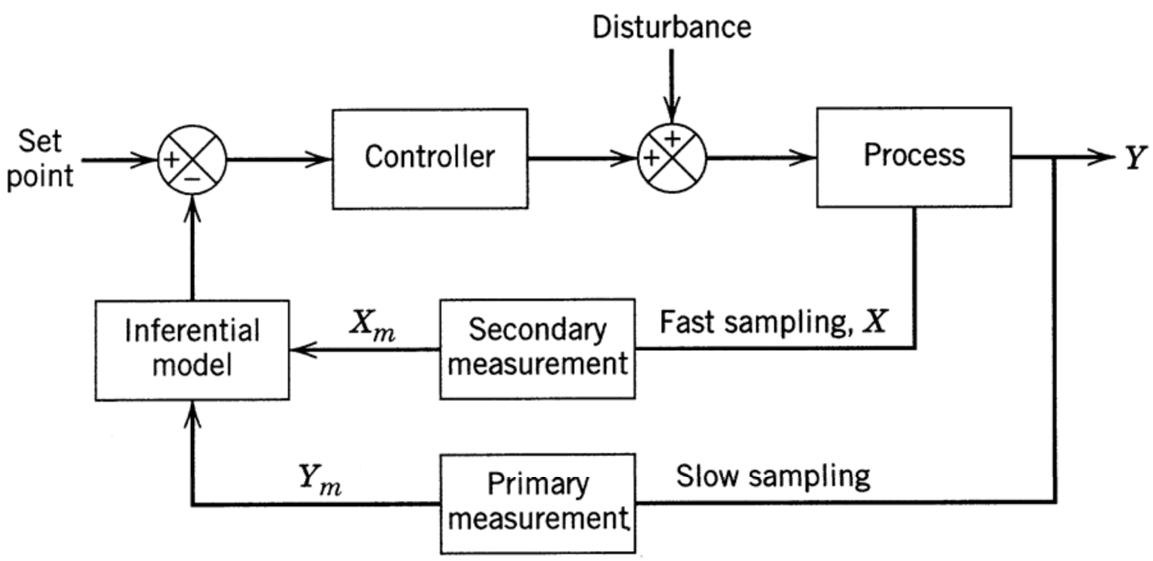


Figure 16.12 Soft sensor block diagram used in inferential control.

Selective Control Systems & Overrides

- For every controlled variable, it is very desirable that there be at least one manipulated variable.

- But for some applications,

$$N_C > N_M$$

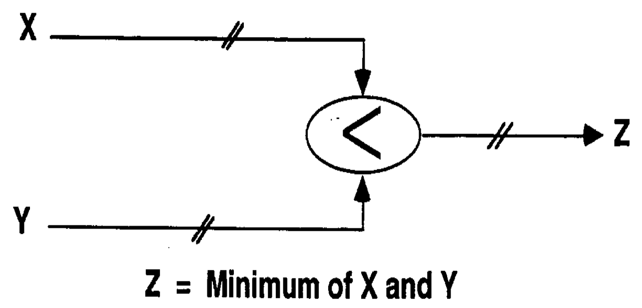
where:

N_C = number of controlled variables

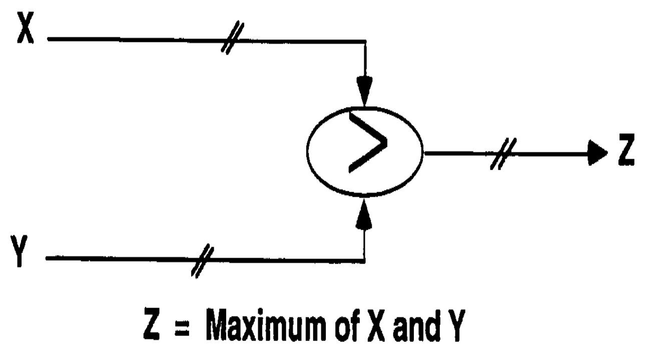
N_M = number of manipulated variables

- **Solution:** Use a *selective control system* or an *override*.

- **Low selector:**



- **High selector:**



- **Median selector:**
  - **The output, Z, is the median of an odd number of inputs**

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Example: High Selector Control System

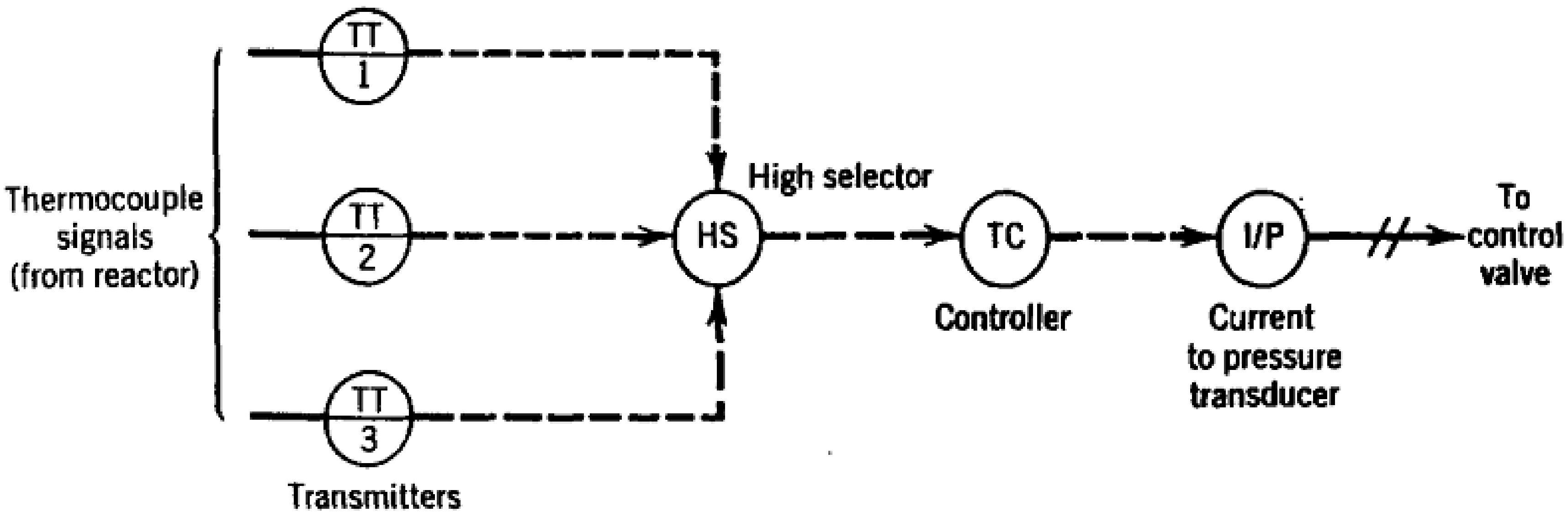

Figure 16.13. Control of a reactor hot spot temperature by using a high selector.

- multiple measurements
- one controller
- one final control element

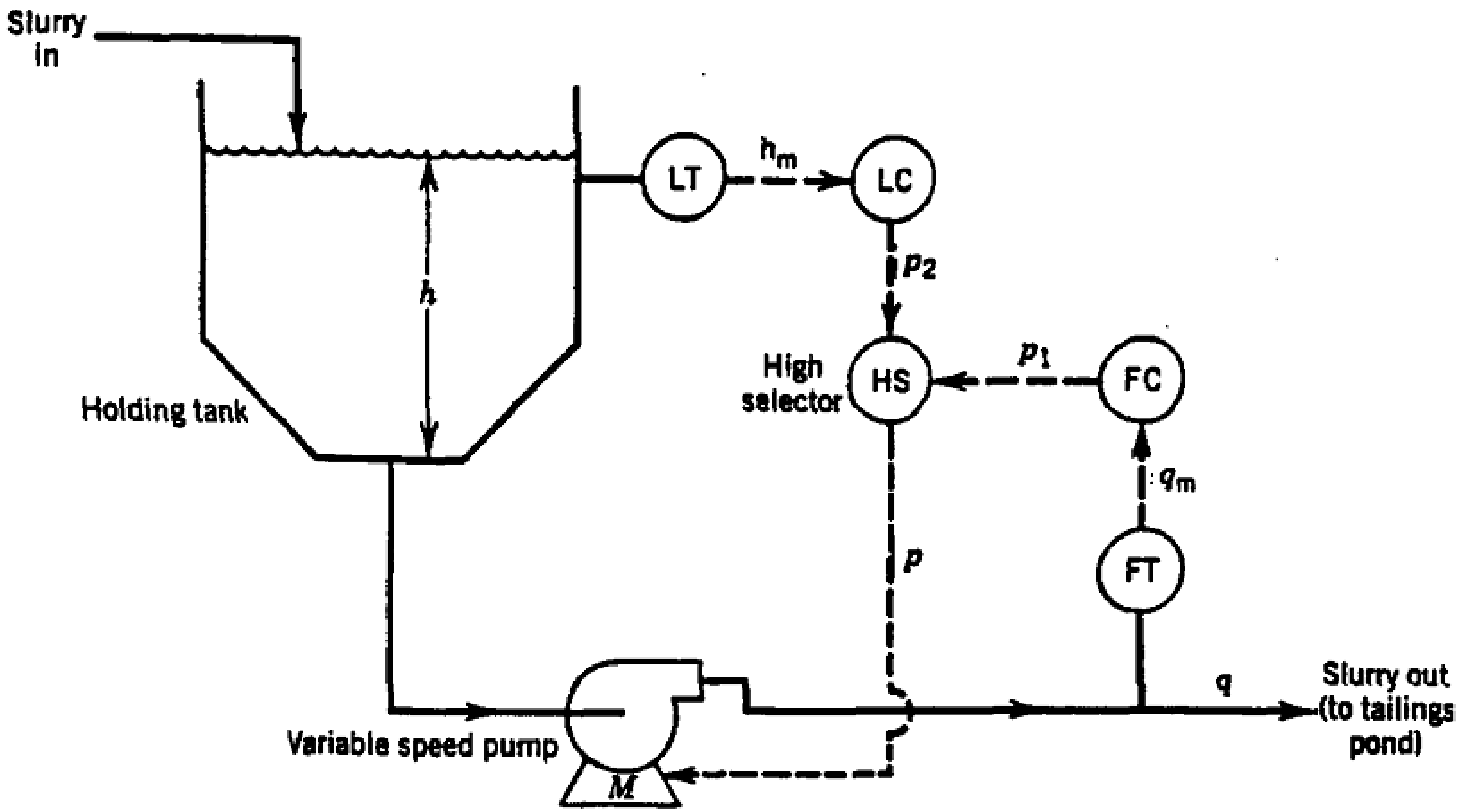

Figure 16.15. A selective control system to handle a sand/water slurry.

2 measurements, 2 controllers,
1 final control element

Overrides

- An *override* is a special case of a selective control system
- One of the inputs is a numerical value, a limit.
- Used when it is desirable to limit the value of a signal (e.g., a controller output).
- Override alternative for the sand/water slurry example?

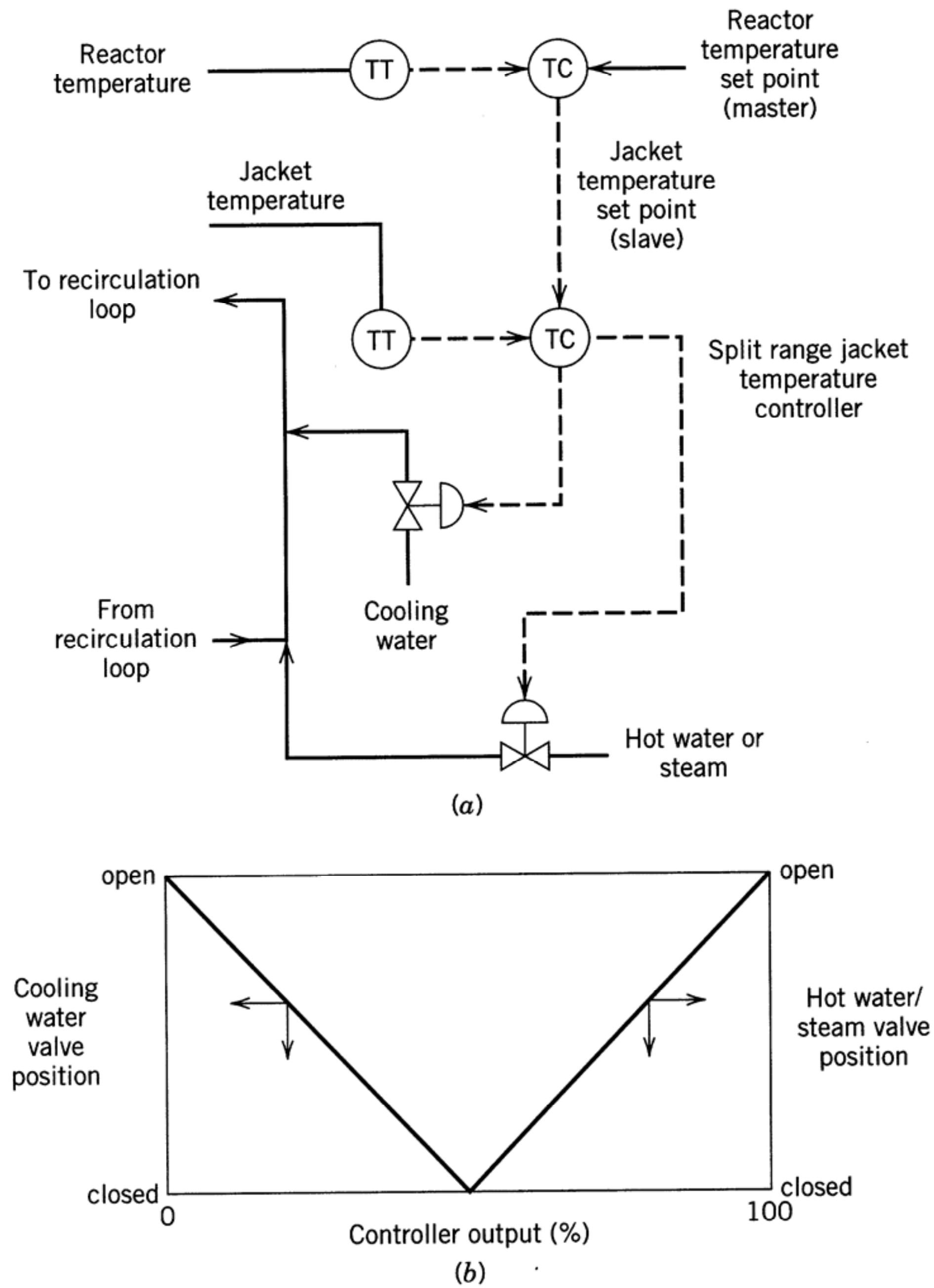

Figure 16.14 Split range control: (*a*) control loop configuration, (*b*) valve position–controller output relationship.

Nonlinear Control Strategies

- Most physical processes are nonlinear to some degree. Some are very nonlinear.

  Examples: pH, high purity distillation columns, chemical reactions with large heats of reaction.

- However, linear control strategies (e.g., PID) can be effective if:

  1. The nonlinearities are rather mild.

  or,

  2. A highly nonlinear process usually operates over a narrow range of conditions.

- For very nonlinear strategies, a nonlinear control strategy can provide significantly better control.

- **Two general classes of nonlinear control:**

  1. Enhancements of conventional, linear, feedback control
  2. Model-based control strategies

  Reference: Henson & Seborg (Ed.), 1997 book.

Enhancements of Conventional Feedback Control

We will consider three enhancements of conventional feedback control:

1. Nonlinear modifications of PID control
2. Nonlinear transformations of input or output variables
3. Controller parameter scheduling such as *gain scheduling*.

Nonlinear Modifications of PID Control:

- **One Example**: nonlinear controller gain

$$K_c = K_{c0}(1 + a|e(t)|) \qquad (16\text{-}26)$$

- K_{c0} and a are constants, and $e(t)$ is the error signal ($e = y_{sp} - y$).
- Also called, *error squared controller*.

Question: Why not use $u \propto e^2(t)$ instead of $u \propto |e(t)|e(t)$?

- *Example*: level control in surge vessels.

Nonlinear Transformations of Variables

- **Objective:** Make the closed-loop system as linear as possible. (Why?)
- **Typical approach:** transform an input or an output.

Example: logarithmic transformation of a product composition in a high purity distillation column. (cf. McCabe-Thiele diagram)

$$x_D^* = log \frac{1 - x_D}{1 - x_{Dsp}} \qquad (16\text{-}27)$$

where x^*_D denotes the transformed distillate composition.

- **Related approach:** Define u or y to be combinations of several variables, based on physical considerations.

<u>*Example:* Continuous pH neutralization</u>

CVs: *pH* and liquid level, h

MVs: acid and base flow rates, q_A and q_B

- *Conventional approach*: single-loop controllers for *pH* and h.
- *Better approach*: control pH by adjusting the ratio, q_A / q_B, and control h by adjusting their sum. Thus,

$$u_1 = q_A / q_B \quad \text{and} \quad u_2 = q_A / q_B$$

Gain Scheduling

- **Objective:** Make the closed-loop system as linear as possible.
- **Basic Idea:** Adjust the controller gain based on current measurements of a "scheduling variable", e.g., *u*, *y*, or some other variable.

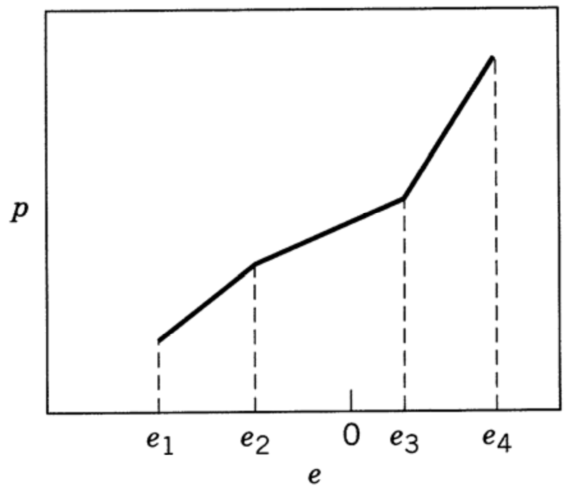

Figure 16.17 A gain-scheduled proportional controller with a controller gain that is piecewise constant.

- Note: Requires knowledge about how the process gain changes with this measured variable.

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Examples of Gain Scheduling

- **Example 1.** Titration curve for a strong acid-strong base neutralization.
- **Example 2.** Once through boiler

The open-loop step response are shown in Fig. 16.18 for two different feedwater flow rates.

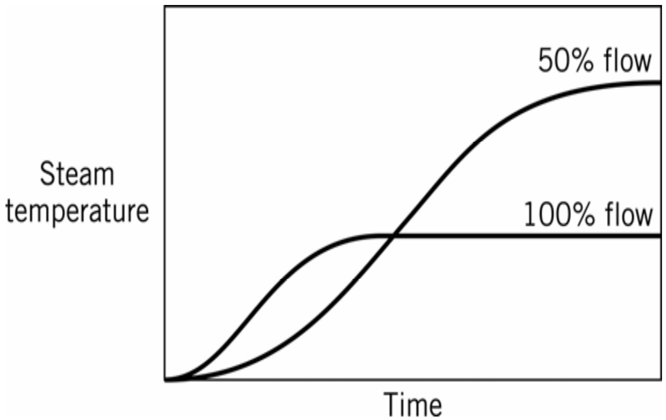

Fig. 16.18 Open-loop responses.

- **Proposed control strategy:** Vary controller setting with w, the fraction of full-scale (100%) flow.

$$K_c = w\bar{K}_c, \quad \tau_I = \bar{\tau}_I / w, \quad \tau_D = \bar{\tau}_D / w, \qquad (16\text{-}30)$$

- Compare with the IMC controller settings for Model H in Table 12.1:

Model H : $G(s) = \dfrac{Ke^{-\theta s}}{\tau s + 1},$

$$K_c = \frac{1}{K}\frac{\tau + \frac{\theta}{2}}{\tau_c + \frac{\theta}{2}}, \quad \tau_I = \tau + \frac{\theta}{2}, \quad \tau_D = \frac{\tau\theta}{2\tau + \theta}$$

Adaptive Control

- A general control strategy for control problems where the process or operating conditions can change significantly and unpredictably.

  Example: Catalyst decay, equipment fouling

- Many different types of adaptive control strategies have been proposed.
- **Self-Tuning Control (STC):**
  - A very well-known strategy and probably the most widely used adaptive control strategy.
  - **Basic idea:** STC is a model-based approach. As process conditions change, update the model parameters by using least squares estimation and recent *u* & *y* data.
- **Note:** For predictable or measurable changes, use gain scheduling instead of adaptive control

  Reason: Gain scheduling is much easier to implement and less trouble prone.

Block Diagram for Self-Tuning Control

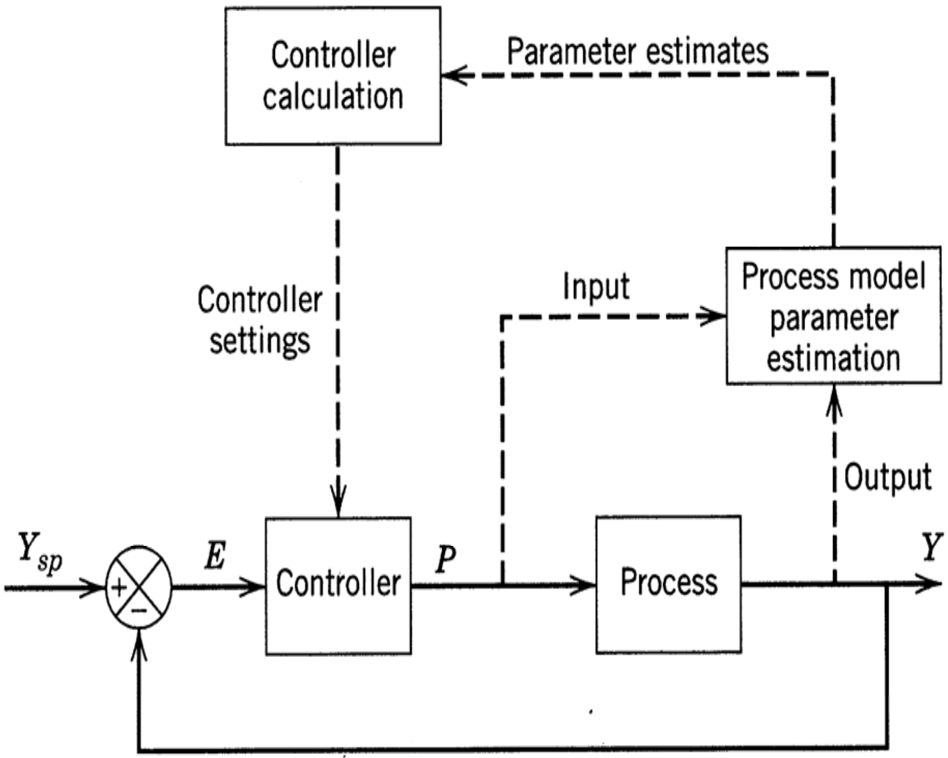

Figure 16.23 A block diagram for self-tuning control.