2.6 Control Systems—Cascade Loops

F. G. SHINSKEY (1970, 1985) **B. G. LIPTÁK** (1995) **R. BARS, J. HETTHÉSSY** (2005)

INTRODUCTION

Cascade control has a multi-loop structure, where the output of the controller in the outer loop (the "primary" or "master") is the set point of a controller in the inner loop (the "secondary" or "slave").

The slave measurement is an intermediate process variable that can be used to achieve more effective control of the primary process variable. In the cascade configuration the process is divided into two parts and therefore two controllers are used, but only one process variable (m) is manipulated.

In Figure 2.6a the primary controller maintains the primary variable y_1 at its set point by adjusting the set point r_2 of the secondary controller. The secondary controller, in turn, responds both to set point r_2 and to the secondary controlled variable y_2. This secondary controlled variable also affects the primary process and therefore the primary controlled variable (y_1), hence the loop is closed.

This cascade loop can also be shown in a more detailed block diagram form (Figure 2.6b). Here, the primary controller (C_1) generates the set point for the secondary controller (C_2), while the secondary controlled variable (y_2) also affects the primary process (P_1) and therefore it also affects the primary controlled variable (y_1).

Cascade control is advantageous on applications where the P_1 process has a large dead time or time lag and the time delays in the P_2 part of the process are smaller. Cascade control is also desirable when the main disturbance is in the secondary loop. This is because with the cascade configuration, the correction of the inner disturbance d_i occurs as soon as the secondary sensor (y_2) detects that upset.

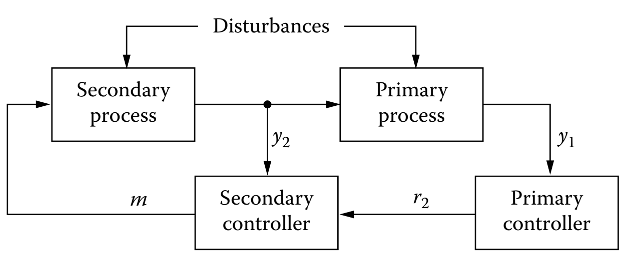

FIG. 2.6a
The cascade control system divides the process into two parts.

Cascade System Advantages

There are two main advantages gained by the use of cascade control:

1. Disturbances that are affecting the secondary variable can be corrected by the secondary controller before their influence is felt by the primary variable.
2. Closing the control loop around the secondary part of the process reduces the phase lag seen by the primary controller, resulting in increased speed of response. This can be seen in Figure 2.6c. Here, the response of the open loop curve shows the slow response of the secondary controlled variable when there is no secondary controller. The much faster closed loop curve describes the response of the secondary controlled variable when the cascade configuration adds a secondary inner control loop.

Other advantages of using cascade control include the ability to limit the set point of the secondary controller. In addition, by speeding up the loop response, the sensitivity of the primary process variable to process upsets is also reduced. Finally, the use of the secondary loop can reduce the effect of control valve sticking or actuator nonlinearity.

COMPONENTS OF THE CASCADE LOOP

The primary or outer control loop of a cascade system is usually provided with PI or PID control modes and is designed only *after* the secondary loop has already been designed. This is because the characteristics of the slave loop have an effect on the master loop.

For example, if the secondary measurement is nonlinear, such as the square root signal of an orifice-type flow sensor, it would produce a variable gain in the primary loop if the square root was not extracted.

For these reasons, the secondary loop requirements will be discussed first.

The Secondary Loop

Ideally, the secondary variable should be so selected as to split the process time delays approximately in half. This

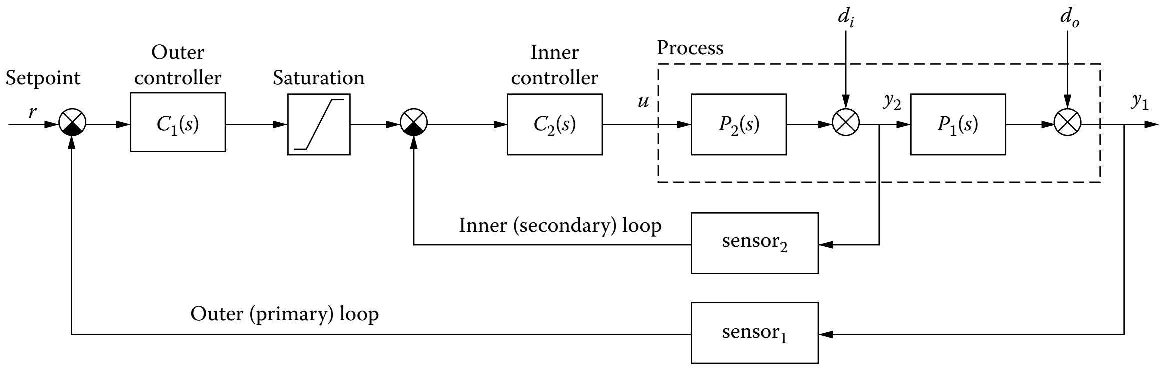

FIG. 2.6b
Block diagram of a cascade control system.

means that the secondary loop should be closed around half of the total time lags in the process. To demonstrate this need, consider two extreme cases:

1. If the secondary variable responded instantly to the manipulated variable (no time delay in the secondary loop), the secondary controller would accomplish nothing.
2. If the secondary loop was closed around the entire process, the primary controller would have no function.

Therefore, the dynamic elements of the process should be distributed as equitably as possible between the two controllers. When most of the process dynamics are enclosed in the secondary loop, that can cause problems. Although the response of the secondary loop is faster than the open loop configuration, in which a secondary controller does not even exist (Figure 2.6c), its dynamic gain is also higher, as indicated by the damped oscillation. This means that if stability is to be retained in a cascade configuration, the proportional band of the primary controller must be wider than it would be without a secondary loop; such de-tuning reduces responsiveness.

The right choice of the secondary variable will allow a reduction in the proportional band of the primary controller when the secondary loop is added because the high-gain region of the secondary loop lies beyond the natural frequency of the primary loop. In essence, reducing the response time of the secondary loop moves it out of resonance with the primary loop.

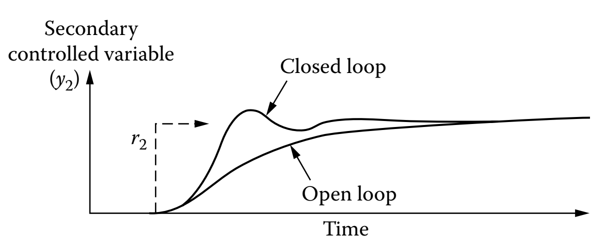

FIG. 2.6c
The response of the secondary variable is much improved when cascade control is used.

Secondary Control Variables

The most common types of secondary control variables are discussed next, discussed in the order of their frequency of application.

Valve Position Control (Positioner) The position assumed by the plug of a control valve is affected by forces other than the control signal, principally friction and line pressure. A change in line pressure can cause a change in the inner valve position and thereby upset a primary variable, and stem friction has an even more pronounced effect.

Friction produces hysteresis between the action of the control signal and its effect on the valve position. Hysteresis is a nonlinear dynamic element whose phase and gain vary with the amplitude of the control signal. Hysteresis always degrades performance, particularly where liquid level or gas pressure is being controlled with integral action in the controller.

The combination of the natural integration of the process, reset integration in the controller, and hysteresis can cause a "limit cycle" that is a constant-amplitude oscillation. Adjusting the controller settings will not dampen this limit cycle but will just change its amplitude and period. The only way of overcoming a limit cycle is to close the loop around the valve motor. This is what a positioner (a valve position controller) does; and this action can be considered to be the secondary loop in a cascade system.

Flow Control A cascade flow loop can overcome the effects of valve hysteresis as well as a positioner can. It also ensures that line pressure variations or undesirable valve characteristics will not affect the primary loop. For these reasons, in composition control systems, flow is usually set in cascade.

Cascade flow loops are also used where accurate manipulation of flow is mandatory, as in the feedforward systems shown in Section 2.8.

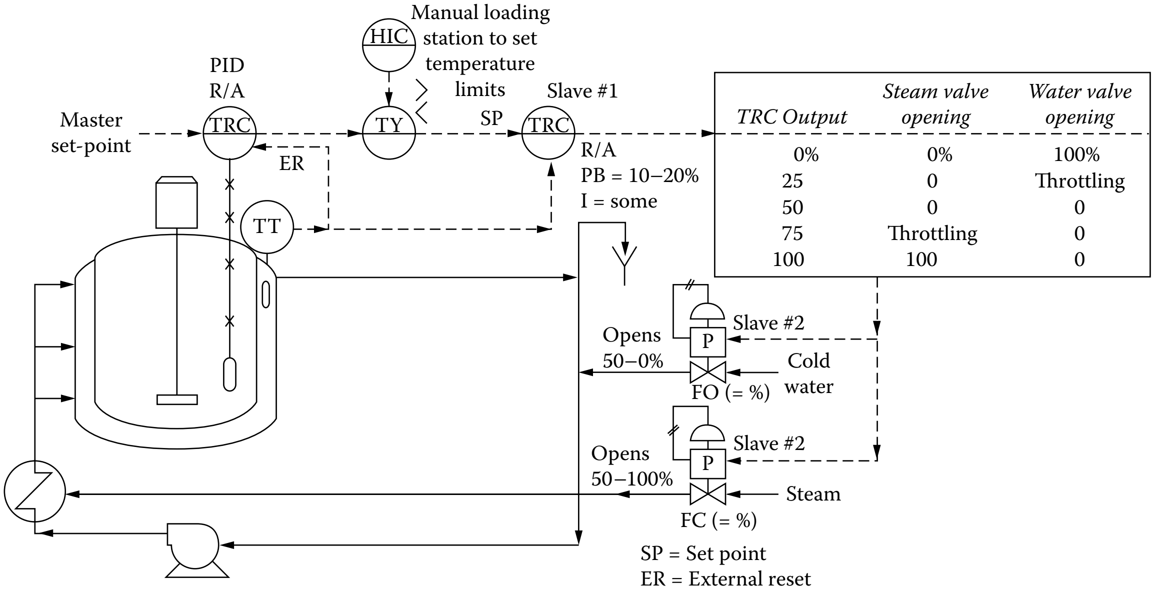

FIG. 2.6d
When cascade control is provided for a stirred reactor, it is recommended to provide the cascade master with a PID algorithm and with external reset from the measurement of the secondary controller. In this case, there are two inner loops; both can be plain proportional. The gain of the temperature controller is usually 5 to 10, while that of the valve positioner is about 20.

Temperature Control Chemical reactions are so sensitive to temperature that special consideration must be given to controlling the rate of heat transfer. The most commonly accepted configuration has the reactor temperature controlled by manipulating the coolant temperature in cascade. A typical cascade system for a stirred tank reactor is shown in Figure 2.6d.

Cascade control of coolant temperature at the exit of the jacket is much more effective than at the inlet because the dynamics of the jacket are thereby transferred from the primary to the secondary loop. Adding cascade control to this system can lower both the proportional band and reset time of the primary controller by a factor of two or more.

Since exothermic reactors require heating for startup as well as cooling during the reaction, heating and cooling valves must be operated in split range. The sequencing of the valves is ordinarily done with positioners, resulting in a second layer of inner control loops or cascade sub-loops in the system.

In Figure 2.6d, the secondary variable is the jacket temperature and the manipulated variables are the steam and cold water flows. Under control, the secondary variable will always come to rest sooner than the primary variable because initially the controller will demand a greater quantity of the manipulated variable than what is represented by the "equivalent" step change in the manipulated variable.

This is particularly true when the secondary part of the process contains a dominant lag (as opposed to dead time). Because with a dominant lag the gain of the secondary controller can be high, closing the loop is particularly effective. The dominant lag in the secondary loop of Figure 2.6d is the time lag associated with heat transfer across the jacket.

Secondary Control Modes Valve positioners are proportional controllers and are usually provided with a fixed band of about 5% (gain of 20). Flow controllers invariably have both proportional and integral modes. In temperature-on-temperature cascade systems, such as shown in Figure 2.6d, the secondary controller should have little or no integral. This is because reset is used to eliminate proportional offset, and in this situation a small amount of offset between the coolant temperature and its set point is inconsequential. Furthermore, integral adds the penalty of slowing the response of the secondary loop. The proportional band of the secondary temperature controller is usually as narrow as 10 to 15%. A secondary flow controller, however, with its proportional band exceeding 100%, does definitely require an integral mode.

Derivative action cannot be used in the secondary controller if it acts on set-point changes. Derivative action is designed to overcome some of the lag inside the control loop and if applied to the set-point changes, it results in excessive valve motion and overshoot. If the derivative mode acts only on the measurement input, such controllers can be used effectively in secondary loops if the measurement is sufficiently free of noise.

Cascade Primary Loop

Adding cascade control to a system can destabilize the primary loop if most of the process dynamics (time lags) are within the secondary loop. The most common example of this is using a valve positioner in a flow-control loop.

Closing the loop around the valve increases its dynamic gain so much that the proportional band of the flow controller may have to be increased by a factor of four to maintain

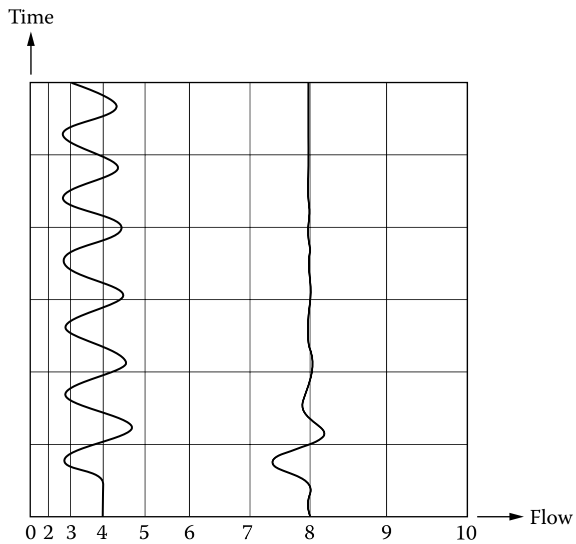

FIG. 2.6e
If cascade flow with a nonlinear orifice sensor is tuned at 80% of range, it will become unstable when the flow drops below 40%.

stability. The resulting wide proportional band means slower set-point response and deficient recovery from upsets. Therefore, in flow control loops, the existence of a large valve motor or long pneumatic transmission lines causes problems, and a volume booster rather than a positioner should be used to load the valve motor.

Instability Instability can also appear in a composition- or temperature-control system where flow is set in cascade. These variables ordinarily respond linearly to flow, but the nonlinear characteristic of a head flow meter can produce a variable gain in the primary loop.

Figure 2.6e compares the manipulated flow record following a load change that has occurred at 40% flow with a similar upset that took place at 80% flow. Differential pressure h is proportional to flow squared:

$$h = kF^2 \qquad \textbf{2.6(1)}$$

If the process is linear with flow, the loop gain will vary with flow, but when h is the manipulated variable, because flow is not linear with h:

$$\frac{dF}{dh} = \frac{1}{2kF} \qquad \textbf{2.6(2)}$$

Thus, if the primary controller is adjusted for 1/4-amplitude damping at 80% flow, the primary loop will be undamped at 40% flow and entirely unstable at lower rates.

Whenever a head-type flow meter provides the secondary measurement in a cascade system, a square-root extractor should be used to linearize the flow signal. The only exception to that rule is if flow will always be above 50% of full scale.

Saturation When both the primary and secondary controllers have automatic reset, a saturation problem can develop. Should the primary controller saturate, limits can be placed on its integral mode, or logic can be provided to inhibit automatic reset as is done for controllers on batch processes.

Saturation of the secondary controller poses another problem, however, because once the secondary loop is opened due to saturation, the primary controller will also saturate. A method for inhibiting reset action in the primary controller when the secondary loop is open (switched to manual) for any reason is shown in Figure 2.6f.

If the secondary controller has integral, its set point and measurement will be equal in the steady state, so the primary controller can be effectively reset by feeding back its own output or the secondary measurement signal. But if the secondary loop is open (in manual), so that its measurement no longer responds to its set point, the positive feedback loop to the primary controller will also open, inhibiting reset action.

Placing the dynamics of the secondary loop in the primary reset circuit is no detriment to control. In fact it tends to stabilize the primary loop by retarding the integral action. Figure 2.6f shows an application of this technique where the primary measurement is at the outlet of a secondary steam superheater, whereas the secondary measurement is at its inlet, downstream of the valve delivering water.

At low loads, no water is necessary to keep the secondary temperature at its set point, but the controllers must be prepared to act should the load suddenly increase. In this example the proportional band of the secondary controller is generally wide enough to require integral action.

When putting a cascade system into automatic operation, the secondary controller must first be transferred to automatic. The same is true in adjusting the control modes, insofar as the secondary should always be adjusted first, with the primary in manual.

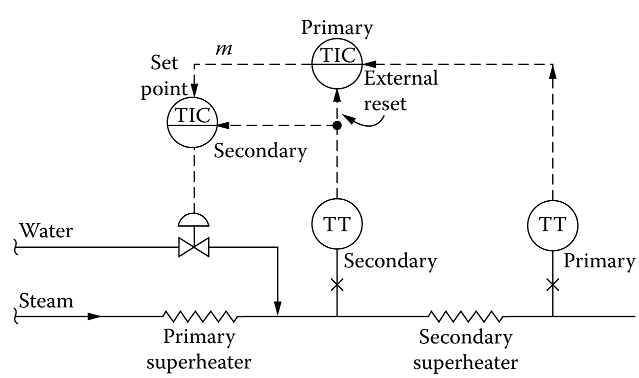

FIG. 2.6f
External reset is provided for the primary controller to prevent integral windup when the secondary controller is in manual.

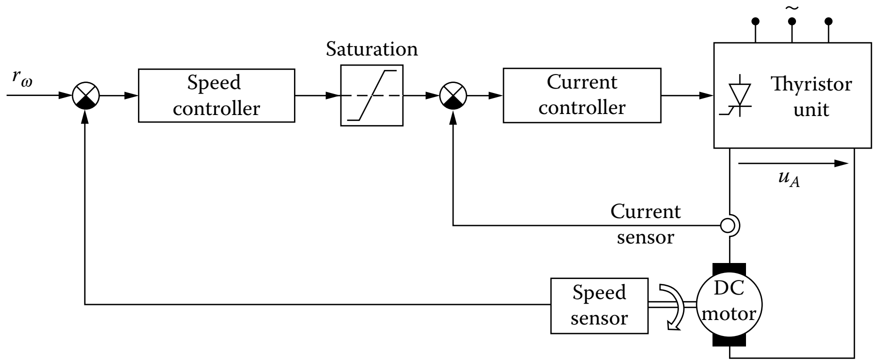

FIG. 2.6g
Cascade control of a DC motor.

Cascade Application Examples

Speed Control of a DC Motor When controlling the speed of a DC motor, the current can reach high peaks during the start-up and shut-down phases or when loading or breaking the motors. Therefore it is usual to apply cascade control with the armature current as the secondary variable. Saturation applied at the current set point (reference value) limits the maximum value of the current. The control system is shown in Figure 2.6g.

Another cascade control example is the control of electrical drives in controlling the position of servomechanisms with velocity feedback. Here the primary measured variable is the position, and the secondary measured variable is the velocity. The manipulated variable is the voltage applied to the servomotor.

Room Temperature Control A possible room temperature control configuration is shown in Figure 2.6h. The room is heated by a steam-heated heat exchanger, which warms the air supply to the room. The manipulated variable is the opening of the steam valve, which determines the steam flow through the heat exchanger. The primary variable is the room temperature. The variation in steam pressure can be the main source of upsets. The secondary variable is the inlet air temperature measured immediately after the heat exchanger, as the disturbances in the steam pressure affect it much sooner than they affect the room temperature.

Adding Valve Positioner to Tank Level Control When the liquid level in a tank is controlled by an outlet valve, placing a positioner on that valve forms a secondary cascade loop. This is desirable because the position of the valve's inner valve is affected by other factors besides the control signal, such as friction and inlet pressure. Variations in the line pressure can cause a change in position and thereby upset a primary variable. Stem friction can have an even more pronounced effect.

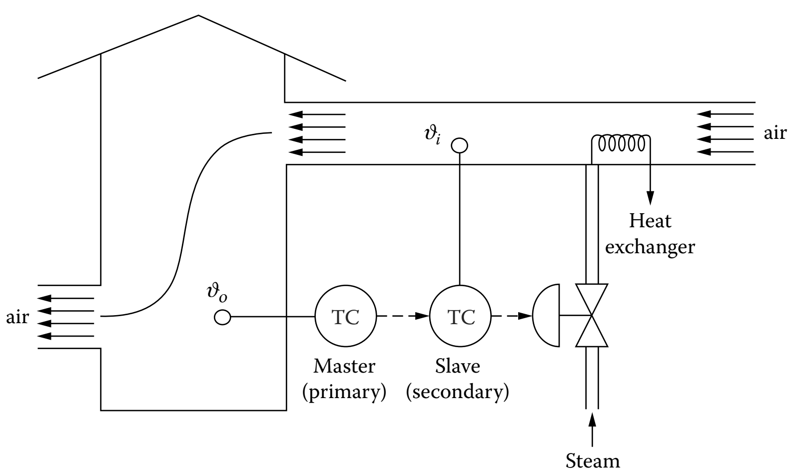

FIG. 2.6h
Cascade control of room temperature.

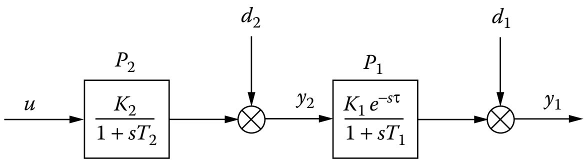

FIG. 2.6i
The process controlled by a cascade control system consists of two segments (P_1 and P_2) and has a measurable inner variable (y_2).

Inner valve sticking due to friction produces a square-loop hysteresis between the action of the control signal and its effect on the valve position. Hysteresis always degrades performance, particularly where liquid level is being controlled with integral action in the controller. The combination of the integrating nature of the process and the integral action in the controller with hysteresis causes a "limit cycle" that results in a constant-amplitude oscillation.

Adjusting the controller's tuning settings will not dampen this limit cycle but will only change its amplitude and period of oscillation. The only way of overcoming this limit cycle is to add a positioner and close the position-controlling slave loop around the valve motor.

Composition Control In composition control systems, flow is usually set in cascade. A cascade flow loop can overcome the effects of valve hysteresis. It also ensures that line pressure variations or undesirable valve characteristics will not affect the primary loop.

Cascade Controller Design and Simulation

The process controlled by a cascade system is shown in Figure 2.6i. The nominal values of the parameters and the uncertainties in their values are as follows:

$$K_{1n} = 1, \quad 0.8 < K_1 < 1.2 \quad K_{2n} = 1, \quad 0.8 < K_2 < 1.2$$
$$\tau_n = 20, \quad 18 < \tau < 22 \quad T_{2n} = 4, \quad 3 < T_2 < 5$$
$$T_{1n} = 10, \quad 8 < T_1 < 12$$

The dual design objectives are:

1. First design a single-series PID controller, with feedback taken from the output signal y_1. Design objectives are stability and good reference signal (set point) tracking properties characterized by zero steady-state error and about 60° of the phase margin to keep the overshoot below 10%, when a step change is introduced in the set point.
2. Design a cascade control system with feedback taken from the primary and secondary (outer and inner) controlled variables (process outputs) y_1 and y_2, respectively. Both control loops have to be stable. The design objectives for tracking the set point (reference signal) are the same as before. Fast response is also required to any upsets in the secondary loop or measurement.

The design should consider the theoretical process model, and the effects of the possible errors or uncertainties in the model should be checked.

Single-Series Primary Controller The block diagram of the single control loop is shown in Figure 2.6j. To ensure accurate tracking of the set point (step reference signal), the controller must have an integral mode. The addition of rate action can accelerate the system response. A PID controller combines these two effects. With pole-cancellation technique the suggested PID controller transfer function is

$$C(s) = A\frac{(1+10s)(1+4s)}{10s(1+s)} \qquad \textbf{2.6(3)}$$

The ratio of the time constants of the PD part was chosen to be four.

The loop transfer function is

$$L(s) = \frac{A}{10s(1+s)}e^{-20s} \qquad \textbf{2.6(4)}$$

Gain A of the controller has to be chosen to ensure the required phase margin. In case of systems with dead time τ, a phase margin of about 60° is obtained if the cut-off frequency ω_c in the loop's Bode amplitude-frequency diagram is at slope of –20 dB/decade and is located at about $1/2\tau$.

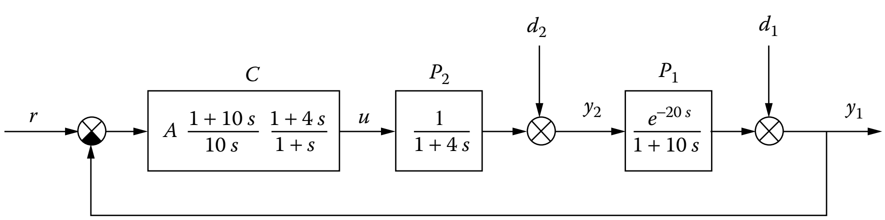

FIG. 2.6j
Series compensation.

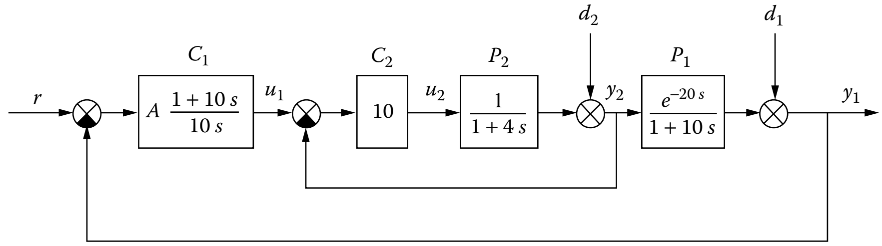

FIG. 2.6k
Cascade compensation.

In our case ω_c ~ 1/40. At the cut-off frequency the absolute value of the frequency function is 1, thus $A/10\omega_c = 1$. Hence $A = 0.25$.

Cascade Controller System Design The block diagram of the control system is shown in Figure 2.6k. First the inner or secondary control loop is designed for fast correction of upsets that might occur in the secondary loop. The inner (secondary) controller can be proportional only. Its gain can be 10 and the cut-off frequency of the secondary loop can be 2.5. In that case the settling time is expected around $3/\omega_c$.

The resulting transfer function of the inner closed loop between signals y_2 and u_1 is

$$\frac{0.909}{1+0.36s} \tag{2.6(5)}$$

In the primary (outer) loop a PI controller is used, which will be responsible for accurate steady-state tracking of a step change in the set point (reference signal). Again, pole-cancellation technique is applied. The transfer function of the controller is

$$C_1(s) = A\frac{1+10s}{10s} \tag{2.6(6)}$$

The loop transfer function of the outer loop is

$$L(s) = A\frac{0.909}{10s(1+0.36s)}e^{-20s} \tag{2.6(7)}$$

The cut-off frequency is chosen again to be $\omega_c = 1/40$. At this frequency the absolute value of the frequency function is 1.

$$\frac{0.909A}{10\omega_c} = 1, \qquad \text{hence } A = 0.275 \tag{2.6(8)}$$

Simulation Results The set-point (reference signal) change is a unit step that occurs at a time point of 10 sec. A unit step inner disturbance is added at a time point of 250 sec. The simulation ends at $t = 500$ sec.

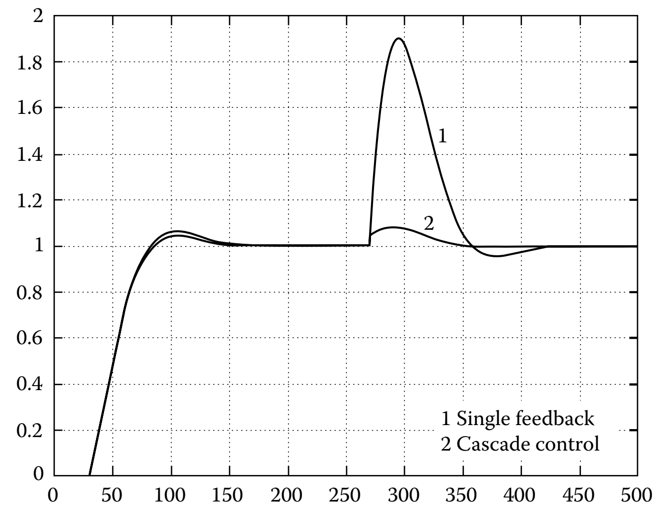

FIG. 2.6l
Tracking a unit change in set point that occurred at t = 10 sec and correcting for an upset of unit step size that occurred at t = 250 sec. Response 1 is that of a single feedback controller and 2 is that of a cascade system.

With the assumed nominal process model described earlier, the controlled variable responses (output signals) are shown in Figure 2.6l. It is seen that the set-point (reference signal) tracking is practically the same for both control configurations. On the other hand, the response to upsets (disturbance rejection) is much better with cascade control.

Figure 2.6m, in its upper part, shows the manipulated variable (u_2, controller output signal) and in the lower part, the controlled variable (y_2, inner variable) response of the secondary loop in the cascade configuration. The peak in the manipulated variable (secondary control signal) ensures the fast disturbance compensation. If the slave controlled variable (inner process variable) maximum value has to be under a given limit, this can be achieved by artificially saturating the output of the primary controller.

Figure 2.6n shows the effect of model parameter uncertainties in cases of the single loop and the cascade system. The controllers are designed based on the same nominal process model, and their responses to upsets are compared when the real process dynamics differ from the assumed

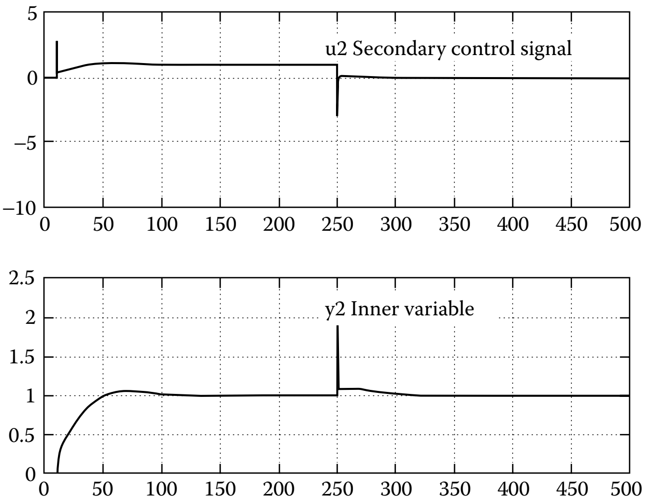

FIG. 2.6m
The response of the manipulated variable (u_2, controller output signal) and the controlled variable (y_2, inner variable) of the secondary loop in the cascade configuration.

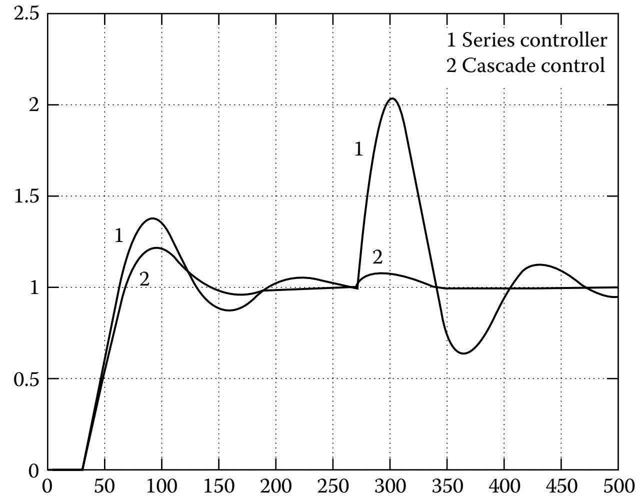

FIG. 2.6o
The controlled variable responses (output signals) of the single loop (series controller) and the cascade control system, when the time constants and the gain of the process are at their highest values.

nominal one that was used during tuning of the controllers (dynamics slower or faster) by various degrees.

Cascade control is more tolerant of model parameter uncertainties or errors in the model than a single controller and it gives particularly better performance when process upsets occur, as its disturbance rejection capability is superior to that of a single-loop controller.

Figure 2.6o describes the controlled variable responses of the single-loop and cascade control systems, assuming all dynamic parameters (time constants) of the process and its gain are both at their highest values.

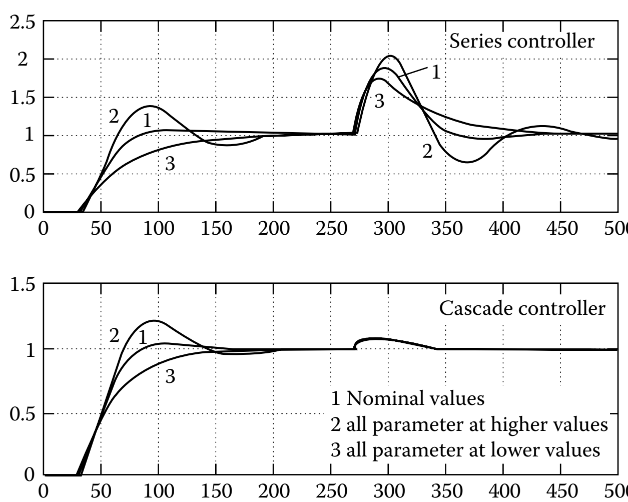

FIG. 2.6n
The response to both a set point step and a process upset of a single-loop controller (top) and a cascade system (bottom) to changes in the process dynamics. (1, correctly tuned; 2, process faster than at time of tuning; 3, process slower).

SUMMARY

Cascade loops consist of two or more controllers in series and have only a single, independently adjustable set point, that of the primary (master) controller. The main value of having secondary (slave) controllers is that they act as the first line of defense against disturbances, preventing these upsets from entering and upsetting the primary process.

For example, in Figure 2.6d, if there were no slave controller, an upset due to a change in steam pressure or water temperature would not be detected until it had upset the master measurement. In this configuration, the cascade slave detects the occurrence of such upsets and immediately counteracts them, so that the master measurement is not upset and the primary loop is not even aware that an upset occurred in the properties of the utilities.

In order for the cascade loop to be effective, it should be more responsive (faster) than the master. Some rules of thumb suggest that the slave's time constant should be under 1/4 to 1/10 that of the master loop and the slave's period of oscillation should be under 1/2 to 1/3 that of the master loop.

The goal is to distribute the time constants more-or-less evenly between the inner (slave or secondary) and outer (master or primary) loops, while making sure that the largest time constant is *not* placed within the inner loop. When that occurs, such as in the case where a valve positioner is the slave and a very fast loop such as a flow or liquid pressure controller is the master, stability will be sacrificed because the valve has the largest time constant in the loop. In such configurations, stability can be regained only at the cost of reduced control quality (sluggish response to load or set-point changes).

Therefore, in such cases one would try to either speed up the valve or avoid the use of cascade loops. If the reason for using a positioner is to increase the air flow to the actuator, one can replace the positioner with a booster relay.

Providing external reset (Figures 2.6d and 2.6f) for the cascade master from the slave measurement is always recommended. This guarantees bumpless transfer when the operator switches the loop from automatic control by the slave to full cascade control. The internal logic of the master controller algorithm is such that as long as its output signal (m) does not equal its external reset (ER), the value of m is set to be the sum of the external reset (ER) and the proportional correction ($K_c e$) only. When $m = \text{ER}$, the integral mode is activated, and in case of a PI controller, the output is:

$$m = K_c(e + 1/T_i e\, dt) + b \qquad \textbf{2.6(9)}$$

Bibliography

Åström, K. J., and Hägglund, T., *PID Controllers: Theory, Design and Tuning,* Research Triangle Park, NC: Instrument Society of America, 1995.

Buckley, P. S., *Techniques of Process Control,* Englewood Cliffs, NJ: John Wiley & Sons, 1964.

Corripio, A. B., *Tuning of Industrial Control Systems,* Research Triangle Park, NC: Instrument Society of America, 1990.

Gibson, J. E., *Nonlinear Automatic Control,* New York: McGraw-Hill, 1963.

Goodwin, G. C., Graebe, S. F., and Salgado, M. E., *Control System Design,* Englewood Cliffs, NJ: Prentice Hall, 2001.

Harriott, P., *Process Control,* New York: McGraw-Hill, 1964.

Jones, B. E., *Instrumentation, Measurement and Feedback,* New York: McGraw-Hill, 1977.

Levine, W. S. (Ed.), *The Control Handbook,* Boca Raton, FL: CRC Press, 1996.

Luyben, W. L., *Process Modeling, Simulation and Control for Chemical Engineers,* 2nd ed., New York: McGraw-Hill, 1990.

Luyben, W. L., and Luyben, M. L., *Essentials of Process Control,* New York: McGraw-Hill, 1997.

Morari, M., and Zafiriou, E., *Robust Process Control,* Englewood Cliffs, NJ: Prentice Hall, 1989.

Seborg, D. E., Edgar, T. F., and Mellichamp, D. A., *Process Dynamics and Control,* New York: John Wiley & Sons, 1989.

Shinskey, F. G., *Process Control Systems: Application, Design and Tuning,* 4th ed., New York: McGraw-Hill, 1998.

Skogestad, S., and Postlethwaite, I., *Multivariable Feedback Control,* New York: John Wiley & Sons, 1996.

Smith, C. A., and Corripio, A. B., *Principles and Practice of Automatic Process Control,* New York: John Wiley & Sons, 1985.

Stephanopoulos, G., *Chemical Process Control: An Introduction to Theory and Practice,* Englewood Cliffs, NJ: Prentice Hall, 1984.