Getting the Best Performance from Challenging Control Loops

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ABSTRACT

While many control loops are easy to tune and present almost no control problems, a few control loops can be very problematic and never seem to control right. These loops exhibit process control problems like oscillations, large deviations from set point, sluggish response to disturbances, spurious upsets, and inability to follow a set point. Control engineers and technicians can spend many hours or even days trying to improve the performance of these challenging control loops, but the results often remain unsatisfactory.

This paper explores various reasons why the optimization of some control loops can be very challenging, and explains how to get the best performance from these loops. It covers several problems originating from within the control loop, such as poor controller tuning, final control element problems, and instrumentation problems. It also covers problems originating outside of the control loop such as interactions and disturbances. The paper presents the appropriate corrective actions to take for solving each of the problems and improving control performance. These include tuning methods, control strategy design, calibration and maintenance practices, and equipment design.

To help with the diagnosis of control problems, the paper presents a fault tree that divides control problems into different categories. It provides steps to follow for differentiating between different problems and diagnosing the causes of poor control. It also presents corrective actions that should be taken to solve each of the problems. Examples from real process plants are given throughout the paper to illustrate many of the typical problems experienced and show improvement results possible through proper problem analysis and corrective action.

HIGH-LEVEL SYMPTOMS OF POOR CONTROL

Poor control loop performance often makes itself evident as excessively large deviations between the process variable and its set point. An operator may notice the deviations on his process trend displays or process alarm system and place the controller into manual mode in an attempt to stabilize the control loop. As a result, poor control loop performance may be indicated by excessive deviations between process variable and set point or controllers being in manual mode.

CONTROLLERS IN MANUAL

Some plants have 30 percent or more of their control loops in manual control mode. Being in manual mode does not necessarily mean the control loop performs poorly in automatic control mode. Many control loops are associated with redundant equipment, or certain operating modes, and can justifiably be in manual mode as a result. However, if a control loop is supposed to be in automatic control mode, but it is in manual mode, further investigation is required. It is important to talk to the operator to find out why the loop is in manual. Historical process trends of times when the loop was in automatic control can be reviewed to gain more insight into the problem, or the loop can be placed into automatic control mode to observe how it responds.

DEVIATIONS FROM SET POINT

If a controller is in automatic control mode and its process variable always remains acceptably close to its set point, there should be no need for concern. However, excessively large deviations between process variable and set point indicate poor control. The problem may be a constant or intermittent, and could originate from within or outside the control loop, but if large deviations occur, it provides grounds for further investigation. Large deviations between process variable and set point can be caused by a rapidly changing set point, process disturbances, loop nonlinearities, interactions, control element saturation, or poor controller tuning. The exact cause can be pinpointed through systematic analysis, using a fault analysis tree like the one below.

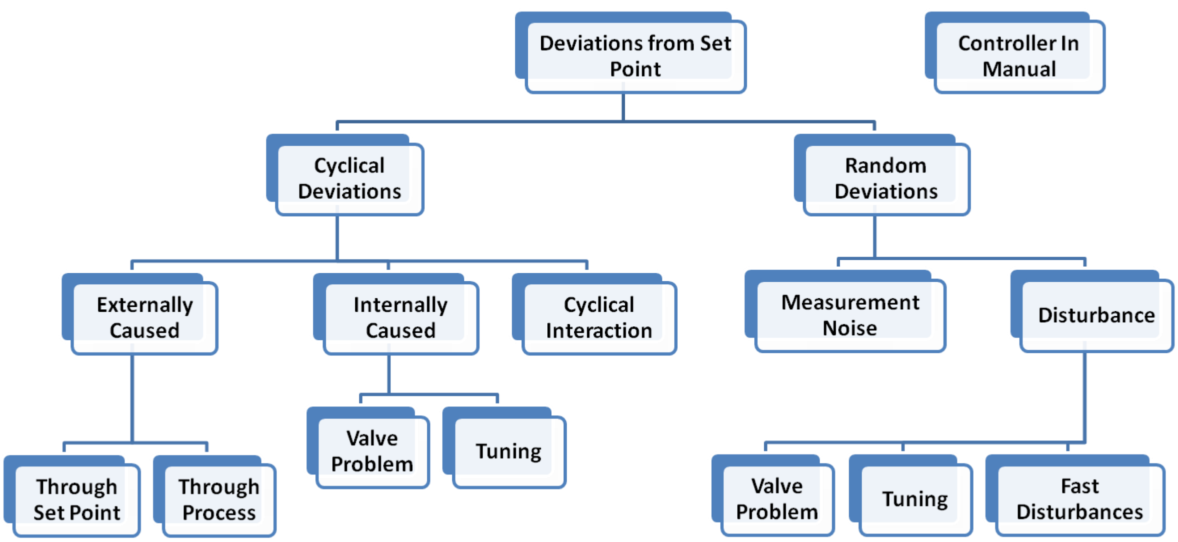

Figure 1. A fault analysis tree for systematically determining the cause of poor control.

CYCLICAL OR RANDOM DEVIATIONS

The first step in analyzing the control problem would be to look at the shape of the deviations on a trend plot: whether they are cyclical (oscillating) or random. This will determine the path for further analysis. Although sophisticated frequency analysis tools can be used to determine if deviations are cyclical or random, it can also be done quite simply by looking at historical time trends of the process variable. Once it has been determined if the deviations from set point are cyclical or not, the next level of analysis can be done.

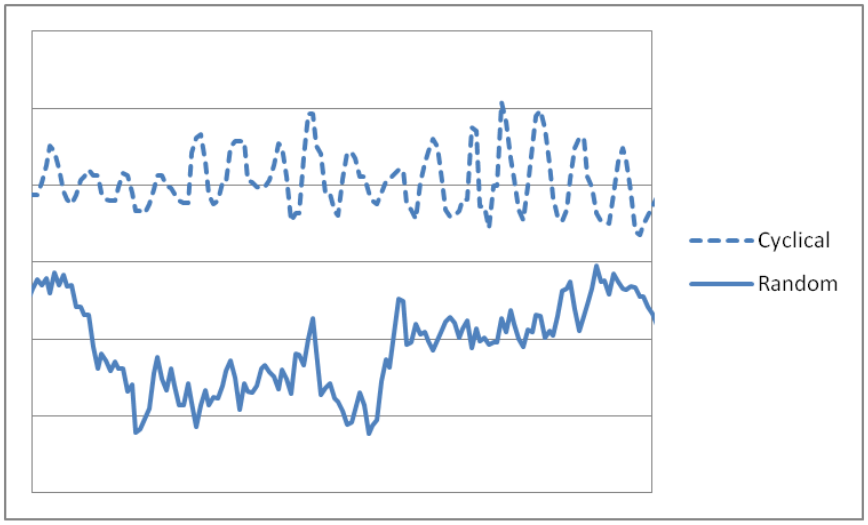

Figure 2. Process variables behaving cyclically and randomly.

CYCLICAL DEVIATIONS

Cyclical deviations (or oscillations) can appear as combinations of sine, saw-tooth, or square-wave patterns, sometimes intermixed with some randomness. Oscillations can originate from within the control loop or it could be caused by external factors. It could also be as a result of a cyclical interaction between two or more control loops. To narrow down the cause of the oscillation, the controller should be placed into manual mode to see if the oscillation would stop. If the oscillation persists when the controller is in manual mode, it originates from outside the loop.

EXTERNALLY-CAUSED OSCILLATIONS

An oscillation with its origin outside the control loop can influence the control loop through its set point or through the process.

OSCILLATING SET POINT

Oscillations entering the loop through the set point are easy to find – simply look at where the loop's set point is driven from. For example, if a steam flow control loop oscillates because its set point (coming from a temperature controller) is oscillating, the problem likely lies with the temperature controller. To verify, place the temperature controller in manual and see if the flow controller stops oscillating. The fault analysis should then be applied to the temperature controller.

OSCILLATING PROCESS

One oscillating loop can cause several other loops on the same plant to oscillate with it. The loops will all oscillate in harmony with the same period of oscillation.

Historical trends or process analysis software can be used to identify all the loops oscillating with the same period. The problem loop can then be isolated through knowledge of the process and its interactions, by looking at phase-shifts between oscillations, or by placing likely culprit loops in manual one at a time. If the loop driving the oscillations is placed into manual control mode, the oscillations will cease on all loops. That loop should then be analyzed further.

Note that this scenario is different from a cyclical interaction in which two or more control loops interact directly with each other in a cyclical fashion. In the case of a cyclical interaction, any of the participating control loops placed into manual will cause all loops to stop oscillating. This will be discussed later.

INTERNALLY-CAUSED OSCILLATIONS

Oscillations generated by a control loop itself can be caused by faulty final control element (e.g. control valve or damper) or by tuning. Generally, if the oscillation is caused by poor tuning, the process variable will oscillate with a reasonably smooth sine-wave pattern, and the oscillation will often grow in amplitude until either the controller output or the process variable periodically runs into its limits. If the oscillation is caused by final control element problems, the trends are more likely to be shaped like a square wave or saw tooth wave. However, this is a guideline and not a definitive test. If the control loop drives a final control element, the performance of the latter should be checked first by doing simple valve diagnostic tests before attempting to tune the controller. This is especially true if the control loop used to work properly, and is now oscillating without any changes to the controller settings.

The most common equipment-based causes of oscillations are control valve (or damper) related. Note that the discussions below sometimes mention only control valves for the sake of brevity. However, dampers can cause the same problems with the same symptoms as control valves. So where only control valves are mentioned, the same arguments will also apply to dampers.

STICTION

A common problem found in final control elements is stiction. This is short for *static friction*, and means that the valve internals are sticky. If the stem of a valve with stiction comes to rest, it tends to stick in that position. Additional force is then required to overcome the stiction.

The integral control mode will continue to change a controller's output in an attempt to get the process variable to its set point. While the valve is sticking, the process remains deviated from set point but additional pressure builds up in the valve actuator. If enough pressure has built up to overcome the static friction, the valve breaks free, and travels to the new controller output which is now far beyond its original value. This causes the process to overshoot its set point. Then the valve sticks at the new position, the controller output reverses its direction of travel and the whole process repeats in the opposite direction. This causes an oscillation, called a stick-slip cycle. If loop oscillations are caused by stiction, the controller output's cycle often resembles a saw-tooth wave, while the process variable may look like a square wave or an irregular sine wave.

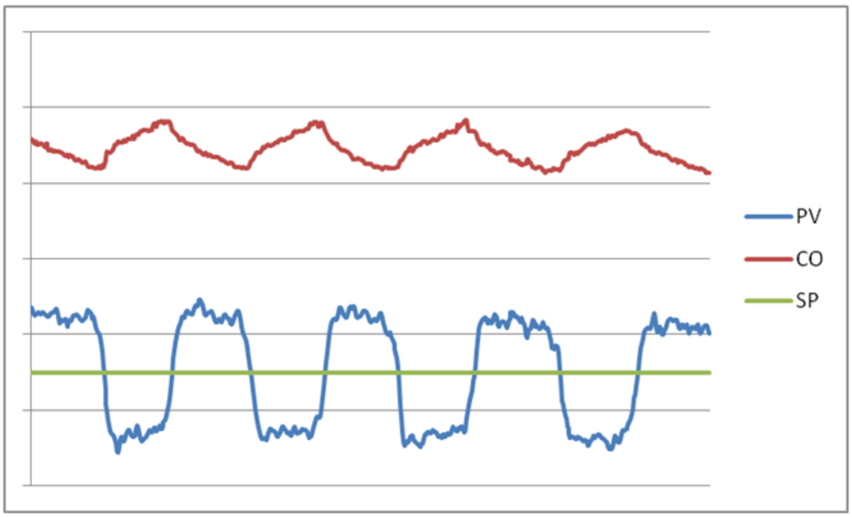

Figure 3. A flow loop with a stick-slip cycle. (PV is process variable, CO is controller output, and SP is set point.)

Stiction might be caused by an over-tight valve stem seal, by sticky valve internals, by an undersized actuator, or a faulty positioner. Stiction can be detected by placing the controller in manual mode and making small changes (0.5% is recommended) in controller output and monitoring the process variable for a resulting change. If the control valve seems to accumulate a few of the controller output changes before the process variable shows movement, it has stiction.

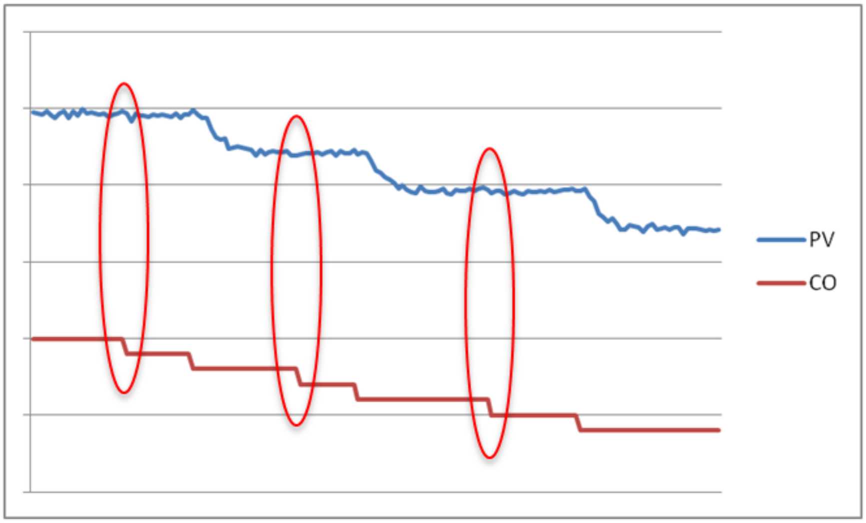

Figure 4. A stiction test revealing the presence of control valve stiction.

POSITIONER OVERSHOOT

Because of the widespread adoption of positioners for accurately positioning control valves and dampers, one problem that is more common now than a decade ago, is that of positioner overshoot. Positioners are fast feedback controllers mounted on the final control element to measure the valve stem or damper vane position and manipulate the actuator until the desired valve position is achieved. Most positioners can be tuned. Some are tuned too aggressively for the valve or damper they are controlling. This causes the device to overshoot its target position after a change in controller output.

The positioner and control valve form a closed-loop system that may cause the valve or damper position to hunt around or even oscillate. The positioner could simply be defective and cause the valve to overshoot. Or the valve can be sticky and the positioner is simply trying to overcome the stiction. If the controller on a fast-responding loop like a flow control loop is tuned aggressively, the combination with positioner overshoot can cause severe oscillations in the control loop. Positioner overshoot can be detected on fast-responding loops by placing the controller in manual and changing the controller output by two to five percent.

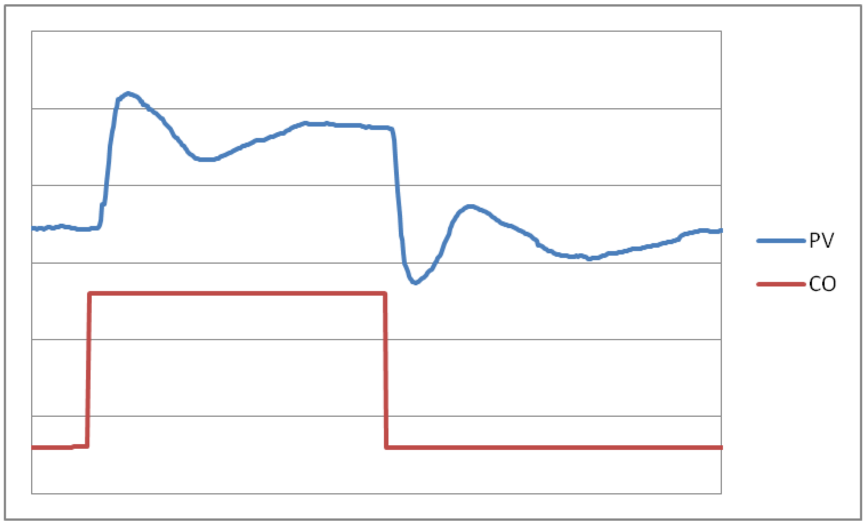

Figure 5. A faulty positioner causing the process variable to overshoot and hunt around after a change in controller output.

DEAD BAND AND LEVEL LOOPS

If a level controller drives a valve directly (i.e. no cascade control), and the valve has dead band, the loop will continuously oscillate. Valve dead band will be discussed later. A level control loop will also oscillate if the controller has an internal dead band around the set point. This is sometimes done to prevent the controller from reacting to process noise, but it should not be done on level loops because of the continuous oscillation it causes.

TUNING

A loop that is tuned too aggressively (overly fast response) can quickly develop oscillations. Step tests should be done on the process to determine the dominant process characteristics: process gain, dead time, and time constant. A step test is done by placing the controller into manual mode and changing its output by a few percent (between two and five percent is normally sufficient). Three or more step tests should be done to compare the results, throw out outliers, and use the average.

Proven, broad-spectrum tuning rules like the Cohen-Coon or Lambda tuning rules should be used to calculate new controller settings. However, many tuning rules are too aggressive in their original form, and it is recommended to use only half of the calculated controller gain. Best practices prescribe using tuning software for analyzing step-test data and calculating new controller settings.

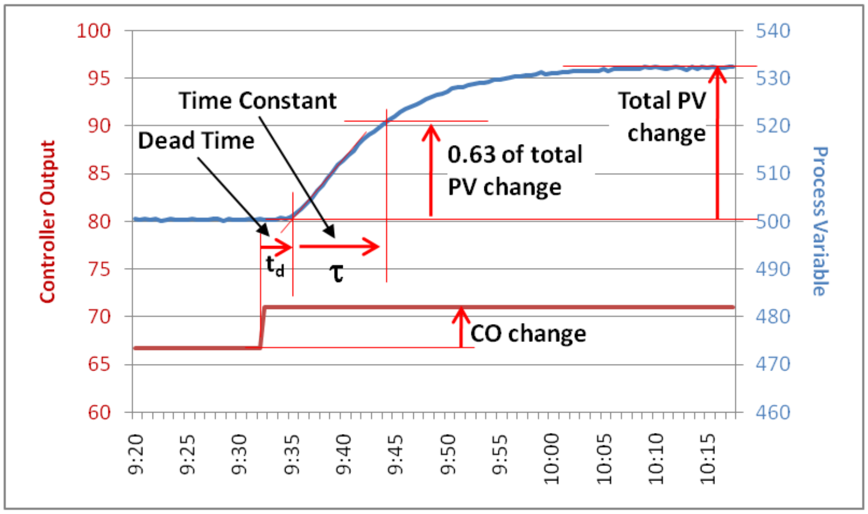

Figure 6. A step test showing measurements to be taken for controller tuning.

CYCLICAL INTERACTION

Interaction between loops with similar dynamics can cause the two loops to "fight" each other. One example of cyclical interactions is when liquid pressure and flow are controlled on the same line. This is often done with a pressure-reducing controller controlling one valve, and a downstream flow controller controlling another valve. Both controllers affect the flow and the pressure, and a cyclical interaction between the two loops can easily occur.

Cyclical interaction is aggravated by aggressive tuning, for example when using the Ziegler-Nichols or Cohen-Coon tuning rules.

To solve problems with cyclical interactions, control loops have to be tuned less aggressively. Using the Lambda method results in very stable control loops. One can think of highly interactive control loops as a tub filled with water. If you drop a stone in the tub, lots of waves result that take a while to stabilize. Using the Lambda tuning method is like replacing the water with oil. Now if you drop the stone into the tub, the oil just absorbs the disturbance and no persistent wave action results. The pulp and paper industry has highly interactive processes, and it has had great success using the Lambda tuning method.

RANDOM DEVIATIONS

In contrast to oscillations that are periodic, poor control can also make itself evident in large but random deviations between the process variable and set point. These could be measurement noise, process disturbances or rapid set point changes. To understand how a control loop is capable of handling disturbances, we need to look at the speed of response of a control loop.

LOOP SETTLING TIME

There are several measurements for loop response; settling time will be used here. Settling time can be defined as the duration of time during which a deviation between set point and process variable is more than 5% of the size of the deviation. The settling time of a control loop cannot be infinitely short. If a control loop is tuned sluggishly, it will have a long settling time. If the tuning is improved, the settling time will be reduced, but only up to a point. If the tuning is made any faster, the loop will become cyclical and the settling time will increase.

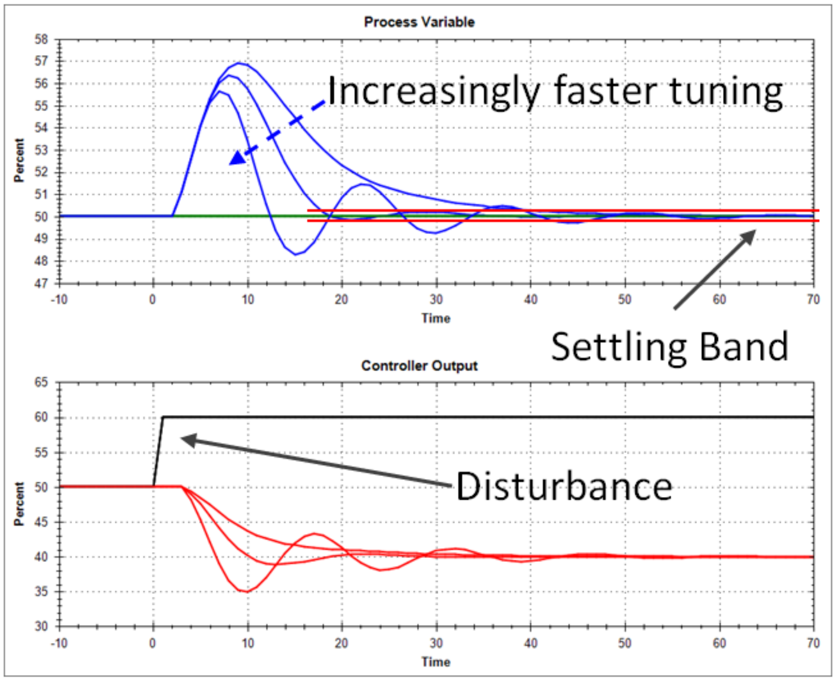

Figure 7. The settling time of a control loop has a minimum limit.

The minimum settling time of a control loop is determined mostly by the amount of dead time in the process. For a flow loop, the settling time is about three times the dead time, for a temperature loop it's between three and four dead times, and for a level loop it is about four dead times.

MEASUREMENT NOISE

Measurement noise is random, rapidly-changing deviations from set point. The rate at which this happens is so much shorter than the loop settling time that it is impossible for the controller to eliminate noise or even reduce its amplitude. A controller responding aggressively to noise will likely increase the average deviation size. The amplitude of noise can be reduced through filtering the process variable with a first-order lag filter or a moving-average filter. It is important to note that a filter increases the apparent dead time of a loop and therefore increases its settling time. Filtering should be applied only when needed, and then as little of it as possible.

DISTURBANCES

A process disturbance can push the process variable away from its set point. Disturbances are often the nemesis of good loop performance. As described above, feedback control is limited in how fast it can eliminate the effects of a disturbance and bring the process back to set point. If a disturbance occurs much slower than the settling time of a control loop, feedback control should be able to significantly reduce its amplitude. If not, it may be a problem with the final control element or the tuning of the controller. One should first check for final control element problems before tuning the controller.

DEAD BAND

Dead band (sometimes called hysteresis), reduces the effectiveness with which a controller can counteract disturbances. Every time the direction of a disturbance changes, the controller output has to traverse the entire dead band before the final control element begins moving. Dead band can be detected very reliably with a simple process test consisting of two controller output steps in one direction and one step in the opposite direction with the controller in manual mode. The second and last steps should be the same size. If the process variable does not reach the same level after the first and third steps, it indicates the presence of dead band. Dead band is a mechanical problem and cannot be addressed with tuning.

Figure 8. A dead-band test revealing the presence of dead band.

SATURATION AND RATE-OF-CHANGE LIMITS

A control loop may also appear to have sluggish response if the controller output becomes saturated at its upper or lower limit. If the controller output is constrained by a rate-of-change limiter, it also may cause sluggish response regardless of how well the controller is tuned. Alarms can warn of these conditions or historical time-trends of the controller output and process variable can be reviewed to find their presence.

TUNING

Once the final control element has a clean bill of health, the controller tuning should be reviewed to see if it is perhaps sluggish tuning that reduces the controller's effectiveness in counteracting disturbances.

The controller-tuning advice given earlier applies here too. Note that the Lambda tuning method results in stable control loops, but often cause a sluggish response to disturbances, especially on slow temperature loops. Cohen-Coon tuning provides faster disturbance rejection.

Although correct tuning methods can go a long way in minimizing the effects of disturbances, disturbances sometimes happen so rapidly that feedback control alone is unable to reduce their effects to reasonable levels. The fastest possible feedback control action is limited by the makeup of the process dynamics. Once this limit has been reached, other solutions must be sought to obtain further improvement in performance. It is sad to hear of personnel spending days and even weeks tweaking a control loop that is already at the limit of its performance capability.

CASCADE AND FEEDFORWARD CONTROLS

If a disturbance occurs faster than the control loop can respond, there is very little the controller can do to reduce its amplitude. In cases like this the feedback controller can be greatly augmented with cascade and feedforward control.

Cascade control should be applied whenever a slow-responding loop like temperature or level controls liquid or gas flow. The flow should be controlled directly with a flow controller of which the set point is set by the level or temperature controller.

Feedforward control should be applied when large, measureable disturbances affect the process variable. The feedforward controller uses the disturbance measurement to generate an equal but opposing control action to minimize the effect of the disturbance on the process variable.

INTERMITTENT PROBLEMS

Some control problems seem to come and go with time – intermittent problems. These problems are more difficult to track down and solve, but it helps to know what causes to look for.

NONLINEAR VALVE CHARACTERISTIC

Many control valves and most dampers have a nonlinear installed characteristic. This means that the flow characteristic of the device changes depending on how much open it is. If tuning is done with the valve or damper at the one end of its travel, the settings might not work at the other end and could cause oscillations or sluggish behavior. If this is the case, a function generator (X-Y curve) can be placed in the path of the controller output to cancel out the control valve or damper nonlinearity.

NONLINEAR PROCESS

Many processes react differently based on operating conditions, product type, production rate, etc. In many cases the differences in process characteristics are large enough to affect control loop performance. For example, the process gain of a heat exchanger is much less at high product flow rates compared to low flow rates. These changes in process characteristics often require different tuning settings for optimal control at various operating conditions. However, this is seldom implemented, leaving the control loop with poor response for most of its operating range. On systems with varying process characteristics, controller tuning should be altered automatically based on the operating conditions. This is accomplished quite effectively by implementing gain scheduling. Gain scheduling uses the operating condition as an input to one or more function generators to dynamically adjust the controller gain, and sometimes also the integral time, and derivative time if used.

CONCLUSION

Obtaining robust and optimally-performing control loops can be challenging at times. Control loops can perform sub-optimally due to a variety of reasons and controller tuning alone is in many cases not the ultimate solution for poor control performance. Through a simple but systematic analysis of the control problem, the root cause of poor control can be established and the problem can be resolved or at least minimized in the most effective way.