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Good Tuning: A Pocket Guide

By Gregory K. McMillan

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Research Triangle Park, NC 27709

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1.0—Best of the Basics

1.1 Introduction

Welcome to the wonderful world of proportional-integral-derivative (PID) controllers. This guide will cover the key points of good tuning and provide more than thirty rules of thumb. First, let's blow away some myths:

Myth 1 – It is always best to use one controller tuning method. False; the diversity of processes, control valves, control algorithms, and objectives makes this impractical.

Myth 2 - Controller tuning settings can be computed precisely. Not so. The variability and nonlinearity in nearly all processes and control valves makes this implausible. Any effort to get much more than one significant digit is questionable because any match to the plant is momentary. If you run a test for an auto tuner or manually compute settings ten times, you should expect ten different answers.

Roughly 75 percent of process control loops cause more variability running in the automatic mode than they do in the manual mode. A third of them oscillate as a result of nonlinearities such as valve dead band. Another third oscillate because of poor controller tuning. The remaining loops oscillate because of deficiencies in the control strategy. A well-designed control loop with proper tuning and a responsive control valve can minimize this variability. Because this means you can operate closer to constraints, good tuning can translate into increased production and profitability.

1.2 Actions Speak Louder than Words

The very first settings that must be right are the controller and valve actions. If these actions are not right, nothing else matters. The controller output will run off scale in the wrong direction regardless of the tuning settings.

The controller action sets the direction of a change in controller output from its proportional mode for every change in the controller's process variable (feedback measurement). If you choose *direct* action, an *increase* in process variable (PV) measurement will cause an *increase* in controller output that is proportional to its gain setting. Since the controller action must be the opposite of process action to provide feedback correction, you should use a direct-acting controller for a reverse-acting process except as noted later in this guide. Correspondingly, you should select *reverse* control action for a *direct*-acting process so an *increase* in process variable measurement will cause a *decrease* in controller output that is proportional to its gain setting, except as noted later. A *direct*-acting process is one in which the direction of the change in the process variable is the same as the direction of the change in the manipulated variable. A reverse-acting process is one in which the direction of the change in the process variable is opposite the direction of the change in the manipulated variable. The manipulated variable is most frequently the flow through a control valve, but it can also be the set point of a slave loop for a cascade control system or variable speed drive.

The valve action sets the display. For example, it determines whether a 100 percent output signal corresponds to a wide open or a fully closed valve. It also determines the direction of a change in the actual signal to the control valve when there is a change in the controller's output. In some analog controllers developed in the 1970s, such as the Fisher AC^2, the valve action affected only the display, not the actual signal. To compensate for this lack of signal reversal for a reverse-acting valve (i.e., an increase-to-close or fail-open valve), the control action had to be the opposite of the action that would normally be appropriate based on process action alone. Fortunately, the valve action corrects both the display and the actual valve signal in modern controllers, so the control action can be based solely on process action. However, the user should verify this before commissioning any loops. In control systems that use fieldbus blocks, the valve action should be set in the analog output (AO) block rather than in the PID controller block. This ensures that the "back-calculate" feature is operational for any function blocks (split range, characterization, and signal selection) that are connected between the PID and AO blocks. The signal can also be reversed in the current-to-pneumatic transducer (I/P) or in the positioner for a control valve. Before the advent of the smart positioner, it was preferable for the sake of visibility and maintainability that any reversal be done in the control room rather than at the valve. It is important to standardize on the location of the signal reversal to ensure that it is done and done only once. Table 1 summarizes how the controller action depends upon both the process and valve actions and on the signal reversal.

Table 1 – Controller Action

Process Action	**Valve Action**	**Signal Reversal**	**Controller Action**
Direct	Increase-Open	No	Reverse
Reverse	Increase-Open	No	Direct
Direct	Increase-Close	Yes	Reverse
Reverse	Increase-Close	Yes	Direct
Direct	Increase-Close	No	Direct
Reverse	Increase-Close	No	Reverse

Which brings us to rule of thumb number one.

Rule 1 – **The controller action should be the opposite of the process action unless there is an increase-to-close (fail-open) control valve for which there is no reversal of the valve signal.** This means that you should use reverse and direct-acting controllers for direct and reverse-acting processes,

respectively. The valve signal can be reversed for a fail-open valve at many places, but it is best done in the AO block of the control system.

1.3 Controller à la Mode

The names for the operational modes of the PID vary from manufacturer to manufacturer. Thus, the Foundation Fieldbus modes listed next provide a uniformity that can be appreciated by all.

- **Auto** (automatic) – The operator locally sets the set point. PID action is active (closed loop). In older systems, this mode is also known as the *local mode*.
- **Cas** (cascade) – The set point comes from another loop. PID action is active (closed loop). This is also known in older systems as the remote mode or remote set point (*RSP*).
- **LO** (local override) – PID action is suspended. The controller output tracks an external signal to position the valve. This mode is typically used for auto tuning or to coordinate the loop with interlocks. In older systems, it is also known as *output tracking*.
- **Man** (manual) – The operator manually sets the output. PID action is suspended (open loop).

- **IMan** (initialization manual) – PID action is suspended because of an interruption in the forward path of the controller output. This is typically caused by a downstream block that is not in the cascade mode. The controller output is back-calculated to provide bumpless transfer.
- **RCas** (remote cascade) – The set point is remotely set, often by another computer. PID action is active (closed loop). This mode is also known in older systems as the *supervisory* mode.
- **ROut** (remote output) – The output is remotely set, often by a sequence or by another computer. PID action is suspended (open loop). In older systems, this mode is known as *direct digital control* (*DDC*).

1.4 Is That Your Final Response?

The contribution that the proportional action makes to the controller output is the error multiplied by the gain setting. The contribution made by the integral action is the integrated error multiplied by the reset and gain setting. The reset setting is repeats per minute and is the inverse of integral time or reset time (minutes per repeat). Foundation Fieldbus will standardize the reset time setting as seconds per repeat and the rate (derivative) time setting as seconds. The contribution made by derivative mode is the rate of change of the error or process variable in percent (%), depending upon the type of algorithm, multiplied by the rate and gain settings.

When derivative action is on the process variable instead of on the control error, it works against a set point change. (The control error is the difference between the process variable and the set point.) The reason for this is that it doesn't know the process variable should be changing initially and that the brakes should only be applied to the process when it approaches set point. Using derivation action that is based on the change in control error will provide a faster initial takeoff and will suppress overshoot for a set point change. This is particularly advantageous for set points driven for batch control, advanced control, or cascade control. The improvement can translate into shorter cycle or transition times, an enhancement of the ability of slave loops to mitigate upsets, and less off-spec product because overshoot has been diminished.

When derivative action is used with a time constant there is a built-in filter that is about one-eighth (1/8) of the rate setting. However, you should use set point velocity limits to prevent a jolt to the output when there is a large step change in error from a manually entered set point. This is particularly important when you are using large gains or derivative action based on control error.

The PID algorithm uses percentage (%) input and output signals rather than engineering units. Thus, if you double the scale span of the input (error or process variable), you effectively halve the PID action. Correspondingly, if you double the scale span of the output (manipulated variable), you double the PID action. In fieldbus blocks, both input and output signals can be scaled with engineering units. Using output signal scaling will facilitate the manipulation of the slave loop's set point for cascade control. For controllers that use proportional band, you need to divide the proportional band into 100 percent to get the equivalent controller gain. Proportional band is the percentage change in error needed to cause a 100 percent change in output.

Proportional mode is expressed by the following equation (note that adjustments are gain or proportional band):

$$P_n = K_c * E_n$$

When the derivative mode acts on the process variable in percent (%PV), the equation becomes as follows:

$$P_n = K_c * D_n$$

Integral mode is expressed by the following equation (note that adjustments are integral time or reset):

$$I_n = K_c * 1/T_i * (E_n * T_s) + I_{n-1}$$

When the derivative mode acts on %PV, the equation becomes as follows:

$$I_n = K_c * 1/T_i * (D_n * T_s) + I_{n-1}$$

The equation for derivative mode is as follows (adjustments are derivative time or rate):

$$D_n = K_c * T_d * (E_n - E_{n-1}) / T_s$$

Or for derivative mode on %PV:

$$D_n = K_c * T_d * (\%PV_n - \%PV_{n-1}) / T_s$$

Note the inverse relationships across controllers!

$$\text{Gain } (K_c) = 100\% / PB$$

Reset action (repeats/minute) = $60 / T_i$

Where:

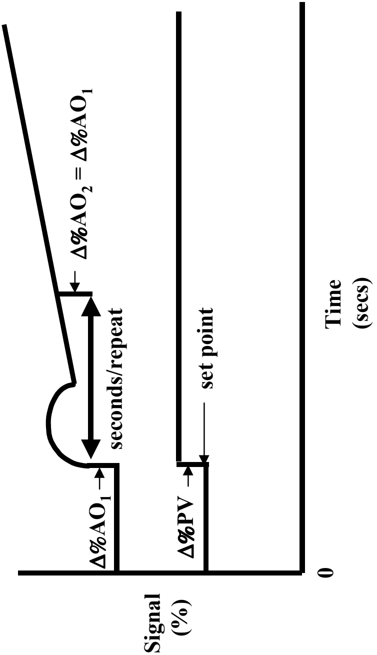
E_n = error at scan n (%)
K_c = controller gain (dimensionless)
PB = controller proportional band (%)
P_n = contribution of the proportional mode at scan n (%)
$\%PV_n$ = process variable at scan n (%)
I_n = contribution of the integral mode at scan n (%)
D_n = contribution of the derivative mode at scan n (%)
T_d = derivative (rate time constant) time setting (seconds)
T_i = integral time or reset time setting (seconds/repeat)
T_s = scan time or update time of PID controller (seconds)

Rule 2 – **If you halve the scale span of a controlled (process) variable or double the span of a manipulated variable (i.e., set point scale or linear valve size), you need to halve the controller gain to get the same PID action.** For controlled variables, the PID gain is proportional to the measurement calibration span. Often these spans are narrowed since accuracy is a percentage of span. For con-

trol valves, the PID gain setting is inversely proportional to the slope of the installed valve characteristic at your operating point. If a valve is sized too small or too large, the operating point ends up on the flat portion of the installed valve characteristic curve. For butterfly valves, the curve gets excessively flat below 15 percent and above 55 percent of valve position.

Figure 1 shows the combined response of the PID controller modes to a step change in the process variable (%PV). The proportional mode provides a step change in the analog output ($\Delta\%AO_1$). If there is no further change in the %PV, there is no additional change in the output even though there is a persistent error (offset). The size of the offset is inversely proportional to the controller gain. Integral action will ramp the output unless the error is zero. Since the error is hardly ever exactly zero, reset is always driving the output. The contribution made by the integral mode will equal the contribution made by the proportional mode in the integral time ($\Delta\%AO_2 = \Delta\%AO_1$). Hence, the integral time setting is the time it takes to repeat the proportional contribution (seconds per repeat). The contribution made by the derivative mode for a step change is a hump because of the built-in filter, which is about one-eighth (1/8) of the rate

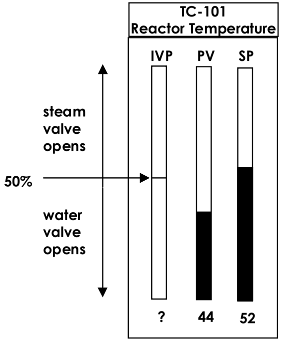
Figure 1 – PID Response

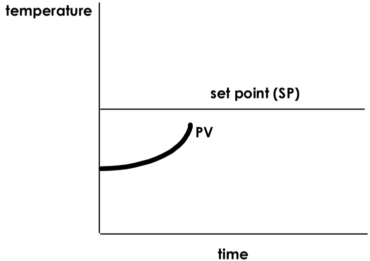


setting. Otherwise, there would be a spike in the output.

If the temperature is below set point for a reactor, as shown in Figure 2, should the steam or water valve be open? After looking at a faceplate or the digital value for temperature on a graphic display, most people think the steam valve is open when the temperature is below set point. Reset provides a direction of action that is consistent with human expectation. However, the proper direction for a change in controller output and the split-ranged control valve depends upon the trajectory of the process variable (PV). If the temperature is rapidly and sharply increasing, the coolant valve should be opening. Gain and rate action will recognize that a set point is being approached and position the valves correctly to prevent overshoot. In contrast, reset has no sense of direction and sacrifices future results for immediate satisfaction. Most reactors, evaporators, crystallizers, and columns in the process industry have too much reset. On two separate applications in chemical plants, for example, it was reported that the controller was seriously malfunctioning because the wrong valve was supposedly open, when in reality it was just gain and rate doing their job to prevent overshoot.

Figure 2 – Reset Gives Them What They Want





Reset action integrates the numeric difference between the PV and SP seen by operator on a loop faceplate Reset works to open the steam valve (reset has no sense of direction).

Proportional and rate action see the trajectory visible in a trend! Both would work to open the water valve to prevent overshoot.

Figures 3a and 3b show the effect of gain setting on a set point response. If the gain is too small, the approach to set point is too slow. If the gain is too large, the response will develop oscillations. If the dead time is much larger than the time constant and the gain action is too large compared to the reset action, the response will momentarily flatten out (falter or hesitate) well below the set point. Loops that are dead-time dominant tend to have too much gain and not enough reset action. For nearly all other loops in the process industry, the opposite is true. The size of the dead time relative to the time constant and the degree of self-regulation (ability to reach a steady state when the loop is in manual) determine which tuning methods you should employ.

It is particularly important to maximize the gain so you can achieve tight control of loops that can ramp away from set point, such as gas pressure. Maximizing the gain is also important for loops that can run away from set point, such as an exothermic reactor temperature. Finally, gain must be maximized to speed up the set point response for advanced control, batch control, and cascade control as well as for startup sequences. However, gain readily passes variability in the process variable to the output, which can increase overall variability and interaction.

Figure 3a – Effect of Gain Setting on Set Point Response for a Time Constant Much Larger than Dead Time

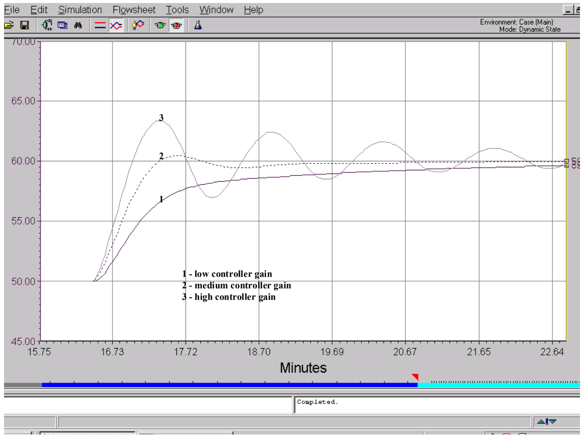
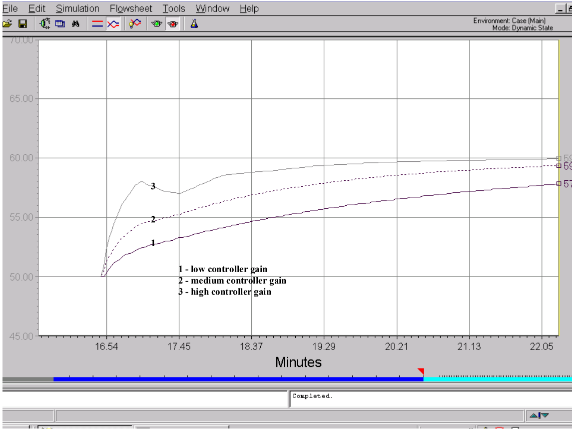


Figure 3b – Effect of Gain Setting on Set Point Response for a Dead Time Much Larger than Time Constant (Dead Time Dominant)



You must set the process variable filter to ensure that fluctuations from measurement noise stay within the dead band of the control valve. The variability amplification by gain can be particularly troublesome if there is inadequate smoothing because of insufficient volumes or mixing. These loops tend to be dead-time dominant, and they exhibit abrupt responses (i.e., no smoothing by a time constant). Thus, you should reduce the gain setting for loops on pipelines, desuperheaters, plug flow reactors, heat exchangers, static mixers, conveyors, spinning (fibers), and sheets (webs).

Rule 3 – **Increase the gain to achieve tight level, stirred reactor, or column control and to speed up the set point response for advanced, batch, cascade, and sequence control.** For high gains it is especially important that you set a process variable filter to ensure that fluctuations in the PID output are smaller than the valve dead band. You should also establish set point velocity limits so the output doesn't jerk when the operator changes the set point.

Rule 4 – **Decrease the gain to provide a smoother, slower, and more stable response; reduce interaction between loops; and reduce the amplification of variability.** Pulp and paper, fiber, and sheet processes use less gain (higher proportional band) and more reset action (smaller reset time). They also are much more sensitive to variability from valve dead band. The dead time is larger than the process time constant for most of these loops.

Rule 5 – **For level control of surge tanks, use a gain, such as 1.5, that is just large enough to avoid actuating the alarms set to help prevent the tank from overflowing or running dry.** You can estimate the gain as follows: it is the maximum needed change in controller output divided by the difference between the high- and low-level alarms. For example, if a 60 percent change in controller output will always keep the level between 30 percent and 70 percent, the correct controller gain is 1.5. You can add an error-squared algorithm, where gain is proportional to error, to effectively attenuate control action near the set point. This is a generally effective way to reduce the effect of noise for all types of level control loops.

Rule 6 - **For residence time control and for material balance control, use as high a gain for level control as possible, such as 5.0, that does not wear out the valves or upset other loops.** Continuous reactors, evaporators, crystallizers, and thermosyphon reboilers often need to maintain a level accurately, particularly if they are being pushed beyond nameplate production rates. Level control also needs to be tight when it manipulates a reflux flow for a column or a reactant makeup flow for a recycle tank.

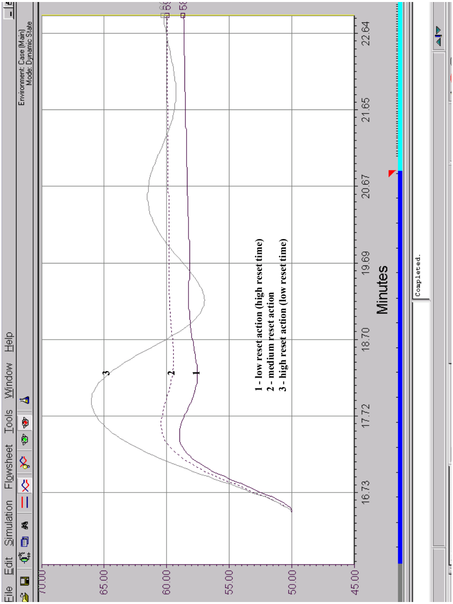
The Ziegler-Nichols (J. G. Ziegler and N. B. Nichols, "Optimum Settings for Automatic Controllers," *Transactions of the ASME*, vol. 64, Nov. 1942, p. 759) tuning method depicts a goal of quarter-amplitude response where each subsequent peak in an oscillatory response of the PID controller is one-fourth of the previous peak. This goal will generally provide a minimum peak error, but the controller is too close to the edge of instability. An increase of just 25 percent in the process gain or dead time can cause severe oscillations. By simply cutting the controller gain in half you will make Ziegler-Nichols tuning sufficiently robust. You will also make it look more like Internal Model Control (IMC) tuning for stirred reactors, evaporators, crystallizers, and columns. If the controller manipulates a control valve instead of a flow set point, you need to reduce the gain for the steepest part of the installed valve characteristic curve.

For loops that will reach a steady state when put in manual, decreasing the gain will always improve their stability. For exothermic reactors where a runaway condition can develop, there is a window of allowable gains in which instability can be caused by a gain that is too small or too large. For level loops with reset action, there is also a window in which a gain that is too small can cause slow, nearly sustained oscillations.

Rule 7 – **If you have a quarter-amplitude response, you should decrease the gain to promote stability.** Generally, halving the controller gain is sufficient. This assumes that the reset action is not excessive (i.e., integral or reset time is too small). If the manipulated variable is a control valve and the operating point moves from the flat to the steep portion of the installed characteristic, you may need to cut the gain by a factor of three or four.

Figure 4 shows the effect of reset setting on a set point response. If the reset action is too small (i.e., the reset time too large), it takes a long time

Figure 4 – Effect of Reset Setting on Set Point Response



to eliminate the remaining error or offset. Whereas insufficient gain slows down the initial approach, inadequate reset action slows down the final approach to set point. If the reset action is too large (reset time is too small), there is excessive overshoot. The addition of reset action reduces controller stability and is dangerous for exothermic reactor temperature loops. Polymerization reactors often use proportional-plus-derivative temperature controllers (i.e., no integral action). For level loops, using a reset time of fewer than 3,000 seconds per repeat makes it necessary to use relatively high gains (5.0 for large volumes) to prevent slow, nearly sustained oscillations.

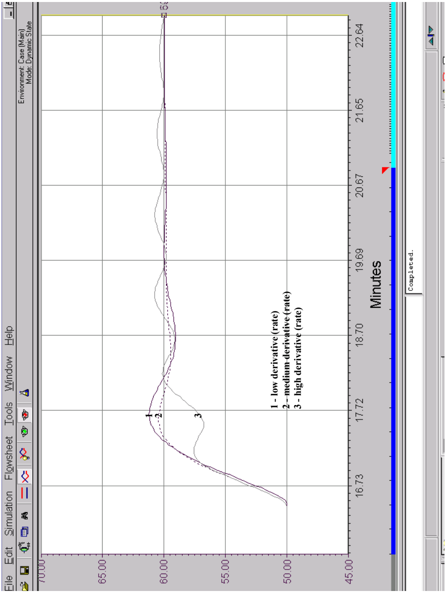
Both reset and gain can cause oscillations, and a high gain setting can aggravate an overshoot caused by reset action. For this reason, checking the ratio of the reset time to the oscillation period can help you distinguish the main culprit. You will need to decrease the reset action (i.e., increase the reset time) of the loop if the ratio is much less than 0.5 for vessel or column temperature, level, and gas pressure control or if it is much less than 0.1 for flow, liquid pressure, pipeline, spin-line, conveyor, or sheet (web) control.

Rule 8 – **If there is excessive overshoot and oscillation, decrease the reset action (increase the reset time).** If the current reset time setting is less than half the oscillation period for loops on mixed volumes or less than a tenth of the period for any loop, there is too much reset action. If the oscillation period changes by 25 percent or more as you change reset action, it is a sign that the oscillation is caused or aggravated by reset action.

Rule 9 – **For level loops with low controller gains (< 5.0), reset action must be greatly restricted (i.e., reset time increased to 50 or more minutes) to help mitigate nearly sustained oscillations.** Level loops are the opposite of most other loops in that you can increase the reset action (decrease the reset time) as you increase the gain (see page 99 for more detail).

Figure 5 shows the effect of rate setting on a set point response. A rate setting that is too small increases any overshoot caused by reset action. A setting that is too large causes the approach to set point to staircase and if large enough it can cause fast oscillations.

Figure 5 – Effect of Rate Setting on Set Point Response



Rate can be used whenever there is a smooth and slow open-loop process response. Rate cannot be used where there is excessive noise, chatter, inverse response, or an abrupt response. It can be used primarily on temperature loops that manipulate heating or cooling where there is a temperature transmitter with a narrow calibration span (large spans cause analog/digital [A/D] chatter). Analyzers that have a cycle time (sample and hold) have signals that are too noisy and abrupt for derivative action. Continuous concentration measurements made by electrodes or inline devices (such as capacitance, conductivity, density, microwave, mass spectrometry, pH, and viscosity for mixed volumes) may allow you to use derivative mode if you add a filter to attenuate the high-frequency measurement noise.

If the measurement is continuous but is used for pipeline control, the control response is too abrupt. This is because the loop dead time is large compared to the largest time constant.

There is noise in every process variable. If the noise is greater than the measurement resolution limit, the loop sees it. If it causes an output change greater than the final element or control valve resolution limit, loop variability will be worse in auto than in manual, particularly if there is any rate action. While it is true that increasing the measurement and valve resolution limit reduces the loop's sensitivity to noise, it also degrades its ability to recover from load upsets. If the measurement doesn't see or correct an upset because the control error is smaller than a resolution limit, the control loop performance deteriorates.

Poor control valve resolution is a problem for any loop. If the valve doesn't move, there is no correction. On the other hand, poor measurement resolution is primarily a problem for PID temperature loops. The large-span ranges found in computer input cards or transmitter calibration can result in a resolution limit of 0.25°F or more. It is impractical to expect control that is tighter than twice the resolution limit or 0.5°F. Also, for a change of 0.25°F per minute, an additional minute of dead time is introduced into the loop. Short scan times also introduce chatter because the actual temperature change is small compared to the noise (i.e., the signal-to-noise ratio is too small). The scan time for most temperature loops is too short. Poor measurement resolution and fast scan times are the main reasons why rate cannot be used in temperature loops. Constrained Multivariable Predictive Controllers (CMPC) are much less sensitive to measurement resolution because the scan time is large and the move in the manipulated variable is based on the complete trajectory and not the current measurement like the PID controller. Poor valve resolution is as much a problem for the CMPC as for the PID, unless you set the minimum size of a move to exceed the resolution limit. The resolution limit is unfortunately a function of direction and position.

Rule 10 – **Rate is primarily used in temperature loops that have narrow span transmitters and slow scan times.** However, it is also used for continuous analytical measurements such as capacitance, conductivity, density, mass spectrometry, microwave, pH, and viscosity for concentration control of mixed volumes. If the temperature control is not intended for a mixed volume, the rate setting is very small (6 seconds) and is mostly used to compensate for the thermowell lag. In every case, you must set the measurement filter large enough to keep output fluctuations within the valve dead band.

Rule 11 – **Most temperature and level loops chatter because the PID scan time is too small.** The true temperature or level change within the scan time must be large compared to the A/D resolution limit (0.05%); otherwise, the signal-to-noise ratio is too small.

1.5 The Key to Happiness

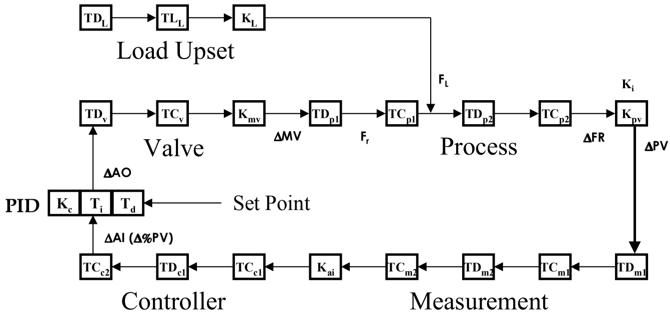
When I was four years old and sitting on my daddy's knee, he said, "Son, I have just one thing to say to you–dead time." Well, it took me forty years to appreciate the significance of his words of wisdom. If a loop has no dead time or noise, perfect control is possible and an infinite gain permissible. There is no tuning issue. Without dead time, I would be out of a job.

Dead time is that period of time from the start of a disturbance until the controller makes a correction that arrives at the same point in the loop at which the disturbance entered. The controller needs to see the upset, react to it, and get the correction to the right place. To appreciate dead time imagine that you go to a party and start drinking. The period of time between the first drink and when you eventually bypass the next round is dead time.

While zero dead time is not possible, a decrease in dead time reduces the effect of nonlinearities, such as changes in the time constants and steady-state gains of the loop, and makes the loop easier to tune. It is desirable that the largest loop time constant be in the process because it will slow down the divergence of the measurement from the set point during an upset and give the controller a chance to catch up. All other time constants are undesirable and create additional dead time. The major time constant can be approximated as follows: it is the time it takes to reach about 63 percent of the final value after the dead time. It takes one dead time and four time constants for a response to reach 98 percent of its final value.

The open-loop steady-state gain is the percentage change in the process variable for a percentage change in output after all transients have died out. The controller is in manual. A high steady-state gain is both a curse and blessing. It can improve the control of the true process variable by making the measurement more sensitive. This is particularly important for inferring composition from temperature or concentration from pH. However, a high steady-state gain can make it more difficult--and sometimes impossible--to control the measured variable. For example, excessive oscillation can result from a small amount of stick and slip in a reflux valve for an acrylonitrile and water distillation or in a reagent valve for sulfuric acid and sodium hydroxide neutralization. A low steady-state gain often results when a throttle position is on the excessively flat portion of the installed valve characteristic. The process variable will often wander and respond to the conditions of a related loop more than from the manipulated variable.

Figure 6 – Loop Block Diagram



- "Open Loop" - loop is in manual (PID algorithm is suspended)
- "Closed Loop" - loop is in auto (PID algorithm is active)

The block diagram in Figure 6 shows all the dead times, time constants, and steady-state gains in a control loop. Feedback control corresponds to the signal making one complete circle around the block diagram. The total dead time is approximately the sum of all the pure dead times and small time constants as you traverse the loop. The dead time from valve dead band is inversely proportional to the rate of change of the controller output, and the dead time from transportation delays is inversely proportional to throughput. The diagram in Figure 6 has many uses--including preventing guests from overstaying their welcome. Just show this slide and start talking about dead time and you will be amazed at how quickly the place empties. There is a safety issue, however, in that guests can get trampled in the rush for the door.

The following list summarizes the three key variables and their relationship to loop performance:

1. *Dead Time or Time Delay (TD)*

   The most important of the three key loop variables

   Delays controller's ability to see and react to upset

Perfect control is possible for zero dead time

Nonlinearities become less important as the dead time decreases

2. *Time Constant (TC)*

Better control is possible for a large *process* time constant

Time constants in series create dead time

Measurement time constants give the illusion of providing better control

3. *Steady-State Gain (K)*

The valve and process steady-state gains are usually nonlinear

High steady-state gain causes overreaction and oscillations

Low steady-state gain causes loss of sensitivity and wandering

1.6 Nothing Ventured, Nothing Gained

There is a lot to be gained from having a better understanding of the overall open-loop gain (K_o). However, if you don't have time to study the relationships revealed by the equations presented in this section, you should skip ahead to Section 2.0 on Tuning Settings and Methods.

The controller gain is inversely proportional to the open-loop gain for all types of loops. For loops that are dead-time dominant (i.e., dead time is much greater than the largest time constant), the controller gain can be simply set as one-fourth of the inverse of the open-loop gain. For these same loops the reset time can be set as one-fourth of the dead time. The result is a smooth stable response similar to what you would get from Lambda tuning, which will be discussed (see Section 2.5) as one of the preferred tuning methods for dead-time dominant loops. For loops with a healthy time constant, the controller gain is also proportional to the ratio of the time constant to the dead time. For these loops, the smooth response that is provided by a large time constant (i.e., residence time) of a mixed volume enables you to use a higher controller gain for tighter concentration, gas pressure, and temperature control.

If the loop time constant is much larger than the dead time (TC / TD >> 1), then the controller gain is proportional to this ratio besides the inverse of the open loop again:

$$K_c \cong 0.25 * TC / (TD * K_o)$$

If the loop is dead-time dominant (TC/TD<< 1), then the controller gain is proportional to just the inverse of the open loop gain:

$$K_c = 0.25 * 1 / K_o$$

The manipulated variable gain is linear for a variable speed drive (VSD). For a control valve it is the slope of the installed characteristic curve and is typically very nonlinear. For operating points on the upper part of the installed characteristic curve of a butterfly valve, the slope can be so flat that the valve gain approaches zero.

Note that for concentration and pressure loops, the process variable should be plotted against a flow ratio (e.g., column temperature versus distillate-to-feed ratio, heat exchanger temperature versus coolant-to-feed ratio, and pH versus reagent-to-feed ratio). The process gain is the slope of this curve and is highly nonlinear. At high flow ratios, the slope can be so flat that the process gain approaches zero. The open-loop gain has a flow ratio gain that is inversely proportional to the feed flow, which is an additional source of nonlinearity for pipeline, desuperheater, static mixer, and exchanger temperature or concentration control. In these applications, an equal-percentage valve characteristic is desirable since the valve gain is proportional to flow and cancels out the flow ratio gain. However, for loops on agitated vessels and columns the controller gain is proportional to the process time constant (residence time), which is inversely proportional to feed flow. Since, as we mentioned previously, the controller gain is also inversely proportional to the open-loop gain, the introduction of an equal-percentage valve characteristic is bad news because the flow ratio gain was cancelled out by the process time constant for the controller gain.

The process gain for flow loops is 1. For pressure loops downstream of a compressor, fan, or pump, the process gain is the slope at the operating point on the characteristic curve of the compressor, fan, or pump. At low flows, the slope can be so flat that the process gain approaches zero.

The analog input gain is the inverse of the calibration span for the controlled variable. It is linear except when a square root extractor is not used on a differential head meter.

Rule 12 – **For dead-time-dominant loops, set the gain equal to about one-fourth of the inverse of the open-loop gain and the reset time (seconds per repeat) equal to about one-fourth of the dead time.** Rate action is not used. This provides a smooth stable response similar to Lambda tuning.

Rule 13 – **Use equal-percentage valve trim for pipeline, desuperheater, static mixer, and heat exchanger control to make the loop more linear.** This assumes that the installed characteristic is close to the inherent equal-percentage trim characteristic.

For concentration and temperature loops, the open-loop gain is a steady-state gain that has a flow ratio gain K_{fr}, which is the inverse of the feed flow:

$$K_o = \frac{\Delta AI}{\Delta AO} = \frac{\Delta MV}{\Delta AO} \bullet \frac{\Delta FR}{\Delta MV} \bullet \frac{\Delta PV}{\Delta FR} \bullet \frac{\Delta AI}{\Delta PV}$$

$$K_o = K_{mv} * K_{fr} * K_{pv} * K_{ai}$$

$$K_{fr} = \frac{1}{F_f}$$

For flow loops, the open-loop gain is a steady-state gain that has a unity process gain and no flow ratio gain (K_{fr} omitted and $K_{pv} = 1$):

$$K_o = K_{mv} * K_{ai}$$

For pressure loops, the open-loop gain is a steady-state gain with a process gain K_{pv}, which is the slope of the pump or compressor curve, and no flow ratio gain (K_{fr} omitted and K_{pv} = curve slope):

$$K_o = K_{mv} * K_{pv} * K_{ai}$$

For level loops, the open-loop gain is a ramp rate and uses K_i instead of K_{pv} to denote an integrator gain. K_i is the inverse of the product of fluid density and the vessel's cross-sectional area. There is no flow ratio gain (K_{fr} omitted):

$$K_o = K_{mv} * K_i * K_{ai}$$

$$K_i = \frac{1}{\text{density} \bullet \text{area}}$$

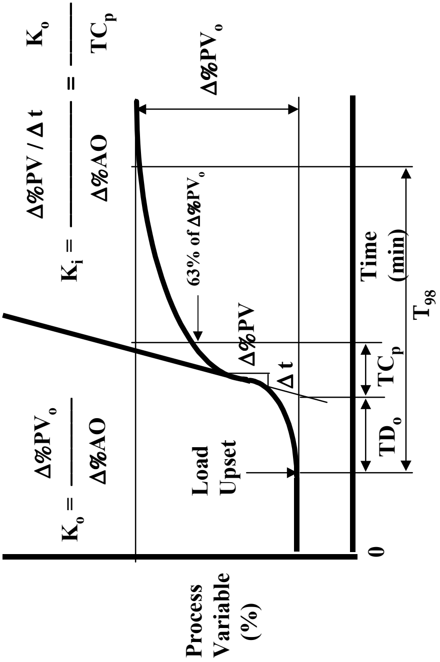
Where:

F_f	=	feed flow (pph)
K_i	=	level loop integrator gain (ft/lb)
K_o	=	overall open-loop gain (one per hour for level, otherwise dimensionless)
K_{mv}	=	manipulated variable (valve or VSD) steady-state gain (pph/%)
K_{fr}	=	flow ratio steady-state gain (inverse of feed flow) (1/pph)
K_{pv}	=	process variable steady-state gain (wtfrac/%) (degC/%) (psi/%)

K_{ai} = analog input steady-state gain (%/wtfrac) (%/degC)(%/psi) (%/pph)(%/ft)
ΔAO = change in the controller's analog output (%)
ΔAI = change in the controller's analog input (%)
ΔPV = change in the process variable (wtfrac) (degC) (psi) (pph) (ft)
ΔFR = change in the flow ratio (pph/pph)
ΔMV = manipulated variable or valve flow (pph)

Figure 7 shows the dead time, time constant, and steady-state gain for a process that reaches a steady state (i.e., self-regulating) and for an integrator such as level. In both cases, the process variable response is to change the controller output with the loop in manual (open-loop response). Figure 7 also shows the equations used to approximate an integrator gain as equal to the initial ramp rate of a self-regulating response. This initial ramp rate is approximately the steady-state gain divided by the dominant time constant. Auto tuners that compute settings from an open-loop response use this approximation to convert an integrator into an equivalent

Figure 7 – Integrator Gain (Ramp Rate)



time constant and steady-state gain because they cannot handle integrator gain. This method works well when the dead time is small compared to the ramp rate. The dead time is the time required for the PV to get out of the noise band.

2.0—Tuning Settings and Methods

2.1 First Ask the Operator

You should test new tuning settings by changing the set point in both directions and observing whether the response is smooth and fast enough for these set point changes and load upsets. Don't leave these new settings over night until you are sure they are right. Before entering new settings or trying tuning methods, ask the operator for permission to tune the loop. Also ask him or her what the maximum allowable size can be for each direction of a step change in the controller set point and output without causing a problem. You should put most analog controllers and some digital controllers in manual before entering a change in gain to avoid bumping the controller output.

2.2 Default and Typical Settings

Tables 2 and 3 show the default and typical tuning settings for controllers whose reset setting is in repeats per minute and seconds per repeat, respectively. The number in front of the parentheses in each column is a default setting, a number that can be used for a download or for a guess if nothing is known about the loop and no test can be performed. The numbers within the parentheses represent a range of typical values. If your tuning setting falls outside the suggested range, this doesn't necessarily mean the tuning setting is wrong. It does suggest that the loop is unusual or has some problems such as interaction, noise, a coated sensor, or a sticky control valve. The last column shows the tuning methods that generally give the best results. The "general-purpose" tuning method logic shown in Figure 8 is the closed-loop method (CLM), which is referenced in Tables 2 and 3. While this method is best for gas pressure, reactor, and level loops, it can be used for most other loops as well, particularly if the loop's set point is being driven for advanced, batch, cascade, or sequence control and there are no interaction issues.

Table 2 – Default and Typical PID Settings (scan in sec, reset in rep/min, and rate in minutes; λ = Lambda, CLM = Closed-loop method; SCM = Shortcut method)

Application Type	Scan (seconds)	Gain	Reset (repeats)	Rate (minutes)	Method
Liquid Flow/Press	1 (0.2-2)	0.3 (0.2-0.8)	10 (5-50)	0.0 (0.0-0.02)	λ
Tight Liquid Level	5 (1.0-30)	5.0 (0.5-25)*	0.1 (0.0-0.5)	0.0 (0.0-1.0)	CLM
Gas Pressure (psig)	0.2 (0.02-1)	5.0 (0.5-20)	0.2 (0.1-1.0)	0.05 (0.0-0.5)	CLM
Reactor pH	2 (1.0-5)	1.0 (0.001-50)	0.5 (0.1-1.0)	0.5 (0.1- 2.0)	SCM
Neutralizer pH	2 (1.0-5)	0.1 (0.001-10)	0.2 (0.1-1.0)	1.2 (0.1-2.0)	SCM
Inline pH	1 (0.2-2)	0.2 (0.1-0.3)	2 (1-4)	0.0 (0.0-0.05)	λ
Reactor Temperature	5 (2.0-15)	5.0 (1.0-15)	0.2 (0.05-0.5)	1.2 (0.5-5.0)	CLM
Inline Temperature	2 (1.0-5)	0.5 (0.2-2.0)	1.0 (0.5-5.0)	0.2 (0.2-1.0)	λ
Column Temperature	10 (2.0-30)	0.5 (0.1-10)	0.2 (0.05-0.5)	1.2 (0.5-10)	SCM

* An error/square algorithm or gain scheduling should be used for level loops with gains < 5

Table 3 – Default and Typical PID Settings (scan in sec, reset in sec/rep, and rate in sec; λ = Lambda, CLM = Closed-loop method, SCM = Shortcut method)

Application Type	Scan (seconds)	Gain	Reset (seconds)	Rate (seconds)	Method
Liquid Flow/Press	1 (0.2-2)	0.3 (0.2-0.8)	6 (1-12)	0 (0-2)	λ
Tight Liquid Level	5 (1.0-30)	5.0 (0.5-25)*	600 (120-6000)	0 (0-60)	CLM
Gas Pressure (psig)	0.2 (0.02-1)	5.0 (0.5-20)	300 (60-600)	3 (0-30)	CLM
Reactor pH	2 (1.0-5)	1.0 (0.001-50)	120 (60-600)	30 (6-30)	SCM
Neutralizer pH	2 (1.0-5)	0.1 (0.001-10)	300 (60-600)	70 (6-120)	SCM
Inline pH	1 (0.2-2)	0.2 (0.1-0.3)	30 (15-60)	0 (0-3)	λ
Reactor Temperature	5 (2.0-15)	5.0 (1.0-15)	300 (300-3000)	70 (30-300)	CLM
Inline Temperature	2 (1.0-5)	0.5 (0.2-2.0)	60 (12-120)	12 (12-60)	λ
Column Temperature	10 (2.0-30)	0.5 (0.1-10)	300 (300-3000)	70 (30-600)	SCM

* An error/square algorithm or gain scheduling should be used for level loops with gains < 5

2.3 The General-purpose Closed-loop Tuning Method

A closed-loop (controller in auto) method has the following advantages over open-loop (controller in manual) methods:

1. It forces the user to find the maximum controller gain to minimize peak error and dead time from dead band. It also forces the user to find out whether reset and rate is even needed, and it ensures rapid set point response and gets inside the window of allowable controller gains for integrating and runaway loops.

2. Loops stay in automatic, which is safer for difficult or very fast loops.

3. It includes the effects of valve hysteresis and dead band.

4. It includes the dynamics and peculiar features of controller algorithms.

5. It includes nonlinearities that are dependent on direction and rate of change.

6. It facilitates the tuning of the master (outer) loop of a cascade system for an oscillating inner (slave) loop.

This closed-loop procedure uses rapidly decaying oscillations as shown in Figure 9 instead of the sustained oscillations proposed by the original Ziegler-Nichols method. The measured period will be larger for damped oscillations in industrial applications mostly because of valve dead band, but the error will be in the safe direction in terms of a larger, more stable reset time setting. The closed-loop procedure should also result in about half of the gain estimated by the Ziegler-Nichols method.

Rule 14 – **Use the general-purpose closed-loop method if the loop must stay in auto or if it is particularly important to maximize the gain for tight control and a fast set point response.** However, for dead-time-dominant loops, you should substantially decrease the factor for the reset time. This will prevent the set point response from faltering because of too much gain action and not enough reset action. For temperature and pressure loops on exothermic reactors, it is especially important to use this method to prevent a runaway.

Figure 8 – General-Purpose Closed-loop Method

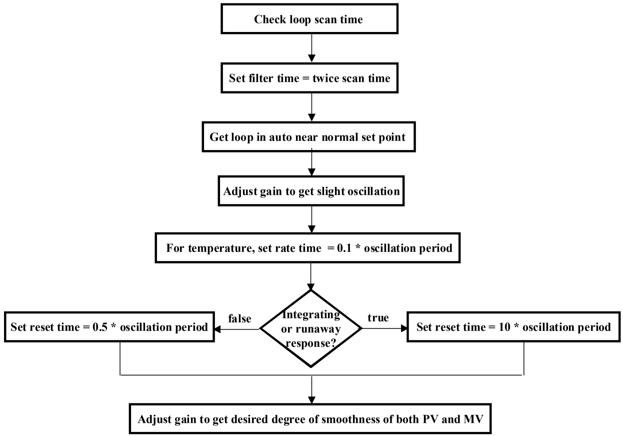
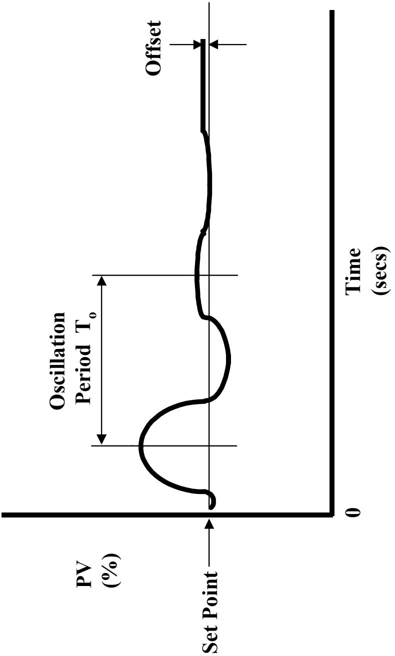


Figure 9 – Slight Oscillation



A list of steps for the "Closed Loop Method" is as follows:

1. Put the controller in automatic at normal set point. If it is important not to make big changes in the manipulated variable (analog output), narrow the controller output limits to restrict valve movement.

2. For level, gas pressure, and reactor loops decrease the reset action (i.e., increase the reset time) by a factor of ten if possible and trend-record the process variable (PV) and analog output (AO).

3. Add a PV filter to keep output fluctuations within the dead band of valve caused by noise.

4. Bump the set point and increase the controller gain if necessary to get a slight oscillation.

5. Stop increasing the gain when the loop *starts* to oscillate or the gain has reached your comfort limit. Then note the period. For gain settings greater than 1, the oscillation will be more recognizable in the analog output (AO). Make sure AO stays on scale within the valve's good throttle range.

6. Reduce the gain until the oscillation just disappears so recovery is smooth.

7. For a temperature loop with a smooth response (no chatter, inverse, square wave, or interaction), use rate action. If the gain is larger than 10, reset and rate action are not needed. If the manipulated flow will upset other loops, decrease the gain or use error-squared control. If a high gain is used, set a velocity limit for the set point and configure the set point so it tracks the PV in manual and DDC (ROUT). This will enable the loop to restart.

8. If rate action is used, set the rate time equal to one-tenth (1/10) of the period and set reset time equal to half the period. If rate action is not used, cut the gain by 50 percent. If the loop is clearly dead-time dominant, increase the reset action so the reset time is about one-eighth (1/8) of the loop period. Make another set point change and adjust the gain to get a smooth response. Do not become any more aggressive than a slight oscillation.

9. If you select gains smaller than 5 for level, decrease the reset action (i.e., increase the reset to more than 3,000 seconds), add feed-

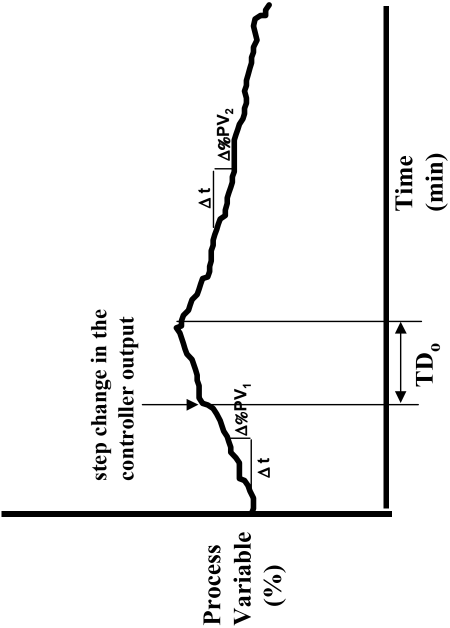
forward, and add rate if there is no inverse response or interaction.

2.4 The Shortcut Open-loop Method

The shortcut method is ideally suited for very slow responses such as column temperature where there is a good control valve and positioner and you need to get a quick estimate of the controller tuning settings. This method looks at the change in ramp rate of the %PV as shown in Figure 10 for about two to three dead times. It doesn't require the loop to be at steady state. However, if there is an upset that causes the ramp rate to change, the results will be inaccurate. In general, you should repeat this test in both directions and use the most conservative settings. Also, if the bump in controller output is much larger than the dead band, the shortcut method doesn't include the dead time from valve dead band. If the changes in controller output per scan approach the control valve dead band in size, you should add the additional dead time from the valve dead band to the observed dead time.

The shortcut method is also effective for pH loops because it can keep the test near the operating point on the titration curve. The use of a closed-loop method can get confusing for pH

Figure 10 – Change in Ramp Rate for Non-Steady State



particularly if the oscillations develop into a limit cycle after being bounced back and forth between the flat ends of the titration curve. The period of such a limit cycle is extremely long and variable, and it will occur for a large range of controller gains.

A list of steps for the "Shortcut Method" are as follows:

1. Adjust the measurement filter to keep the controller output fluctuations caused by noise within the valve dead band.

2. Note the magnitude of output change for each reaction to typical upsets. With the controller in manual near set point, make a step change in the controller analog output (Δ%AO) of about the same magnitude as the output change you noted, but larger than twice the valve dead band.

3. Note the observed dead time and the change in ramp rates. If the process was lined out before the test, then the starting ramp rate is zero ($\Delta\%PV_1 / \Delta t = 0$).

4. Divide the change in ramp rate by the change in valve position to get the pseudo

integrator gain (K_i). Then compute the dead time from the dead band.

5. Use the following equations. For a master or supervisory loop, omit TD_v.

$$K_i = \frac{\left|(\Delta\%PV_2/\Delta t) - (\Delta\%PV_1/\Delta t)\right|}{|\Delta\%AO|}$$

$$K_c = \frac{K_x}{K_i \bullet TD_o}$$

$$TD_v = \frac{DB}{K_x \bullet \Delta\%AVP} \bullet TD_o$$

$$\Delta\%AVP = |\Delta\%AO| - \frac{DB}{2}$$

$$T_i = c_i \bullet (TD_v + TD_0)$$

$$T_d = c_d \bullet (TD_v + TD_0)$$

Where:

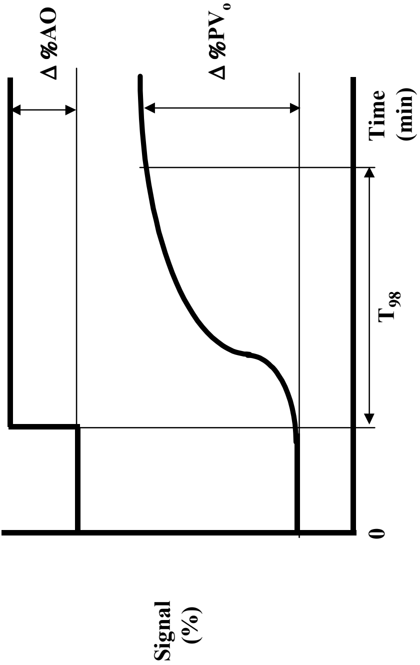
$\Delta\%AVP$ = change in actual valve position (%)
$\Delta\%AO$ = change in the controller analog output (%)
DB = dead band from valve hysteresis (%)
K_c = controller gain (dimensionless)
c_d = rate time coefficient (1.0 for back-mixed and 0.0 for plug flow volumes)
c_i = reset time coefficient (4.0 for back-mixed and 0.5 for plug flow volumes)
K_x = gain factor (1.0 for Ziegler-Nichols, 0.5 for IMC, and 0.25 for Lambda)
K_o = open-loop gain (dimensionless)
K_i = pseudo integrator open-loop gain (1/sec)
$\Delta\%PV$ = change in process variable (%)
Δt = change in time (sec)
TC = largest loop time constant (sec)
TD_o = dead time seen in open-loop test (sec)
TD_v = dead time from control valve dead band (sec)
T_d = derivative (rate) time setting (seconds)
T_i = integral (reset) time setting (seconds/repeat)

Rule 15 – **Use the shortcut method when you want a quick estimate for a very slow or nonlinear loop, provided that the valve dead band is 0.25 percent or less and the step size keeps you near the operating point.** Make sure there are no load upsets during the test and that you measure the new rate of change of the PV for at least two dead times. You should repeat the test for both directions and use the most conservative tuning.

2.5 Simplified Lambda Tuning

If the loop is dead-time dominant, Lambda tuning is the best method. It also helps to minimize interactions and suppress oscillations. Lambda tuning is an open-loop method that is particularly effective for relatively fast loops, such as in pipelines, desuperheaters, static mixers, exchangers, conveyors, spin lines, and sheet (web) lines, or wherever there is plug flow. It provides a closed-loop time constant that approximates the open-loop time constant. Figure 11 shows the step change in controller output and the open-loop response of the process variable. The user simply needs to note these changes and the time required to reach 98 percent of the final response (T_{98}). This simplified form of the equation for Lambda

Figure 11 – Lambda Test



tuning is only suitable for self-regulating processes. EnTech has expanded the actual Lambda tuning rules to cover a wide variety of processes.

The following is a list of steps for the “Simplified Lambda Tuning Method”:

1. Adjust the measurement filter to keep the controller output fluctuations caused by noise within the valve dead band.

2. Note the magnitude of output change to determine the reaction to typical upsets. With the controller in manual near set point, make a step change in the controller analog output (Δ%AO) of about the same magnitude as the output change you noted, but larger than twice the valve dead band.

3. Note the observed dead time as the time it took to reach the first change in the process variable (Δ%PV) outside of the noise band.

4. Note the open-loop gain (the percentage change in measurement divided by the percentage change in controller output).

5. Note the response time (the time from the bump to 98 percent of the final value).

6. The reset setting (repeats/minute) is set equal to the inverse of one-fourth of the response time.

7. The controller gain is adjusted to be one-fourth of the inverse of the open-loop gain ($\lambda = 4$).

The equation for reset time setting is as follows:

$$T_i = T_{98} / 4 = TD_o / 4 + TC \text{ (seconds/repeat)}$$

The equation for gain setting is as follows:

$$K_c = (\Delta \%AO / \Delta \%PV) * (1/ \lambda)$$

(λ is a tuning factor that is increased to provide a slower and smoother response.)

Rule 16 – **Use the Lambda tuning method for dead-time-dominant or fast loops that can be left in manual.** You should repeat the test for both directions and use the most conservative tuning.

Rule 17 – **In Lambda tuning, decrease λ to speed up the controller's response to upsets and set point changes (minimize Δ PV) and increase λ to slow down the response and reduce loop interaction (minimize Δ %AO).** This simplifies tuning to a single knob.

3.0–Measurements and Valves

3.1 Watch Out for Bad Actors

The analog/digital converter (A/D) chatter from large temperature measurement spans and short scan times is the factor most frequently cited to explain why derivative action cannot be used in PID controllers. It is also a considerable source of dead time because the measurement must get out of the noise band if the PID controller is to discern a true load change from the chatter. There is a similar signal-to-noise ratio problem for level measurements, particularly if the signal is noisy due to sloshing or bubbles.

The next most frequent problem is sensor coating and drift. Resistance temperature detectors (RTDs) and smart transmitters can reduce drift by an order of magnitude. Keeping the velocity that is passing a sensor above 5 fps is the most effective way to keep a sensor clean.

Rule 18 – **Narrow span ranges and slow scan times should be used for temperature and level measurements to minimize the**

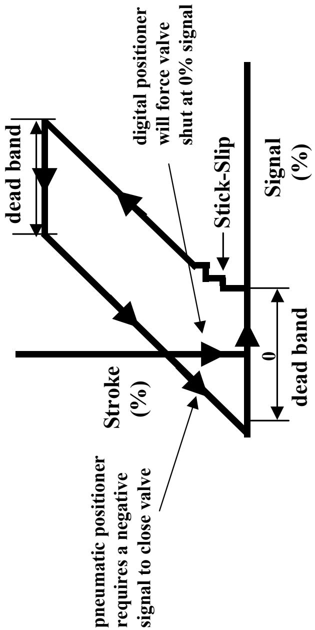
dead time and chatter from the analog/digital converter (A/D). For level measurements, you should minimize sensitivity to bubbles, sloshing, and coating.

Rule 19 – **Use RTDs and smart transmitters to reduce drift and calibration requirements.** They will more than pay for themselves through lower maintenance costs and more accurate process operating points.

3.2 Deadly Dead Band

According to EnTech studies, most loop variability is caused by poor tuning and control valve resolution. Figure 12 shows the dead band that occurs whenever a control valve changes direction and the staircasing that is the result of stick-slip action. The dead band and stick slip are usually largest near the valve seat. Valve resolution is the minimum change in signal in the same direction that will result in a flow change and is usually about half of the valve dead band. As you approach the resolution limit there is a dramatic increase in the loop dead time from dead band, as illustrated in Figure 13. This additional dead time can be about five times larger for pneumatic positioners than for digital positioners. Figure 14 shows that it is

Figure 12 – Dead Time from the Valve



worse for piston actuators and rotary valves. Most valve manufacturers will choose a valve position and step size that is near the minimum shown in Figures 13 and 14. For large step sizes, the time required to move large amounts of air into or out of the actuator increases the response time.

Another problem with rotary valves is that the feedback to the positioner is from the actuator shaft. What might seem to be minor gaps in key-lock connections and twisting in long shafts actually means that the butterfly disk or ball may not move, even though the positioner sees a movement of the actuator shaft. To minimize this problem, you should use splined connections and short large-diameter shafts, and you should conduct actual flow tests on all rotary valves. All tests on control valves should use step sizes that approximate what is expected to occur as changes in the controller output from one scan to another ($< 0.5\%$). You should also use very large step sizes for pressure-relief valves and for compressor antisurge valves. There is nothing in a valve specification that requires that the valve will actually move. To ensure that a valve will respond to meet the needs of a control loop, you should add the dynamic classes in Table 4 to the valve specification sheet when purchasing a control valve.

Figure 13 – Dramatic Rise in Response Time as You Approach the Resolution Limit

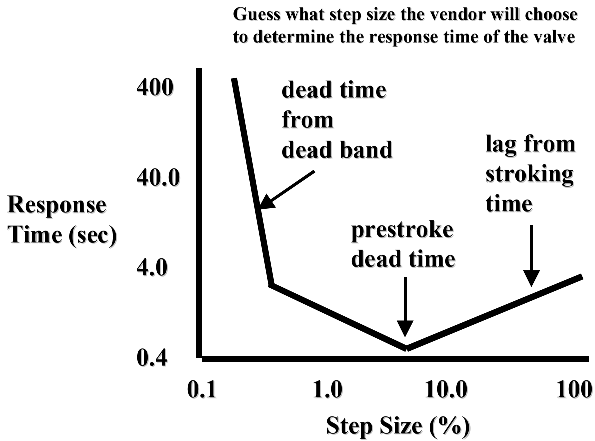
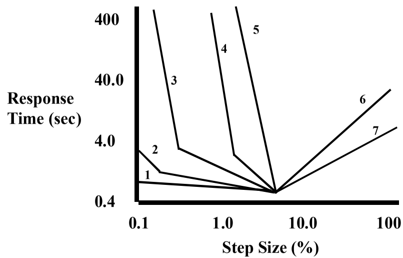


Figure 14 – Response Time Depends upon the Valve, Shaft, Size and Connections, Actuator, and Positioner



1 - sliding stem valve with diaphragm actuator and digital postioner with pulse width modulated solenoids.
2 - sliding stem valve with diaphragm actuator and digital positioner with nozzle flapper
3 - sliding stem valve with diaphragm actuator and pneumatic positioner
4 - rotary valve with piston actuator and digital positioner
5 - rotary valve (tight shutoff) with piston actuator and pneumatic positioner
6 - very large rotary valve (>6”) with any type of positioner
7 - sliding stem valve with a digital positioner

Table 4 – Dynamic Classes of Control Valves (Four classes A → D for each of the four categories 1 → 4)

1-Minimum Step Classes Class A - 3.0% ± 0.3% Class B - 1.0% ± 0.1% Class C - 0.5% ± 0.1% Class D - 0.2% ± 0.1%	**2-Maximum Step Classes** Class A - 5% ± 0.5% Class B - 10% ± 1.0% Class C - 20% ± 2.0% Class D - 50% ± 2.0%
3-Response Time Classes Class A - 15 sec Class B - 5 sec Class C - 2 sec Class D - 1 sec	**4-Minimum Position Classes** Class A - 30% Class B - 20% Class C - 10% Class D - 0%
Note: The response time is the time it takes the trim (not the actuator) to stay within 10 percent of step or within 0.1 percent of span, whichever is largest. (Overshoot is OK if recovery is within this offset from the desired position of the trim)	

Referring to Table 4, consider a DACB class control valve. It will respond to signals larger than 0.2 percent and smaller than 5.0 percent in less than 2 seconds above a position of 20 percent. If you don't care about how well a valve responds, specify class AAAA--almost any valve will meet it.

A general-purpose application would be BAAB. Examples of other classes for various types of control loops are given in the following list. To choose a typical class, use the loop type that actually throttles the control valve. For cascade loops, you should use the slave loop.

- Vessel or Column Temperature Control => CABC
- Exchanger Temperature Control => CACC
- Pipeline Temperature Control => CCCC
- Liquid Pressure or Flow Control => CACC
- Vessel Level Control => CABC
- Gas or Steam Pressure Control => CBDC
- Compressor Surge Control => CCCD
- Vessel pH Control => DABD
- Pipeline pH Control => DBCD
- Pressure Relief => BCCD
- For split range, use class D for last category (minimum throttle position)

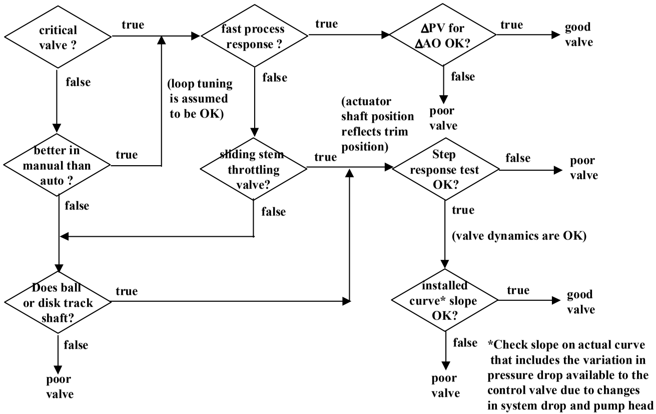
The best way to see if the flow actually changed for a small change in controller output is to make a sensitive flow measurement with very low noise. This is particularly important for rotary valves or pneumatic positioners. Another clue that the valve has a problem is if the control loop oscillations are smaller in manual than in auto (this assumes you have tuned the controller). If it is a critical rotary valve and there is no flow measurement, you need to take the valve out of the pipeline. In the shop, you need to then attach a travel indicator to the butterfly disk or ball to check whether they track the changes in the actuator shaft.

Finally, the change in the flow measurement for a change in controller output will also show if the control valve is on a portion of the installed characteristic that is too steep or flat. If there is no flow measurement, ask the valve manufacturer to generate an installed valve characteristic for your piping system. Signal characterization to linearize the valve will decrease valve dead time from its resolution limit for the flat portion of the curve, but it will increase this dead time for the steep portion of the curve.

Figure 15 summarizes the maintenance tests for a good or bad valve. The change in PV can be any process variable, but it works best if it is a fast measurement such as flow or pressure. Slow process variables such as temperature and level respond much too slowly, especially for small changes in controller output, to make it practical to determine valve dead band.

Rule 20 – **To achieve the tightest control (i.e., least variability), use sliding stem valves and digital positioners.** If you use rotary valves, you should have a flow

Figure 15 – Maintenance Tests for a Good or Poor Control Valve



measurement, splined shaft connections, and a short large-diameter shaft.

Rule 21 – **If you must operate on the flat portion of an installed valve characteristic, use signal characterization to make the valve more sensitive.** Be careful to avoid increasing the valve dead time from dead band on the steep portion of the curve. Also make sure you don't eliminate an equal-percentage characteristic used to compensate for a flow ratio gain.

4.0–Control Considerations

4.1 Auto Tuners

The relay method auto tuner shown in Figure 16 is effective and simple. It keeps the loop under control by constantly forcing the PV back and forth across the starting point of the test. Its effectiveness depends upon the noise band being set correctly so that fluctuations from noise are not measured as loop oscillations. Figure 17 shows a single step in controller output that uses the shortcut tuning method to provide an estimate of tuning settings and then a series of successive steps to provide an automatic identification of valve dead band.

4.2 Uncommonly Good Practices for Common Loops

For flow and pressure loops, first make sure the scan time is fast enough. For gas pressure, the gain must be maximized. For level loops it is most important that you decide whether the level control must be tight or loose and that you minimize the reset action. For temperature loops it is critical that you use a narrow calibration span and slow scan time. For concentration loops you must make sure that the sensor is not

Figure 16 – Relay Method Auto Tuner

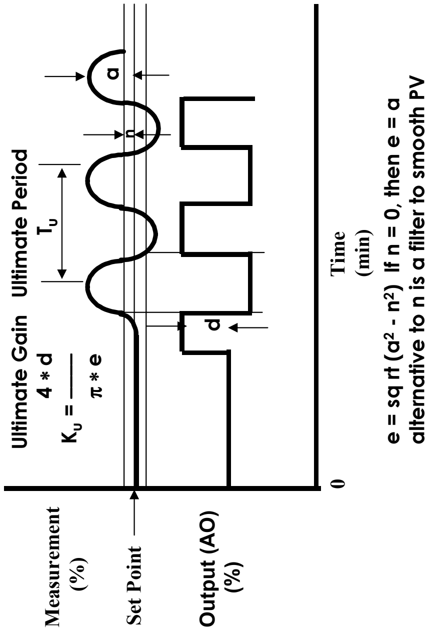
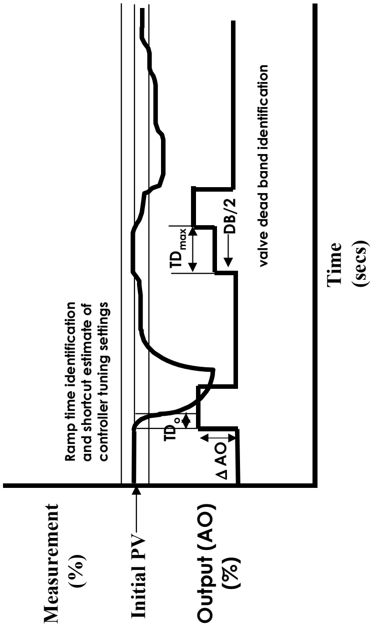


Figure 17 – Enhanced Pretest and Pretune



coated (velocity $>$ 5 fps) and that the signal is fast (minimum transportation delay) and smooth (no noise). For pH loops with an operating point on a steep titration curve, the greatest need is for an exceptionally precise control valve (dead band $<$ 0.15%). Characterizing the signal of the process variable according to the titration curve can help reduce the oscillations.

4.3 Dead-Time Compensation and Warp Drive

When I first left home, my dad said, "Be as honest as the day is long, don't talk when you should listen, and don't be fooled into thinking a dead-time compensator can eliminate process dead time." A dead-time compensator such as a Smith Predictor can cancel out the effect of dead time for changes in the controller output and make possible higher gains and faster reset action (smaller reset times). However, it is a common misconception that it eliminates dead time when correcting disturbances from feedback action. The minimum peak error still corresponds to the excursion of the process variable in one total loop dead time. The disturbance and correction must still make a complete traversal of the block diagram. Dead time in the plant cannot be eliminated by an algorithm without violating the laws of physics. Unless you have Scotty and warp drive, you are stuck with the dead time caused by equipment, piping, instrumentation, and control valves in your loop.

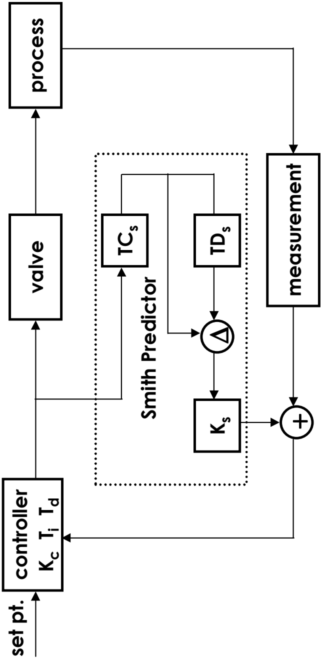
A dead-time compensator is very sensitive to an overestimate of the dead time. A dead time that is 25 percent larger than actual can cause instability. A major source of dead time is a transportation delay, which can be computed for a pipeline as the volume divided by the flow or for a sheet as the distance divided by the speed. Overestimates of the process gain and underestimates of the time constant are also problems but to a lesser degree.

To derive the full advantage of a Smith Predictor, you should increase the controller gain and the reset action. For a negligible time constant, the reset time can be set equal to one-fourth of the uncorrected dead time to provide a reset action that is an order of magnitude greater. A dead-time compensator is most effective on dead-time-dominant loops where the dead time can be accurately calculated and updated in the predictor.

Most dead-time compensators can be reduced to some form of the Smith Predictor shown in Figure 18. The PID controller output passes through a dead time, single time constant, and steady-state gain model of the valve, process, and sensor dynamics. The model output with and without the loop dead time is subtracted from and added to the measurement of the process variable, respectively. This leaves the model output without the dead time as the controlled variable. The dead time has been removed from the loop as far as changes in the controller output are concerned. Note that the controlled variable is no longer the actual process variable.

Rule 22 – **To maximize the stability of a dead-time compensator, update the dead time as quickly and accurately as possible and make sure you always underestimate the dead time and process gain and overestimate the time constant.** You should include the change in dead time resulting from a change in a transport delay by using a computation in which the dead time is inversely proportional to flow or speed.

Figure 18 – Smith Predictor

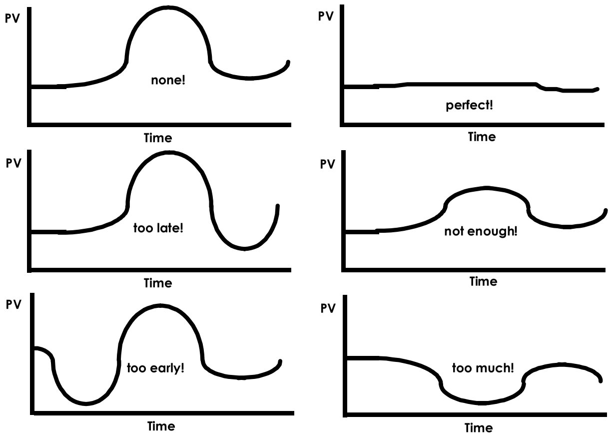


Rule 23 – **The reset time should be set to be about one-fourth (¼) of the uncompensated dead time to get the most out of a dead-time compensator.** The reset time can be decreased by an order of magnitude by adding a dead-time compensator that has just a 10 percent underestimate of the dead time.

4.4 I Have So Much Feedforward, I Eat before I Am Hungry

A feedforward signal that is accurate both in gain and in timing can make an impressive impact on control loop performance, especially for unit operations with large dead times such as distillation columns and large reactors. Ideally, the compensating effect from the feedforward signal should arrive in the process at the same time as the load change and be equal but opposite to the upset. For fast loops, the timing is tight, and the feedforward signal may arrive early and cause an inverse response. If it is too late, the feedforward signal creates a second disturbance. For fast loops, the timing is more critical than the gain, but dynamic compensation is often neglected. Figure 19 shows the patterns of response that are a clue to how you should adjust the feedforward gain and timing.

Figure 19 – Feedforward Tuning



You can greatly improve a loop's set point response by adding the percentage change in set point into the controller output as a feedforward signal that has a gain of about half of the controller gain, an action opposite to the control action, and a slight filter. This provides the kick you would normally get from a high gain setting. The loop can then rely on reset action to make the rest of the transition to the new set point.

Rule 24 – **Both the feedforward gain and the dynamic compensation (dead time and lead lag) must be set properly to get the most benefit.** For fast processes the timing is more critical. It is better for the signal to arrive a little late than too early because inverse response is extremely disruptive.

Rule 25 – **Add a feedforward signal of a set point in percent to the controller output with a gain about half of the controller gain, an action opposite of the controller action, and a small filter.** The feedforward action is opposite the controller action because the controller action works on error, which is the measurement minus the set point. For nonlinear valves, you may need to add signal characterization.

4.5 Cascade Control Tuning

Cascade control is a type of control in which a secondary (slave or inner) loop is added that gets a set point from a primary (master or outer) loop. If the secondary loop response (both dead time and time constant) is five times faster than the primary loop response, there is no interaction between the loops, and the secondary loop can correct for upsets it can measure before they affect the primary loop. If the secondary loop is not fast enough, you must increase the scan time and the PV filter time of the primary loop or you will have to decrease the controller gain and rate action of the primary loop to reduce the interaction. Besides catching an upset quicker, the secondary loop may also help linearize the response of the primary loop. For example, in the cascade of reactor temperature to coolant jacket exit temperature, the secondary loop of jacket temperature will make the process gain for reactor temperature linear in addition to sensing coolant temperature disturbances before they affect the reactor temperature. Similarly, the cascade of exit coolant temperature to a flow loop on coolant makeup will remove the nonlinearity of the installed characteristic and dead band of the valve in addition to reducing the effect of coolant pressure changes. The most common type of secondary loop is the flow loop. It can considerably improve the performance of a primary loop for concentration, level, and temperature control. The flow loop is not generally recommended for liquid pressure control because the speed of response of the liquid pressure and flow are about the same. Another type of cascade control that should be used more frequently is the cascade control of still, reactor, or evaporator temperature to steam pressure control. In this kind of control, the steam pressure loop compensates not only for steam supply pressure upsets but also for changes in the condensing rates (heat load and transfer) as reflected in the steam coil or jacket pressure.

You should tune the secondary loop for an immediate response before tuning the primary loop. An offset in the secondary loop is usually of no consequence since the sole purpose of the secondary loop is to meet the demands of a primary loop. You should tune the secondary loop with maximum gain action and minimum reset action and use a feedforward of the set point. This is particularly important for cascade control of gas pressure control to flow. If the abrupt changes in the secondary loop output upset another important loop, then you may need to use Lambda tuning to make the secondary loop response smoother and more gradual.

The most common mistake is to forget to properly set the output limits of the primary loop. You must set the output limits on the primary loop so they match the set point limits on the secondary loop. Also, the primary loop must not wind up when the secondary loop output is at its output limits.

Rule 26 – **If the dead time and time constant of the primary loop are not five times slower than they are for the secondary loop, the primary loop must be slowed down.** The scan time and filter time must be increased or the controller gain and rate action must be decreased in the primary loop.

Rule 27 – **You must tune the secondary loop first and use gain and/or set point feedforward action in it so the secondary loop can immediately respond to the set point changes from the primary loop.** Reset action is too slow to be the major source of control action in the secondary loop. If you use a feedforward of the set point, make sure it is scaled properly. If the secondary loop fights with other important loops, use Lambda tuning to minimize the interaction.

Rule 28 – **Make sure the output limits of the primary loop match the set point limits of the secondary loop.** In a fieldbus-based system, the primary loop output limits use the engineering units of the secondary loop. In older systems, the primary loop output limits are usually expressed as a percentage of the secondary loop scale.

5.0—Troubleshooting

5.1 Patience, Heck, I Need to Solve the Problem

Figures 20 and 21 show some diagnostics for loops in manual and auto. In these diagrams, reset action (repeats per minute) is referenced rather than reset time (seconds per repeat). If there are fewer oscillations with the loop in manual, the problem is either a poor control valve or controller tuning. If there are also fewer oscillations in other loops, then they are probably caused by an interaction between this loop and other loops. If the oscillations persist and are fast, they are probably due to electromagnetic interference (EMI), sensor noise, pressure waves, or resonance. If the oscillations persist and are slow, then they are periodic upsets from other loops that have poor tuning or valves or are caused by on-off actions (level switches), steam traps, pressure regulators, burps (column flooding), or flashing. If the oscillations dissipate when the valve is closed, the oscillations were caused by pressure fluctuations at the valve.

With a loop in manual, you can find the valve dead band by making a step change of 0.25 percent in the valve and then waiting more than the dead time to see if the actual process variable responds to the change in the controller output. For slow processes or in situations where you are reasonably sure the valve is really lousy, you can increase the step size to 0.5 percent or even 1.0 percent. This will allow the test to be completed before your patience wears thin. If there is no response, repeat the step. The steps should all be in the same direction until there is response of the process variable that is outside of the noise band. Then repeat test for the opposite direction. The absolute magnitude of the total number of step changes needed to get a response in both directions is a measure of the valve dead band.

Rule 29 – **If the oscillation goes away when you put the loop in manual, then the loop is the cause of the oscillation.** The culprit could be the control valve, tuning, or loop interaction.

Rule 30 – **If the oscillation only goes away when you close the valve, then it was caused by pressure fluctuations at the valve.** The culprit could be an oscillating pres-

sure loop, on-off actions, or oscillating users on the same header.

Rule 31 – **To track down the source of an oscillation, put each loop in manual and stop each on-off action one at a time.** When the oscillation stops, you know the culprit was the last loop you put in manual or the last on-off action you stopped.

If the loop is in automatic and there are fast periodic upsets, the oscillations are probably caused by EMI, sensor noise, pressure waves, or resonance. If you experience slow periodic upsets and a period much greater than four times the dead time, then you should suspect on-off actions, steam traps, pressure regulators, burping, or flashing as the source. Otherwise, tuning is probably the culprit. When PV wanders in automatic it is a symptom of the control loop operating on the flat portion of an installed valve characteristic. If the PV falters for a dead-time-dominant loop (see Figure 3b), it is an indication of too much gain action. If the PV staircases (see Figure 5), it is an indication of too much rate action. If the PV overshoots and develops a slow oscillation whose period varies with the reset setting (see Figure 4), it is a sign of too much reset action (i.e., the reset time too small). If the loop period is very sensitive to reset action, it is a sign that there is excessive valve dead band. For sticky valves, you can make the oscillation significantly faster by using more rate and gain action. This will get the controller output through the dead band more quickly.

If the period of an upset is near the natural period of the loop, which is two to four times the dead time, then the loop will amplify the disturbance, and the oscillation will be worse when the loop is in automatic. If the period is much less than the natural period it is uncontrollable noise. For a loop to have a chance of attenuating the upset, the period of the upset must be more than twice as large as the natural period of the loop.

Note that on Figures 20 and 21 "reset too hi" means the reset action is too high, which means the reset time is too small.

Rule 32 – **If the oscillation period for the loop in automatic drastically increases for less aggressive controller tuning, the valve has excessive dead band.** It takes longer for the controller output to work through the dead band if the PID action is slower due to less gain, reset, and rate action.

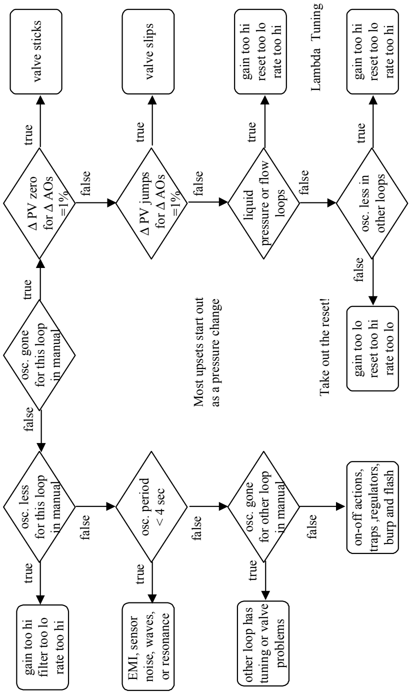
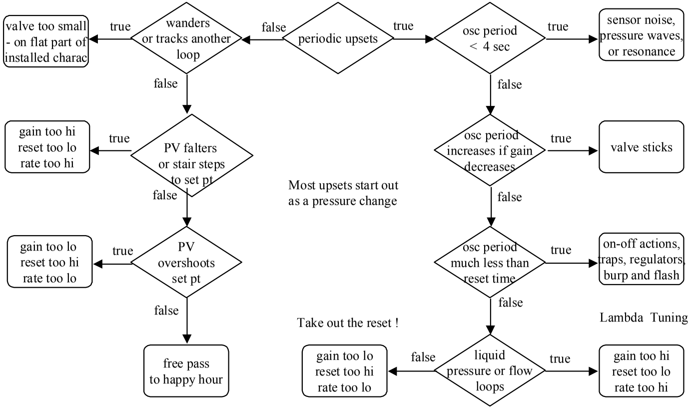


Figure 20 – Diagnostics for Loops in Manual

Figure 21 – Diagnostics for Loops in Automatic



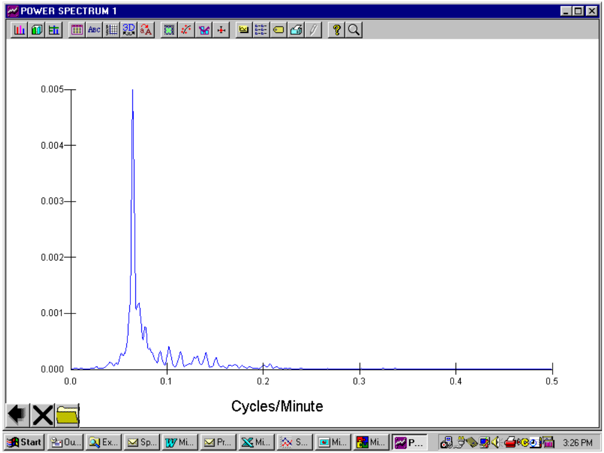
The oscillation can be made much faster and tighter by using higher gain and rate action.

Figure 22 is a spectrum analysis for a level, which shows that the predominant frequency is 0.08 cycles per minute. By analyzing the power spectrum of process variables in the application depicted in the figure, Walsh Automation was able to find the source of a cycle with a twelve-minute period in a series of distillation columns. It was an oscillation in the feed, which was caused by level switches on a feed tank that turned pumps on and off. This power spectrum analysis tool can quickly point to common periodic oscillations that noise and upsets would make it difficult to spot in trend recordings.

Rule 33 – **A power spectrum analyzer can rapidly point you to the culprit by indicating which loops experience significant peaks in the power at the same frequency.**

The first step is to enter the data for loops in automatic into the power spectrum analyzer. The data gathering must be done quickly enough to prevent aliasing. For chemical processes, it is sufficient to use data from a historian with no compression and an update time of 1 second. For sheets (webs), you may need to store the data using a device with an exceptionally fast

Figure 22 – Power Spectrum Analysis (Source: Courtesy of Walsh Automation, Toronto)



scan time (50 milliseconds or less) that is directly connected to the analog input terminals.

5.2 Great Expectations and Practical Limitations

When all is said and done, a loop will at best only pass on variability from its PV to its output. Variability in the output of one loop can greatly disturb another loop. Loops will actually increase the variability of the PV if there is a sticky valve, overly aggressive tuning settings, or upsets that have a period close to the natural period of the loop (two to four times the loop dead time). Most periodic upsets are caused by on-off actions and oscillating loops (which are often associated with level switches or level controllers that have too much gain or reset action). You can avoid a lot of headaches and wasted effort by first resolving valve problems and periodic upsets.

The most troublesome practice is using reset action correctly. It is not widely understood that the reset time as a factor of the loop period varies by an order of magnitude depending upon the degree of dead-time dominance and self-regulation. For example, for a process that tends to ramp or run away, the reset time should be larger than ten times the loop period. For a self-regulating process with a small dead-time-to-time-constant ratio, the reset time should be between 0.5 and 1.0 times the loop period. For pure dead-time processes, the reset time can be one quarter (1/4) of the dead time. Since the natural period is twice the dead time for a dead-time-dominant loop, the reset time ends up being about one-eighth (1/8) of the loop period. If a dead-time compensator is added to this loop, the reset time can be set to be about one quarter of the uncompensated dead time. If the dead time is underestimated by 10 percent, the reset time can be as low as one eightieth (1/80) of the loop period for a dead-time compensator.

To make it even more interesting, the rule of thumb to set the reset time greater than ten times the loop period for integrating and run-away processes is designed to essentially eliminate the possibility of oscillations due to reset action. There is a gain window for such non self-regulating processes. If the gain is below the low limit or above the high limit, there is a loss of control. The closer the controller is to the low gain limit, the more sensitive it is to reset action. Proportional action makes the response more self-regulating and thus, more tolerant of reset action. Some consultants recommend proportional-only controllers for level loops and proportional plus derivative controllers for exothermic reactor temperature loops to protect against instability caused by reset. In practice, some reset is useful to eliminate offset and facilitate startup. For level loops, it has been found that to prevent oscillations, the product of the controller gain and reset time must be greater than 4 times the fastest time for the level to ramp full scale. While the actual minimum product of gain and reset that triggers oscillations will change depending upon the amount of loop dead time and valve dead band, the relationship is useful to determine how to simultaneously adjust the gain and reset for level loops. If the loop response is smooth and the controller gain is doubled, the reset time can be halved and still keep the product of the two adjustments the same. It is critical to note that the opposite is true (the reset time must be doubled) if the loop is approaching the upper gain limit. If you are confused, don't feel alone. It is safe to say that 99% of users who don't carry this guide don't even have a clue that this strange twist of the rule on tuning reset even exists.

The following equation by Harold Wade shows how to eliminate oscillations near the low gain limit for a level controller if the dead time and valve dead band are negligible:

$$K_C * T_I > 4 * T_L$$

$$T_L = V_S / \Delta F$$

Where:

ΔF	=	maximum difference between inlet and outlet flows (gpm)
K_C	=	controller gain
T_I	=	controller reset time (minutes/repeat)
T_L	=	fastest level full scale ramp time (minutes)
V_S	=	volume for level measurement span (gals)

Finally, in order to tune the loop you need to be able to see the response. If the update time is too slow, the period of an oscillation and any time measurements will appear longer than actual. If 10% accuracy is desired, then the update time must be less than one tenth of the dead time or time constant. The scan times of any devices that supply data should be twice as fast as the trend update time to ensure sufficient over sampling. In order to get good accurate estimates of gains, the compression (value for exception reporting) should be smaller than the measurement or valve resolution. Otherwise the gains will be in error to the degree that the compression approaches the step size.

Rule 34 – **The update time of trend recordings should be less than one tenth of the loop dead time or time constant.** The compression of data should be less than one tenth of the step size and less than the measurement or valve resolution, whichever is largest.

If you understand and practice everything in this pocket guide, you might become the next CEO of your company. If this is not realistic, maybe you will be given stock options for every loop that is tuned. If this is a pipe dream, the improvement in loop performance might lead to a promotion. Well, maybe you will get a gift certificate to Burger King.

Seriously, I wish you all the best of luck in tuning loops.

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