How to Avoid Common Tuning Mistakes

Table of Contents Blurb: Ten PID tuning mistakes frequently occur in automation systems that can be avoided by a better understanding of the PID and tuning principles.

Fast Forward:

- Severe problems can be avoided by recognizing the PID Form and units
- Primary loops are oscillating from a counterintuitive effect of a low PID gain
- Measurements lags can make things look better while the process is worse

Headline: Why do most vessel control loops need the reset time increased by two or more orders of magnitude?"

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History

The PID controller is the common key component of all control loops. Basic control systems depend upon the PID to translate the measurement signals into setpoints of secondary loop controllers, digital valve controllers, and speed controllers for variable frequency drives. The success of advanced control such as model predictive control depend upon the basic control system foundation and hence the PID.

The first example of the PID was developed by Elmer Sperry in 1911 and the first theoretical analysis was published by Nicolas Minorsky in 1922. Ziegler and Nichols published papers on the ultimate oscillation method and reaction curve method for controller tuning in 1942 and 1943. While the parameters chosen as factors in the tuning settings provided overly aggressive control, the basic premise of an ultimate gain and ultimate period is essential to the fundamental understanding of the limits of stability. The identification of the slope in the reaction curve method is a key to the use of the near-integrator concept that we will find here to be critical for most composition, temperature, and pH loops to improve tuning settings and dramatically shorten test times.

Bohl and McAvoy published a paper in 1976 that showed the PID can provide nearly optimal control for unmeasured load disturbances. Shinskey wrote many books detailing the knowledge of process dynamics and relationships essential for the best application of PID control. Shinskey developed the original equation for integrated error from disturbances as a function of tuning settings as detailed in the January/February 2012 InTech article "PID tuning rules". Shinskey also published a book dedicated to PID controllers that showed that the simple addition of a dead time block in the external reset feedback path could further enhance the PID performance by dead time compensation.

Internal model control (IMC) and lambda tuning rules were developed based on pole and zero cancellation to provide a good response to setpoints and disturbances at the process outlet. However, most of the improvement in setpoint response could have been achieved by a setpoint lead-lag or PID structure and these tuning rules do not perform well for the more common case of disturbances on the process input (load upsets) particularly for lag dominant processes. Skogestadt developed significant improvements to IMC tuning rules. Bialkowski showed that always using lambda rather than lambda factors, relating lambda to dead time, and treating lag dominant processes as near-integrators enable the PID to provide good non-oscillatory control for load upsets besides dealing with the many different difficulties and objectives that lambda tuning was originally designed for. Not realized is that most methods converge to the same basic expression for the PID gain and reset time when the objective is load disturbance rejection and that a tuning parameter that is the closed loop time constant or arrest time is set relative to dead time. Also not recognized is how PID features, such as structure, external reset feedback, an enhanced PID for analyzer and wireless, a simple calculation of a future value, a valve position controller, and a "full throttle" setpoint response can increase process efficiency and capacity as noted in the ISA 2013 Book *101 Tips for a Successful Automation Career*.

Overload

The user is confronted with considerable disagreement of tuning rules not realizing most of them can be adjusted by factors or a near-integrator concept to achieve good control as seen in the 400 pages of tuning rules in the 2006 book by O'Dwyer.

The modern PID has many more options, parameters, and structures that greatly increase the power and flexibility of the PID but most are underutilized due to insufficient guidance. Additionally, ISA Standard Form used in most modern control system is not the Parallel Form shown in most text books and is not the PID Series form pervasively used in the process industry until the 1990s.

All of this can be quite overwhelming to the user, particularly since tuning is often done by a generalist with many other responsibilities faced with rapid changes in technology. My goal in my recent articles, books, and columns including blogs that are more extensive and less supplier specific than white papers is to provide a unified approach and more directed guidance based on the latest PID features that are missing in the literature. The recently completed *Good Tuning: A Pocket Guide, Fourth Edition* seeks to concisely present the knowledge needed and simplify the tuning by switching between just two sets of tuning rules largely depending upon whether the PID is a primary or secondary controller. A primary PID for vessel or column composition, gas pressure, level, pH, and temperature control uses integrating process tuning rules where the lambda arrest time is set. A secondary PID for liquid pressure, flow, inline pH and heat exchanger temperature control, uses self-regulating process tuning rules where the closed loop time constant is set. In both situations, lambda rather than a lambda factor is used and chosen relative to the dead time to provide the degree of tightness of control and robustness needed.

The best thing a user can do is to use good tuning software, attend supplier schools, and get a consultant onsite to provide the onsite solutions and practice. Also important is to take responsibility for avoiding common tuning mistakes. Here we step back to make sure we are not susceptible to oversights and misunderstandings. The following compilation has the most common and disruptive potentially unsafe mistakes near the top but all can be come into play and be important.

Mistakes

(1) Use of the wrong control action: In analog controllers and in many early distributed control systems (DCS) and Programmable Logic Controllers (PLC), the valve action affected only the display of the output on the station or faceplate. The specification of an "increase-to-close" valve action for a fail-open valve reversed the display but not the actual output. Consequently, the control action had to take into account the valve action besides the process action. If the valve was "increase-to-open" (fail close), the control action was simply the reverse of the process action (direct control action for reverse acting process and vice versa). If the valve was "increase-to-close" the control action was the same as the process action (direct control action for direct acting process and vice versa) if not reversed in the Current to Pneumatic (I/P) transducer or positioner. In today's systems, the user can specify "increase-to-close" in the PID block or analog output (AO) block besides the digital valve controller, enabling the control action to be set as the opposite of the valve action. The challenge is realizing this and making sure the increase-to-close valve action is only set in one place. If you don't get the control action right, nothing else matters (the PID will walk off to its output limit).

(2) Use of PID block default settings: The settings that come in with a PID block as it is dragged and dropped into a configuration must not be used. When first applying PID to dynamic simulations of new plants, typical settings based on process type and scale span can be used as a starting point. However, tuning tests must be done and settings adjusted before operator training and loop commissioning.

(3) Use of Parallel Form and Series tuning settings in the ISA Standard Form: A Parallel Form that uses integrator gain and derivative gain settings that are put into the ISA Standard Form as reset time and rate time settings can be off by orders of magnitude. A Series Form can provide good control with the rate time equal to or greater than the reset time because interaction factors inherently reduce the PID gain, and rate time and increase the PID reset time to prevent oscillations from the derivative mode contribution being greater than the contribution from the other modes. Using a rate time equal to or great than the reset time in an ISA Standard Form can cause severe fast oscillations.

(4) Use of the wrong units for tuning settings: Here we consider just the Series Form and ISA Standard Form. Controllers can have a gain or proportional band setting for the proportional mode. The gain setting is dimensionless and is 100% divided by the proportional band. Some PID algorithms in control studies and actual industrial systems have the gain setting in engineering units that leads to a very bizarre setting. The integral mode setting can be repeats per second, repeats per minute, minutes per repeat, or seconds per repeat. The units of these last two settings are commonly given as just minutes or seconds. The omission of the "per minute" can cause confusion in the conversion of settings. The conversion of the rate time is simpler because the units are simply minutes or seconds.

(5) Use of the wrong units for output limits and anti-reset limits: In analog controllers and in many early DCS and PLC systems, the output and consequently the output limits and anti-reset windup limits were in percent. In todays modern control system, the output is in engineering units and the limits must be set in engineering units. For valves, the units are usually percent of valve stroke. For a primary (upper) PID that is sending a setpoint to a secondary (lower) PID, the primary PID output is in the engineering units of the secondary PID process variable.

(6) Tuning of level controllers: If you calculate the calculate the product of the valve, gain, process gain, and measurement gain where the process gain is simply the inverse of the product of the fluid density and vessel cross sectional area, you realize the open loop integrating process gain is very small (e.g., 0.000001 1/sec) leading to a maximum PID gain for stability that is more than 100. For surge tank level control, a PID gain closer to unity is desired to absorb fluctuations in inlet flows without passing them on as changes to a manipulated outlet flow that will upset downstream users. Users do not like a high PID gain even when tight level control is needed. The decreasing of the level controller gain without a proportional increase in the reset time will cause nearly sustained slow rolling oscillations. Further decreases in the PID gain only make the oscillations worse. Most oscillations in production plants and poor performance of distillation columns can be traced back to poorly tuned level controllers. The solution is to choose an arrest time (lambda for integrating processes) to either maximize of the absorption of variability (e.g., surge tanks level control or distillate receiver level control where distillate flow is manipulated) or maximize the transfer of variability (e.g., reactor level for residence time control or distillate receiver level control where reflux flow is manipulated for internal reflux control). The integrating process tuning rules with arrest time first sets the reset time and then the PID gain to prevent violation of the window of allowable PID gains.

(7) Violation of the Window of Allowable Controller Gains: We can all relate to the fact than too high of a PID gain will cause oscillations. In practice what we see more often is oscillations from too low of a PID gain in primary loops. Most concentration and temperature control systems on well mixed vessels are vulnerable to a PID gain that violates the low PID limit causing slow rolling nearly undamped oscillations. These systems have a highly lag dominant (near-integrating), integrating, or runaway process response. All of these processes benefit from the use of integrating process tuning rules to prevent the PID gaining from being less than twice the inverse of the product of the open loop integrating process gain and reset time preventing the oscillations shown in Figures 1a, 1b, and 1c. The oscillations in the figures could have been stopped by increasing the reset time. In industrial applications, the reset time in vessel control loops often needs to be increased by two or more orders of magnitude. Note that the oscillations get worse as the process loses internal self-regulation going from a near-integrating (low internal negative feedback) to an integrating (no internal feedback) and to a runaway (positive feedback) open loop response. For runway processes there is also a minimum gain setting independent of reset time that is the inverse of the open loop runaway process gain. The identification of the open loop integrating process gain can generally be done in the about 4 dead times, greatly reducing the test time reducing the vulnerability to load upsets.

(8) Lack of recognition of sensor lag, transmitter damping or filter setting effect: A slow measurement response can give the illusion of better control. If the measurement time constant becomes the largest the time constant in the loop, the PID gain can be increased and oscillations will be smoother as the measurement is made slower. This occurs all the time in flow control, pressure control, inline pH control, and temperature control of gas volumes since the process time constant is less than a second. The real process variability has increased and can be estimated via a simple equation. For more on this wide spread problem see the 12/02/2014 Control Talk Blog "Measurement Attenuation and Deception". For details on how to prevent this in temperature control systems see the 2/16/2015 ISA Interchange post "Temperature Sensor Installation for Best Response and Accuracy"

(9) Failure to do tuning tests at different times, setpoints and production rates: The installed characteristics of most control valves and most concentration, pH, and temperature processes are nonlinear. The process gain varies with operating point and process conditions including relatively unknown changes in catalyst activity, fouling and feed compositions. The valve gain varies with system resistances and flow required. For operating point nonlinearities, the open loop process gain identified depends upon the step size and direction and the split range valve being throttled. Temperature process time constants also tend to vary with the direction of the change. For more details see the 1/19/2015 Control Talk Blog "Why Tuning Tests are Not Repeatable".

(10) Failure to increase PID gain to decrease backlash limit cycle amplitude: An attempt to decrease oscillation amplitude by decreasing gain will make the oscillation worse when the oscillation is a limit cycle from backlash (deadband). The amplitude from backlash is inversely proportional to the PID gain. The limit cycle period from backlash or stiction is also increased as the PID gain is decreased reducing the attenuation from the filtering effect of process volumes. The same equation noted in item (8) can be used to estimate the attenuated amplitude at the outlet of a well-mixed volume by using the residence time (volume divided by throughput flow) as the filter time constant.

About the Author

Gregory K. McMillan is a retired Senior Fellow from Monsanto-Solutia and an ISA Fellow. McMillan received the ISA "Kermit Fischer Environmental" Award for pH control in 1991, the *Control* magazine "Engineer of the Year" Award for the Process Industry in 1994, was inducted into the Control "Process Automation Hall of Fame" in 2001, was honored by *InTech* in 2003 as one of the most influential innovators in automation, and received the ISA Life Achievement Award in 2010. McMillan is the author of numerous books on process control and has been the monthly "Control Talk" columnist for *Control* since 2002. He has a Control Talk blog at http://community.controlglobal.com/controltalkblog and provides posts on ISA Interchange at http://automation.isa.org/author/gregmcmillan/.

War Stories

1. The trend charts of phosphorous furnace pressure from faster pressure transmitters installed looked worse even though the number of high pressure reliefs had been dramatically reduced. Fortunately, the older slower transmitters were left installed showing that the amplitude of the pressure excursions had actually decreased after the faster transmitters were used for furnace pressure control.
2. A plant operated for several years with default tuning settings of a gain and reset settings (repeats per minute) both equal to one for all of the PID controllers. Nearly every loop was oscillating but the plant ingeniously managed to run by setting output limits to reduce oscillation amplitudes.
3. When a plant converted from analog controllers to a DCS, the plant was amazed at the improvement in the distillation column control. It turns out the configuration engineers didn't realize the difference between PID gain and proportional band (PB). The analog controller for column overhead receiver level manipulating reflux had a PB of 100% that was then set as a gain of 100 in the DCS PID. The tight level control and consequential great internal reflux control stopped the slow rolling oscillations from violation of the PID low gain limit and rejected disturbances from "Blue Northerner" cold rain storms.