EFFECTIVE USE OF KEY PID FEATURES

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Key words: advanced regulatory control, analyzer control, analyzer failures, anti-reset windup, communication failures, external-reset feedback, integral deadband, nonlinear gain, output tracking, PID structure, smart reset, setpoint filter, setpoint rate limit, setpoint tracking, threshold sensitivity, valve position control, wireless control

ABSTRACT

The PID controller has evolved over the decades to include a large number of features to address common application opportunities. Many of the features are not effectively used due to lack of a consolidation of functional understanding and best practices. The number of options and parameters can seem overwhelming and the best choices unknown.

The capability of the PID is experiencing a renaissance triggered by the advent of smart wireless instrumentation. A particularly key advancement is an enhancement for wireless that has unexpected benefits for simplifying and stabilizing tuning for loops with large and highly variable update times from at-line and off line analyzers, stopping limit cycles, reducing interactions, and recovering from failures. An emerging capability for smart reset action offers the ability to specify when a PID output comes off an output limit and to prevent overshoot or faltering in an approach to setpoint.

There is also an increasing recognition that a key feature where the controller output cannot change faster than the manipulated variable can enable the user to tune for maximum disturbance rejection and still meet other objectives such as coordination and decoupling of loops. Furthermore the PID can be easily extended to unit operation optimization problems usually reserved for model predictive control.

This paper offers guidance for getting the most out of your PID extending its functionality from basic control to advanced control by the use of important existing and new features for applications encountered in the process industry. The goal is not only to show the flexibility but also how an understanding of the functionality and synergism of features can result in a simplification of tuning the PID.

INTRODUCTION

The basic control applications requirements vary due to the diversity of unit operations and the process industries (beverages, crops chemicals, films, foods, health and beauty consumer products, industrial chemicals, mining, oil and gas, petrochemicals, pharmaceuticals, pulp and paper, sheets, textiles, utilities, and waste treatment). The PID has developed an incredible spectrum of features to address these application requirements. This paper is an attempt to provide a consolidated source of guidance on how to take advantage of the incredible flexibility of the PID.

ANTI-RESET WINDUP

All industrial PID controllers have an anti-reset windup algorithm. The method of implementation varies with the automation system supplier. Most limit the contribution from the integral mode so that the total output does not go above the high output limit or below the low output limit. Before the advent of smart digital positioners, the low output was set below 0% (e.g. -10%) to ensure the valve would be completely closed. The high output limit was set above 100% (e.g. 105%) to ensure the valve could be fully opened. The calibration shifts in pneumatic positioners was large and unknown. Now the output limits should generally be set at 0% and 100% for valves. For variable frequency drives the low output limit must be set to prevent the pump discharge head going below the highest static head (e.g. destination pressure) so that a disastrous case of reverse flow cannot occur. The output limits should also match the setpoint limits of secondary loops, an option available in most PID controllers.

Some Distributed Control Systems (DCS) offer an anti-reset windup limit separate from the output limit. One particular DCS supplier increases the integral action by a factor 16 when the PID output was between the output limit and anti-reset windup limit. This was effectively used for valves with high degrees of stick-slip near the closed position not adequately addressed by a positioner. For example, some tight shutoff rotary valves had such a high breakaway torque requirement, that the valve would remain stuck closed until the controller output was larger than 25%. In this case, the anti-reset windup low limit was increased to 25%. In other cases, the extreme flattening of the installed characteristic at high valve positions was addressed by decreasing the anti-reset windup limit to 75%. In all cases, the correct solution is to get a properly sized actuator and control valve with less stiction, a better installed characteristic, and a smart digital positioner. Presently, the anti-reset windup limit should be set equal to the output limit so that conventional tuning rules and smart reset algorithms can be used.

There is an optimum point for when a controller output will come off an output limit to achieve a new setpoint or to return after a large disturbance to an existing setpoint. Good tuning can achieve the optimum but the settings can be difficult to estimate especially for integrating processes where there is a non-intuitive effect caused by a window of allowable PID gains and reset times. The product of the

PID gain (K_c) and reset time (T_i) needs to be greater than or equal to twice the inverse of the integrating process gain (K_i) per Equation 1 to prevent slow rolling oscillations.[1,2]

$$K_c * T_i \geq 2 / K_i \quad (1)$$

In most cases, overshoot is undesirable. The PID output must come off of the limit based on the rate of change of the process variable (PV) in the approach to setpoint (SP) and the deadtime in the system to prevent overshoot. The effect of any change in PID is not seen until after the total loop deadtime. How far the PV travels based on previous actions is the rate of change of the PV multiplied by the deadtime. Thus, there needs to be some anticipatory intelligence. However some users want to simply specify a trigger point for when the PID output comes off the output limit. The trigger point is then biased to prevent overshoot. If the trigger point is when the PV is at SP, the setpoint is then biased to prevent the PV from reaching an undesirable limit. The classic case is the anti-surge controller. Some users do not want the surge valve to open until the PV crosses the SP. In this case the SP is shifted sufficiently away from the surge curve so that overshoot does not drive the compressor into surge.

As the PV approaches the SP, the contribution from the proportional and derivative modes will be working to get the PID output off of the limit but the integral mode will be driving the output into the output limit. To get off of the output limit, the contribution from the integral mode must be less than the contribution from the proportional and derivative modes. The smart reset section shows how the reset time can be adapted to accomplish this specification and to prevent overshoot and faltering in the approach to setpoint.

EXTERNAL-RESET FEEDBACK

If the output of the PID changes faster than whatever the PID is manipulating can respond, the PID can burst into oscillations. To make the situation confusing, the response can be smooth for small setpoint changes and for small disturbances. The situation can arise due to a slow exponential response or rate limit in the response of a large control valve or damper, a variable frequency drive, or a secondary loop tuning. Control valves with large actuators have a rate limited exponential response where the stroking time can be more than a minute. Variable frequency drives often have rate limits in the setup to prevent overloading the motor. Secondary loops can have a slow response due to rate of change limits and filters on the setpoint, detuning, and low process or valve gains. For example the valve gain; slope of the installed valve characteristic; can become exceptionally low; curve extremely flat; for large positions of rotary valves. For heat exchangers, the process gain for large ratios of coolant to process flow can become exceptionally small for a low heat transfer coefficient.

Most publications show an integrator for the implementation of integral action in a PID. The positive feedback implementation of integral action offers many advantages. A filter inserted in the positive feedback path provides the integrating action. The filter time constant is the integral time. When the error is zero, the input at point 1 in Figure 1 is zero and the output of the filter is equal to the input to the filter stopping a further change at point 2. The constant filter output suspends integral action and leads to constant output from the integral mode.

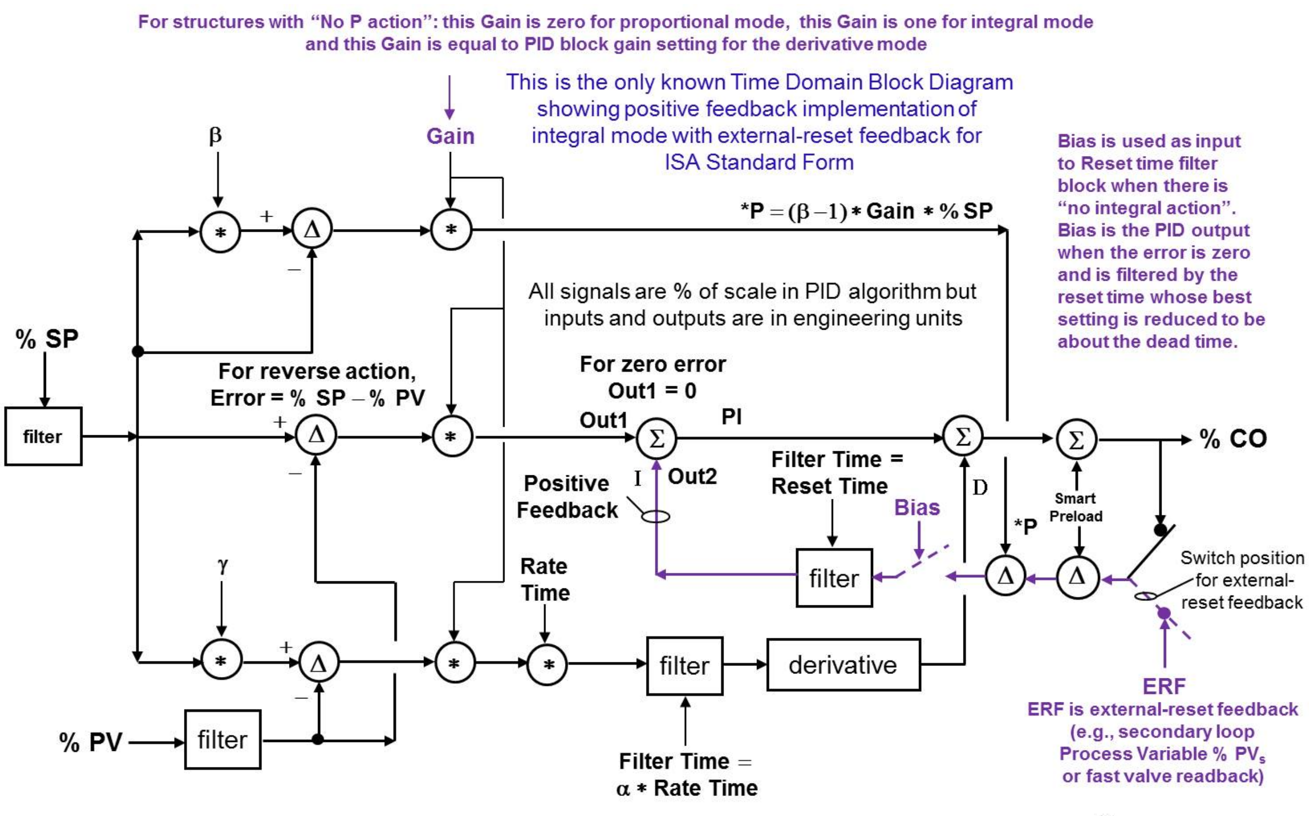

Figure 1 Positive feedback implementation of integral action for standard form PID

If the external-reset (E-R) feedback comes from a secondary process variable (%PV_s) of whatever the controller output (%CO) is manipulating, the controller output cannot change faster than the %PV_s can respond. In some PID controllers, this option is called a "dynamic reset limit." The most common secondary loop is a flow because nearly all types of composition, level, pH, and temperature loops can benefit from a secondary loop compensating for pressure disturbances and valve nonlinearities and affording flow feedforward. For crystallizer or reactor temperature control, a secondary coolant, exchanger, or jacket temperature loop isolates coolant disturbances and process nonlinearities from the primary temperature loop. In all cases of cascade control, an external-reset feedback signal can enable the primary loop to be tuned for best disturbance rejection; fastest response; without worrying about the consequences of a secondary loop not being able to keep up with the changing demands from the primary loop.[3] The only requirement is to have integral action configured in the secondary controller so that the primary's %CO and %PV_s are equal in steady state.

The final control element is most often a control valve. Most control valves bought this century have a smart digital positioner that is secondary valve position control loop. Today's positioners now have gain, reset time, and rate time settings whereas pneumatic positioners were proportional-only. For small actuators, the positioner response is normally fast enough to meet the demands of flow loops. The main need for external-reset feedback in loops with smart digital positioners is for large actuators and a positioners tuned for too slow of a response. Unfortunately, the use of secondary HART variable may not be fast enough. The update time of valve positioner readback must be faster than the valve pre-stroke deadtime and the valve resolution or threshold sensitivity divided by the valve slewing rate.

For liquid or polymer pressure control, the control valve may not be fast enough. Here a variable frequency drive (VFD) should be considered because there is essentially no deadtime and no backlash or stick-slip. Increasingly, a VFD is becoming an option to save energy and to improve the resolution and threshold sensitivity for flow and pressure control[4,5]. The speed and/or torque control loop should be in the drive to keep the VFD loop much faster than the flow or pressure loop. If rate limiting is introduced in the drive or the speed loop is moved to the control room, speed should be used as an external-reset feedback to prevent the process loop outrunning the VFD loop.

External reset (ER) can be used in cascade loops to keep the primary PID from winding up as the error gets integrated. ER stops integration when the secondary PID PV is constant and thereby converts the integral contribution to a constant bias[12]. External-reset feedback also enables the use of setpoint filters and directional rate limits on analog output AO blocks and secondary PID controllers. In these cases the process variable of the AO and PID being manipulated is configured to be the source of the external-reset feedback. The opportunity for coordination of loops, simplification of tuning, optimization of setpoints, and the minimization of overshoot and interactions will be discussed in the sections on Setpoint Filter and Rate Limits.

STRUCTURE

Table 1 Major PID Structure Choices

(1) PID action on error ($\beta = 1$ and $\gamma = 1$)
(2) PI action on error, D action on PV ($\beta = 1$ and $\gamma = 0$)
(3) I action on error, PD action on PV ($\beta = 0$ and $\gamma = 0$)
(4) PD action on error, no I action ($\beta = 1$ and $\gamma = 1$)
(5) P action on error, D action on PV, no I action ($\beta = 1$ and $\gamma = 0$)
(6) ID action on error, no P action ($\gamma = 1$)
(7) I action on error, D action on PV, no P action ($\gamma = 0$)
(8) Two degrees of freedom controller (β and γ adjustable 0 to 1)

The structure choices use β and γ set point weighting factors shown in Figure 1. A controller that has both factors adjustable is called a “two degree freedom controller.” Other structures have the β and γ factors set equal to 0 or 1. The user can also omit a mode entirely to get P-only, I-only, ID, and PD control with various assigned factors. PI control is achieved by simply setting the derivative (rate) time to zero. In general, the user must not set the controller gain equal to zero in an attempt to get I-only or ID control or set the integral (reset) time to zero in an attempt to get P-only or PD control. Note that the use of P-only or PD control requires additional choices of how to set the bias and its ramp time. Table 1 lists 8 choices offered by one major DCS supplier.

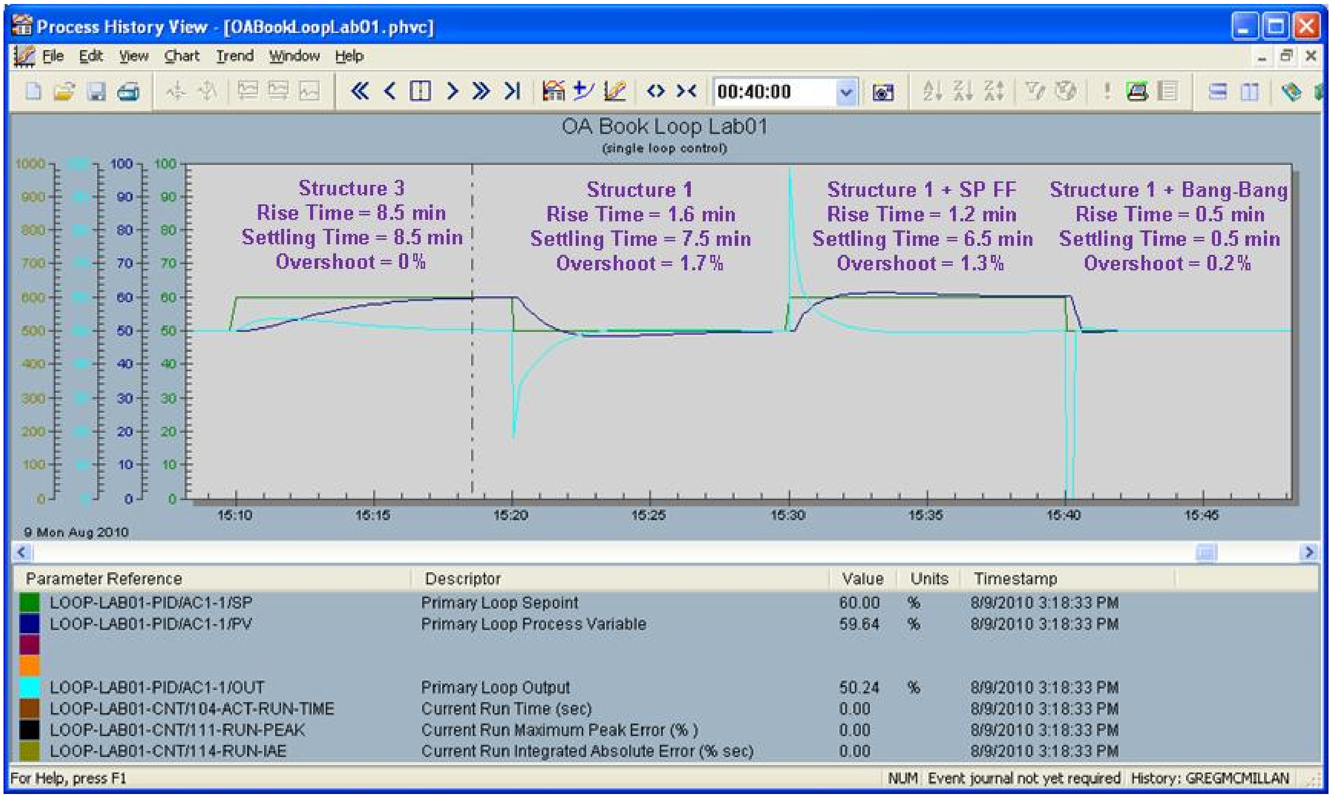

Figure 2 Effects of structure, setpoint feedforward, and Bang-Bang logic on setpoint change for an integrating process

Structure 1 (PID action on error) provides the fastest approach to a new setpoint by virtue of a bump in the controller output from the setpoint change as seen in the second response in Figure 2. For small setpoint changes and low controller gains, the bump can help get through significant valve backlash and stick-slip to get the valve moving. The bump appears to be a spike on trend charts with a large time spans. The abrupt change in output is often seen as disruptive by operators when they make setpoint changes. If the burst of flow through the control valve does not affect other users of the process or utility fluid, the bump is more of a psychological than a process concern. The bump can be made smaller by decreasing the gamma factor (γ). At any rate, the reduction in rise time; time to reach setpoint whether the PV is increasing or decreasing; is marginal for good control valves or higher controller gains and larger setpoint changes.

The PV overshoot of SP in Figure 2 response 2 that results from load disturbance tuning can largely be eliminated by adding a setpoint filter equal to the reset time as discussed in the Setpoint Filter section with a marginal increase in rise time. The use of setpoint feedforward can slightly reduce the overshoot and rise time as seen in response 3 in Figure 2. The advantage of setpoint feedforward is more significant when the controller gain is relatively low. Figure 3 shows the flow of information in a control loop configured with a SP filter and SP feedforward.

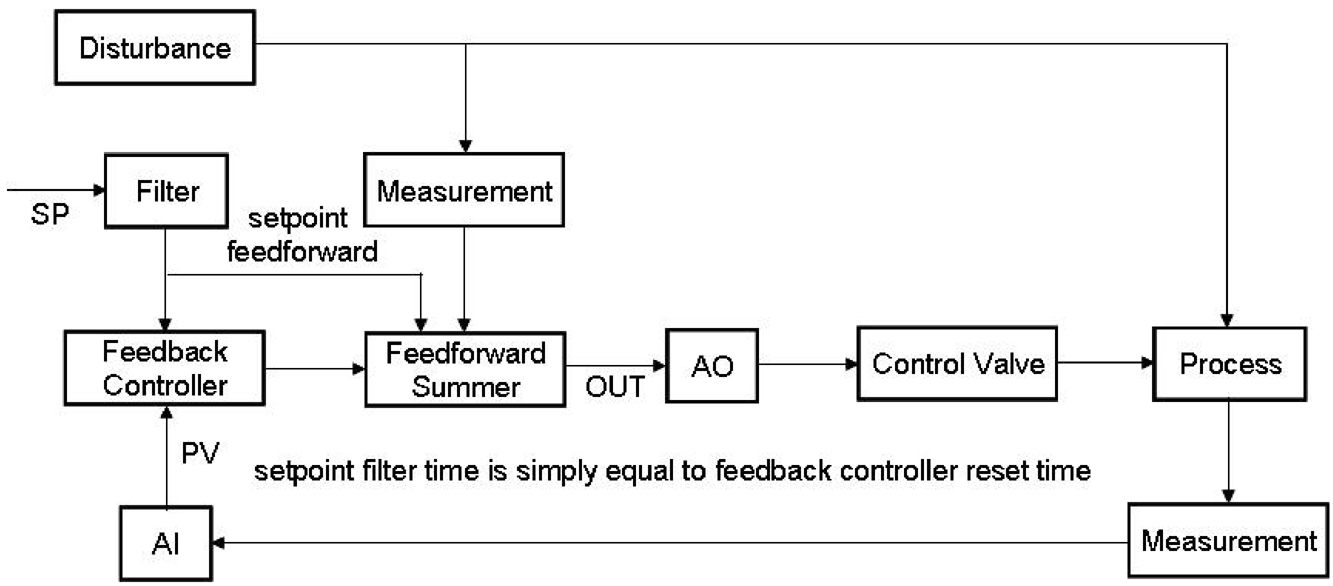

Figure 3 Configuring a SP filter in the feedback controller prevents overshoot and adding SP feedforward reduces rise time[11].

The biggest improvement is offered by a relatively simple Bang-Bang logic described in the Output Tracking section. In response 4 of Figure 2, Bang-Bang logic almost completely eliminates overshoot and provides the fastest possible rise time.

Structure 2 (PI action on error, D action on PV) is the structure most often used. Structure 2 eliminates the bump from derivative action for a setpoint change by setting gamma to zero (γ=0). The increase in rise time going from structure 1 to 2 is negligible for the more important loops, such as column and vessel temperature where derivative action is used. The step in the output from the proportional mode on a setpoint change is large because of the high controller gain. Increases in process gain or deadtime will increase the overshoot unless the controller gain is decreased accordingly. If the elimination of setpoint overshoot is much more important than rise time, then structure 3 may be best.

Structure 3 (I action on error, PD action on PV) eliminates overshoot but with quite a sacrifice in speed of approach to the setpoint. In Figure 2 for an integrating process, response 1 for structure 3 has a rise time 5 times larger than the rise time for structure 1 in response 2 for the same tuning settings and process dynamics. In Figure 4 the rise time for structure 3 is only twice as large as the rise time for structure 2 because the process is self-regulating. A setpoint filter set equal to the integral time could have essentially eliminated the overshoot for all of the cases shown in Figure 4.

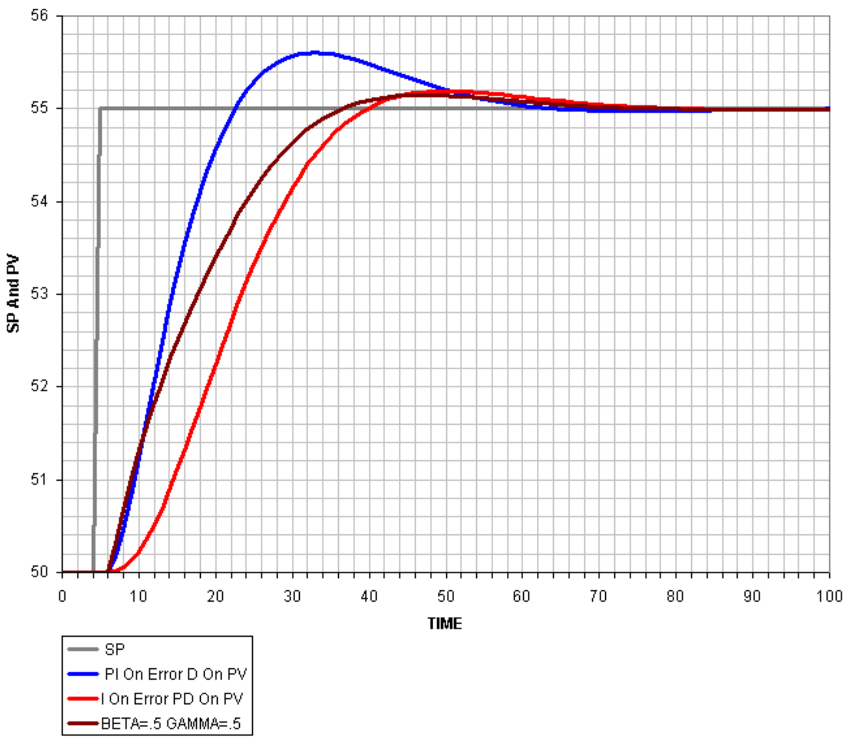

Figure 4 Self-regulating process response to a setpoint change for three different structures[14].

Structure 4 (PD action on error, no I action) is used on processes adversely affected by integral action. The temperature control of severely exothermic polymerization reactors use structure 4 because integrating action in the controller increases the risk of a runaway. If integral action is used, the reset time should be increased by a factor of 10 for these positive feedback processes to be safe. Users may not be aware of this requirement leading to overshoot that can trigger a runaway. The bias for structure 4 is set equal to the normal PD controller output when the PV is at setpoint.

Structure 4 is used for total dissolved solids (TDS) control of boiler drums and vessel level control to eliminate the slow reset cycles from too small of a reset time and to prevent slow rolling oscillations from violation of Equation 1. For drums, the boiler blow down may be discontinuous making control of the TDS integrating response more difficult with integral action. For reactors, the increase in level from proportional-only control for a decrease in reactant feed flow provides a more constant residence time. However, the setting of the bias in these applications is confusing to the user.

Structure 4 is used on batch processes that respond in only one direction. For example, in bringing a batch pH up to a setpoint by the addition of a base where the base is not consumed in a reaction, the batch will only respond in the direction of increasing pH. The pH will overshoot setpoint if integral action is used. If split ranging is added with an acid reagent, there will be some wasted reagent due to cross neutralization of reagents and limit cycling across the split range point from stiction that is greatest near the closed position. For structure 4 and a single reagent, the bias is set for zero reagent addition when the PV is at setpoint.

Structure 5 (P action on error, D action on PV, no I action) is used for the same reasons as structure 4. As with structure 2, the spike from rate action for setpoint changes is eliminated.

Structure 6 (ID action on error, no P action) is used for valve position controllers (VPC) to eliminate the interaction with process controller whose valve position is being optimized. The VPC could be optimizing the coarse adjustment from a large control valve in parallel with a small control valve manipulated as fine adjustment by the process controller. In the example below the process controller FC215 adjusts the small valve input to the process. The output to the small valve is also used as a PV of the VPC ZC215,. The small valve provides precision and the large valve gives rangeability.

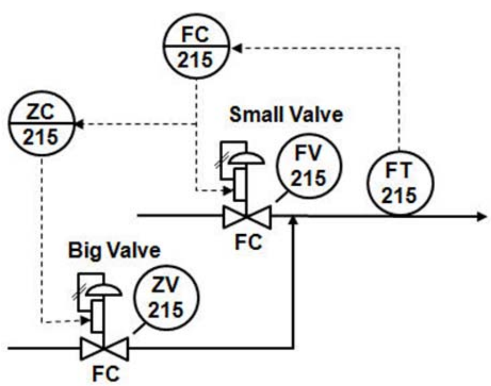

Figure 5 Valve position control increases precision and rangeability [13]

The VPC could be optimizing utility supply pressure or temperature to minimize energy use or optimizing feed rate to maximize production rate. Normally, the rate time is zero. The tuning of this integral-only controller is problematic.[7] Tuning rules often cited are the integral time should be larger than 10 times the product of the gain and reset time of the process controller and 10 times the residence time of the process to eliminate interaction. However, the VPC response for this tuning is too slow to prevent the process controller from getting into trouble for large disturbances. Feedforward action can be added to help, but a more flexible and easier to tune solution is to use external-reset feedback, AO block directional rate limits, and a PID developed for wireless. This solution is described in the Setpoint Rate Limits section.

Structure 7 (I action on error, D action on PV, no P action) is used for the same reasons as structure 6. As with structure 2, the spike from rate action for setpoint changes is eliminated.

Structure 8 is used to provide a balance between a fast rise time and minimal overshoot. A setpoint filter equal to the reset time is a simpler solution. This structure can be used to get the structures with proportional or derivative action on PV by zeroing the respective β and γ set point weighting factors.

INTEGRAL DEADBAND

Limit cycles; sustained equal amplitude oscillations regardless of tuning settings; will occur from stick-slip and backlash if there are at least one or two integrators, respectively, in the process or the control system. Integral deadband; the positive and negative error setting for complete suspension of integral action; will stop limit cycles from stick-slip and backlash if there is zero and one remaining integrator, respectively. The integrators can originate from integral action in control systems; e.g. PID controllers or digital positioners; or an integrating process response; e.g. level, vessel pressure, and batch temperature. For a triple cascade control loop; e.g. reactor temperature to coolant temperature to coolant flow; there can be three integrators in this control system. Since integral action is now offered in digital positioners, there is a fourth possible integrator.

For an integrating process response, integral deadband cannot stop a limit cycle from stick-slip. In this process, integral deadband can stop a limit cycle from backlash if there are no integrators in the control system. For a self-regulating process, integral deadband can stop a limit cycle from stick-slip if integral action is suspended in every PID controller. In this process, integral deadband can stop a limit cycle from backlash if there is only one integrator in the control system. Users need to realize that choosing to use integral action in a positioner necessitates the selection of an integral deadband in the positioner.

The integral deadband needs to be set greater than the half amplitude of the limit cycle assuming the limit cycle is centered about the PID or positioner setpoint. The stick-slip limit cycle amplitude is equal to the stick-slip multiplied by the open loop gain.[6] The degree of stick-slip changes with position and age of the valve and the open loop gain changes with valve characteristic and process operating point. Shaft windup from high rotary valve seal friction, causes a large unpredictable slip. The deadband limit cycle amplitude is the deadband divided by controller gain. Deadband can vary considerably with position when a linear actuator is used on a rotary valve. For these and other reasons, getting the correct integral deadband setting is difficult. An easier alternative is enabling the feature developed for wireless discussed in the enhanced PID section.

In applications where loss-in-weight feeders (LIWF) are used to maintain a downstream blender's level around a given setpoint, the integral deadband is a useful feature. Loss-in-weight feeders require relatively large and less frequent step changes to its setpoint in order to deliver error free mass flow. By configuring the integral deadband the level controller only acts for large corrections thus helping to reduce the frequency of rate changes to the LIWF controller. In some applications an extra deadband is configured in the controller's output. The feeders are only updated with a new rate setpoint when the change exceeds the deadband. However, a deadband on the feeder setpoint will introduce a limit cycle because level is an integrating process.

Integral deadband has also been found to be effective at reducing noise in motors as it lowers the frequency of small corrective commands to hold position. The integrator stops worrying about small amounts of windup and only kicks in for larger deviations[15].

NONLINEAR GAIN

A nonlinear gain reduces the controller gain when the PV is within the nonlinear gap or notch around the setpoint. The use of a nonlinear gain can reduce the reaction to noise and can help reduce the oscillations for pH systems where the setpoint is on the steepest portion of the titration curve (e.g. 7 pH setpoint). The concept here has been applied to surge level control but users often forget an increase in reset time must accompany a significant reduction in controller gain per Equation 1 to prevent slow rolling oscillations. Since reset times are often too small to begin with on integrating processes, the use of a nonlinear gain makes the oscillations more sustained. The oscillations can propagate through downstream equipment. The only saving grace is that these oscillations are generally extremely slow.

Since all of the tuning settings can be written to programmatically in today's DCS, a much more flexible scheduling of tuning settings is possible than what is offered by the nonlinear gain. The nonlinear gain was needed more in the 1980s DCS where writing to the controller gain would bump the controller output.

OUTPUT TRACKING

The most common use of output tracking is to make a bumpless transfer to cascade control. The PID output of the primary loop tracks the PV or SP of the secondary loop. If upsetting the secondary loop is more of a concern than upsetting primary loop in the transition to cascade, the option to track PV is typically selected provided the PV is in a reasonable range. For startups where secondary setpoints are initialized to a desirable operating point, tracking of SP is normally selected. This case commonly occurs in extrusion processes where primary PID controllers take turns to control one same secondary control loop during startup. These primary controllers are normally in pressure control loops that manipulate motor or gear pump speeds. Tracking of SP is selected to ensure a smooth transition between operating points and master controllers.

Output tracking is also used for prepositioning a control valve. The PID is put in output tracking with more optimum output(s) and then released to feedback control after an optimum time. This strategy is particularly useful on startups and product transitions in continuous processes, setpoint changes in batch operations, and open loop backups to prevent environmental excursions and compressor surge. The basic logic is outlined for the following applications:

(1) **“Head-Start” logic for startup and batch setpoint changes**: For a setpoint change the PID output tracks a desired output for the expected rise time minus the total loop deadtime. The desired output is the final settling value; final PID output; for the last or best batch or startup.

(2) **“Bang-Bang” logic for startup and batch setpoint changes**: For a setpoint change the PID output tracks the appropriate output limit until the future PV is close to the new setpoint. At this time the PID output tracks the final settling value for one total loop deadtime.[8] The future PV is a PV predicted one deadtime into the future by the use of a deadtime block whose input is the new PV and whose output is the old PV. The old PV is subtracted from the new PV to create a delta PV. The delta PV is simply added to the new PV to create the future PV.

(3) **“Open Loop Backup” to prevent compressor surge:** When compressor flow PV drops below surge setpoint or a precipitous drop occurs in flow, the PID output tracks a surge valve signal that provides a flow more than large enough to compensate for the loss in downstream flow for a time larger than the total loop deadtime plus the surge period.

(4) **“Open Loop Backup” to prevent RCRA pH violation:** When an inline pH system PV approaches the RCRA pH limit (e.g. 2 and 12 pH), the PID output tracks a signal that incrementally opens a reagent valve (e.g. 0.25% per sec) until the pH sufficiently backs away from the RCRA limit (trigger point plus deadband).

SETPOINT FILTER

A setpoint filter can eliminate overshoot and coordinate the timing of feeds for flow ratio control. If the setpoint filter is set equal to the reset time, tuning for maximum disturbance rejection can be used with minimal overshoot.[1] The rise time can be reduced by adding a lead that is about 1/10 the filter time (lag time) but the improvement may not be worth the introduction of a lead-lag.

A setpoint filter can coordinate the timing of feeds so that there is not a temporary unbalance or transient error in the flow ratio control for changes in production rate. If flows are ratioed to the setpoint of the leader flow and setpoint filters on all the flow loops are set to make the closed loop time constant the same for all the flows, the ratio is more tightly enforced. The setpoint filter time is set to match the slowest flow loop response. The elimination of transient errors in the ratio of feed is particularly important for blending systems because there is no back mixing to smooth out the transients. For pulp and paper and plastics extrusion, the transients and temporary unbalances end up as variations in sheet quality. For reactors, the accumulated excess or deficiency in reactants from less than optimum ratios during transients can total to a decrease in yield.

The primary process loop manipulating the lead flow setpoint must have external-reset feedback of the secondary loop PV so that the primary loop tuning is independent of the setpoint filter time. For cascade control where the main purpose of the secondary loop is secondary disturbance rejection, the introduction of setpoint filter will slow down the ability of the secondary loop to do its job. Also, a secondary PID setpoint filter time that is greater than ¼ the closed loop time constant of the primary PID will cause a violation of the cascade rule where the secondary loop should be 4 times faster than the primary loop resulting in an interaction between the two loops. For these and other reasons, the indiscriminate use of setpoint filters in secondary loops is inadvisable.

SETPOINT RATE LIMITS

Setpoint rate limits can be used to provide a slow approach to an optimum and a fast getaway from problems. The use of external-reset feedback eliminates the need to retune the PID for these limits.

The addition of directional rate limits on the setpoint of the pressure, temperature, or feed rate being optimized coupled with external-reset feedback in the VPC enables move suppression commonly touted in model predictive control (MPC) with the additional benefit not found in MPC of the suppression being different for an increase versus a decrease in controller output. For example, a VPC to increase reactant feed rate to an exothermic reactor to maximize coolant valve position, would slowly increase feed rate for a decrease in coolant valve position but would rapidly decrease feed rate for an increase in coolant valve position so the reactor temperature loop does not run out of coolant valve. External-reset feedback enables the rate limits to be set and adjusted without retuning the VPC.

The addition of directional rate limits on the setpoint of analog output block of a surge controller enable a slow approach to the surge curve via a slow closing of the surge valve and fast get away from an impending surge by a fast opening surge valve. This fast opening of the fail open surge valve was originally accomplished by a quick exhaust valve. The tuning of the controller suffered from the discontinuity from the tripping of the quick exhaust valve.

Setpoint rate limits could also be used in startups where temperatures, pressures and speeds are being ramped to operational levels making the configuration simpler.

In process rolls where pipes are welded internally, if the internal pipe heats up or cools down much quicker than the shell of the roll, the welds on the pipe could break because of the tension or compression from the cooling or heating. The implementation of a setpoint rate limit in the temperature controller prevents damage to the equipment.

SETPOINT TRACKING

The user has the choice of the setpoint remaining at last value or tracking the PV when in manual. Setpoint tracking of PV is sometimes not used in order to retain a setpoint. However, there are some consequences that might justify saving and rewriting the setpoint. The tracking of PV eliminates the bump in the loop upon transition to auto, cascade, or remote cascade. In extruders where torque and pressure control loops are enabled or disabled at different stages of the startup, the setpoint tracking of PV avoid controller's output to kick in and pressure and torque to runaway. As the control loop action is enabled, the SP and PV are close enough and setpoint can ramped to the desired operating point.

Often more importantly, the change in setpoint may be beneficial in that proportional action on error will cause a step change in output to get the PV to setpoint faster. If setpoint tracking of PV is not used, the approach to setpoint is slow and high gain action on level and temperature loops and low integral action may cause the PV to falter or hesitate in the approach to setpoint. In some cases where tanks were being filled on startup, the retention of the setpoint while in manual caused the feed valve to close when the level started to rise after the level controller was switched to auto.

SMART RESET

The use of deadtime for the closed loop time constant for maximum disturbance rejection gives the same expression for controller gain for the major tuning methods. However, the reset time for non-oscillatory maximum disturbance rejection, varies tremendously for self-regulating processes depending upon ratio of the open loop time constant (largest time constant in loop) to total loop deadtime. For ratios much less than 1.0 (deadtime dominant), the optimum reset time approaches ¼ of the deadtime. For ratios much greater than 1.0 (lag dominant), the optimum reset time approaches 4 times the deadtime. Furthermore, pole cancellation tuning methods give a reset setting that can be more than a magnitude larger than the optimum for disturbance rejection.

Too much reset action can cause overshoot. Too little reset action can cause a faltering in the approach to setpoint. Additionally, the violation of Equation 1 can cause slow rolling oscillations and a more sustained overshoot when the controller gain is decreased for integrating processes.

Smart reset action can address these problems. The consequences of a wrong reset time adjustment are generally not as immediately severe as for a wrong gain adjustment. The minimum and maximum reset time is set based on a factor the user enters for the deadtime.

Smart reset action can get the PID output off the output limit at a specified deadband about setpoint. The ramp rate of the PV is used to compute the rise time to get within the deadband. The current error minus the deadband is divided by the ramp rate to get the rise time. The ramp rate and rise time is used to compute the relative contribution of the each PID mode. The output will come off of the output limit when the contribution of the integral mode exceeds the contribution from the proportional and integral modes. Equations 2a - 2h show the derivation of the calculation that sets the reset time based on the user choice of deadband (error that causes PID output to come off limit). If the PID output is between the ARW and output limits, the calculation must take into account the 16x multiplier of reset time.

$$E_{db} = |SP - PV| - DB \quad \text{(2-a)}$$

$$P_{PID} = K_c * |E_{db}| \quad \text{(2-b)}$$

$$I_{PID} = K_c * (1/T_i) * |E_{db}/2| * T_r \quad \text{(2-c)}$$

$$D_{PID} = K_c * (1/T_i) * |E_{db}/T_r| \quad \text{(2-d)}$$

$$R_r = |\Delta PV / \theta| \quad \text{(2-e)}$$

$$T_r = |E_{db}/R_r| \quad \text{(2-f)}$$

$$I_{PID} \le P_{PID} + D_{PID} \quad \text{(2-g)}$$

$$T_i \ge (T_r/2)/(1 + T_d/T_r) \quad \text{(2-h)}$$

DB = deadband for smart reset (%)
D_{PID} = derivative mode contribution to PID output (%)
E_{db} = error less deadband (%)
K_c = controller gain
K_i = integrating process gain (%/sec/%)

I_{PID} = integral mode contribution to PID output (%)
P_{PID} = proportional mode contribution to PID output (%)
PV = process variable (%)
R_r = ramp rate (%/sec)
SP = setpoint (%)
T_r = rise time (sec)
T_d = derivative time (rate time) (sec)
T_i = integral time (reset time) (sec)
ΔPV = delta PV in total loop deadtime for ramp rate calculation (%)
θ = total loop deadtime (sec)

Smart action can prevent overshoot or faltering of approach. If the PV predicted two deadtimes into the future is going to be within or exceed the deadband, the reset time is increased over the next deadtime. If the ramp rate is faltering or even reversing direction in the approach, the reset time is decreased.

An essential detail in the implementation of adaptive reset (and also Bang-Bang logic in the Output Tracking section) is the use of a deadtime block to compute a continuous train of ramp rates and future values every execution of the deadtime block with a good signal to noise ratio. The deadtime block creates an old PV one loop deadtime in the past. The old PV subtracted from the new PV is a delta PV that when divided by the deadtime is the ramp rate. The delta PV multiplied by the loop deadtime and added to the new PV is a predicted PV one deadtime into the future.

THRESHOLD SENSITIVITY

In wireless measurement devices the data is updated periodically or by exception. In exception reporting the information is refreshed only if an exception is generated when the change in data has exceeded a defined limit. The *threshold sensitivity* is set as large as possible to conserve battery life. In this non-continuous update scenario the PID output continues to ramp due to the integral term; when an update is received, the traditional PID considers the entire change to have occurred within the PID execution time interval and, if derivative mode is used, the rate of change of the measurement becomes the difference between the new and old measurement divided by the PID execution time interval. The result is a spike in the controller output.

A threshold sensitivity used in conjunction with the enhanced PID for wireless suspends control action until the valve moves automatically stopping limit cycles from backlash and stick-slip if the rule on the number of integrators is met. The threshold sensitivity and enhanced PID also suspends control action until an analyzer result is available. Large and variable update times from at-line and off-line analyzers will not cause instability.[9] In fact, the tuning of the enhanced PID becomes incredibly simple. The PID gain can be the inverse of the open loop gain (product of valve, process, and measurement gains). The PID reset time can be as small as the execution time of the PID. The threshold sensitivity; smallest change in PV that initiates an update of the PV for the PID; is set so the PID does not execute in response to noise. If the PID output is in a reasonable range, the threshold sensitivity is set to zero.

WIRELESS ENHANCEMENTS

The enhanced PID option for wireless enables the following features:

(1) Positive feedback implementation of reset with external-reset feedback (dynamic reset limit)
(2) Immediate response to a setpoint change or feedforward signal or mode change
(3) Suspension of integral action until there is a change in PV
(4) Integral action is the exponential response of the positive feedback filter to the change in controller output for the time interval since last update
(5) Derivative action is the PV or error change divided by the time interval since the last update multiplied by the gain and rate time

The enhanced PID will not become oscillatory if the wireless update time or analyzer cycle time increases. The enhanced PID will go into a holding mode for a loss of communication, stuck control valve, or failure of an electrode, analyzer, or sample system providing a smooth continuation upon restoration of communication and analyzer results.[9] For offline analyzers, late lab results do not cause a problem. For interacting loops and VPC, a threshold sensitivity setting enables the enhanced PID to eliminate reactions to insignificant changes that could cause fighting between loops. Table 2 shows the function and advantages of the enhanced PID and other PID features that make the simple addition of a VPC PID a quick and easy solution for small optimization problems.[10]

Table 2 Key PID Features for Valve Position Control[10]

Feature	Function	Advantage 1	Advantage 2
Direction Velocity Limits	Limit VPC Action Speed Based on Direction	Prevent Running Out of Valve	Minimize Disruption to Process
Dynamic Reset Limit	Limit VPC Action Speed to Process Response	Direction Velocity Limits	Prevent Burst of Oscillations
Adaptive Tuning	Automatically Identify and Schedule Tuning	Eliminate Manual Tuning	Compensation of Nonlinearity
Feedforward	Preemptively Set VPC Out for Upset	Prevent Running Out of Valve	Minimize Disruption
Enhanced PID	Suspend Integral Action until PV Update	Eliminate Limit Cycles from Stiction & Backlash	Minimize Oscillations from Interaction & Delay

CONCLUSION

PID features in the modern DCS enable the PID to achieve a higher level of performance and automation. The PID is not longer relegated to basic control. The new PID capability extends applications to small optimization problems and is able to handle more real world non-ideal behavior such as poor valve response, wireless update rates, plugged sample lines, broken electrodes, and missing analyzer results. This extended functionality can promote better and more accurate performance in large feeding systems, protect equipment from mechanical failure, and set the basis for smother startups and transitions. Tuning can focus on maximum disturbance rejection and still meet needs for coordinating loop responses, reducing interactions, and minimizing overshoot and faltering in setpoint responses.

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