

# **A UNIFIED PID METHODOLOGY TO MEET PLANT OBJECTIVES**

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## **Introduction**

Plant objectives for basic and advanced regulatory control can be quite diverse. PID control plays an important role in maximizing safety, environmental and equipment protection, process efficiency and process capacity. There can be many different sources of process variability that control loops need to deal with, such as raw material and recycle composition, weather, utility temperature and pressure, operators, interactions, optimization, on-off actions, startups, transitions, shutdowns, measurement and process noise, and limit cycles. Process nonlinearities, process non-self-regulation, dead time dominance, slow measurements and valves, analyzer cycles times, deadband, and threshold sensitivity or resolution limits can make the task much more difficult. The techniques for applying PID control to achieve plant objectives are largely individualistic than systematic. Each application tends to be treated as unique requiring special expertise and tuning. The result has been over 100 tuning rules to meet different perceived and actual needs of the process. Consultants are often required and PID features are not effectively utilized. In some plants PID loops have been pushed into model predictive control (MPC) because PID function application practices are largely heuristic and undocumented.

A unified approach to tuning has been found that enables a common and simplified method for setting PID tuning parameters. Key features can be used to eliminate the need for retuning to deal with different dynamics and objectives. This paper shows a methodology that integrates a unified tuning approach and key features that minimizes implementation and maintenance efforts. A step by step method will be used that will address the myriad of dynamics and objectives in PID applications.

## **Types of Process Responses**

There are 4 major types of process responses. Early tuning rules were developed with one type of process response in mind. Tuning rules developed in the 1990s have expanded the scope of processes addressed but few effectively deal with all types of responses. The tuning requirements are so different for each type of response that users can readily end up confused. The first step to resolving this deficiency is recognition of the distinctive characteristics of each type of process response.

## **Response Methodology and Terminology**

The effect of automation system dynamics will be included and distinguished in the process response. A first order plus dead time (FOPDT) approximation is used that consists of a dead time, a time constant, and a gain. For a near-integrating process the gain is divided by the time constant to give an integrating process gain used for integrating processes. The identification or approximation of a secondary time constant is added to improve the setting of the rate time for the derivative mode.

In the literature these parameters to define the dynamic response are referred to as process dead time, process time constants, and process gain. Since in industrial applications these parameters are determined by the automation system besides process dynamics, more general terms will be used.

The total loop dead time is the sum of all the pure time delays plus the fraction of the small time constants that is equivalent dead time. All time constants smaller than the largest (primary time constant) and next largest if a secondary time constant is identified create an effective dead time. The delays and time constants as shown in Figure 1 in the final control element (e.g. control valve or variable speed drive), process, measurement, and controller must be included.

In the literature primary and secondary time constants are considered to be in the process as shown in Figure 1. In industrial applications these time constant can be in the automation system. In this paper the largest time constant in Figure 1 is termed the open loop time constant.

The open loop gain is the product of the final control element, process, and measurement gains shown in Figure 1. For self-regulating processes, the open loop gain is dimensionless (%/%). For integrating processes, the resulting gain called an integrating process gain has units of 1/sec (%/sec/%). Processes with a large process time constant are termed near-integrating. The near-integrating process gain is attained by dividing a self-regulating process gain by the primary process time constant.

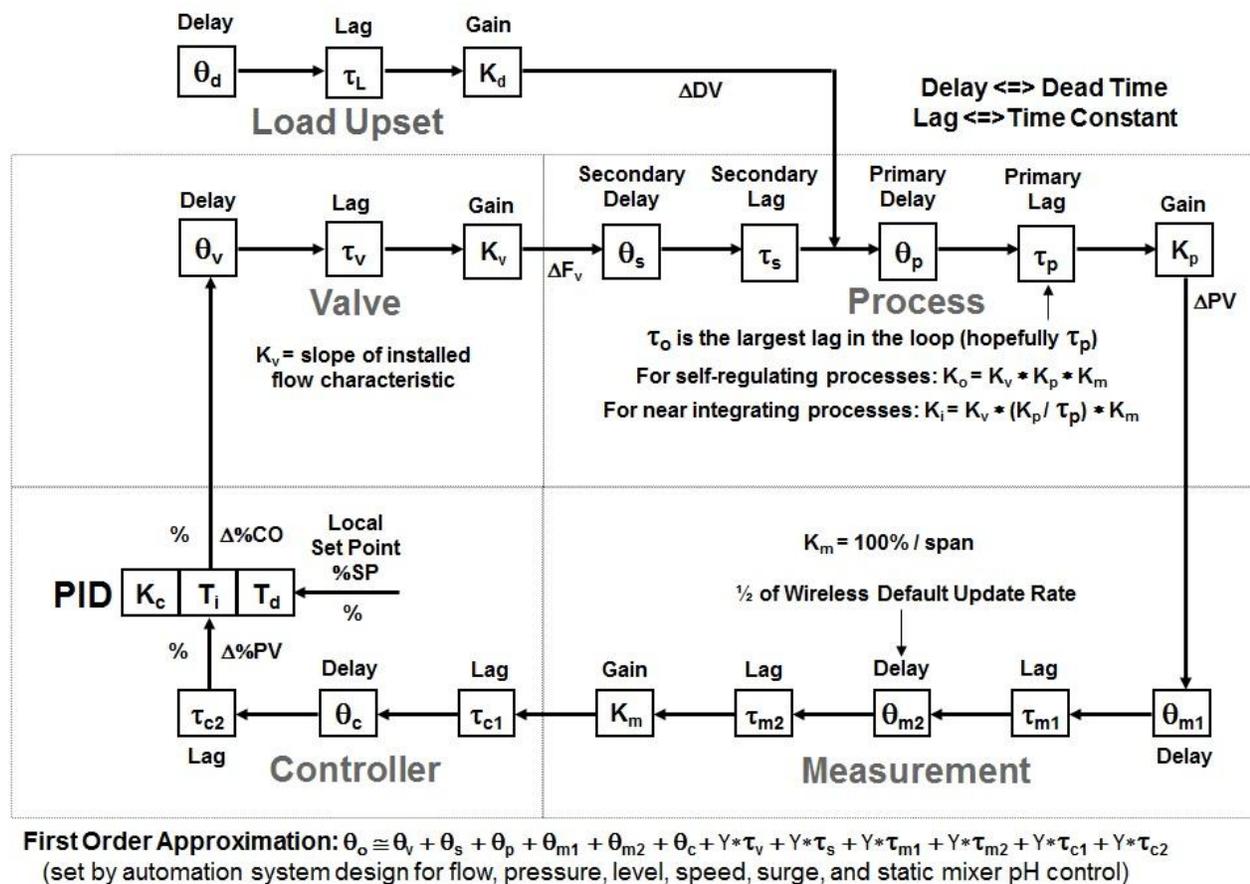


Figure 1 – Block diagram reveals the myriad sources of dynamics in the FOPDT response

## **Dead time Dominant Self-Regulating Responses**

A self-regulating process will reach a new steady state if a manual change in the PID output is made provided there are no disturbances. The steady state is the result of internal self-regulation from a negative feedback process time constant. The actual time to steady state observed will depend upon the open loop time constant (largest time constant in the loop), which may be in the automation system. If the total loop dead time exceeds the open loop time constant, the process is dead time dominant.

The process dynamic response of conveyors, sheets and spin lines is essentially a pure dead time from the transportation delay. However, if die bolt heaters are used instead of hydraulic actuators to control sheet thickness, the thermal lag introduced can be large enough that the process is no longer dead time dominant.

Blenders, desuperheaters, extruders, static mixers, and plug flow reactors have a negligible process time constant due to the lack of back mixing. However, the temperature response is often not dead time dominant due to the thermal lags in the equipment heat transfer surface and thermowell.

The addition of a large wireless default update rate in a flow loop can make the flow loop dead time dominant. A large wireless default update rate in a static mixed pH loop or the use of an at-line or off-line analyzer with a large cycle time can make these composition loops dead time dominant. For these cases where the dead time dominance is due to a discontinuous update of an automation system component, we will see that a PID feature can be enabled to prevent instability and simplify tuning.

## **Lag Dominant Self-Regulating Response**

If the self-regulating process response has an open loop time constant larger than the total loop dead time, the process is termed lag dominant. Continuous operations with mixing such as columns, crystallizers, evaporators, neutralizers, and reactors have a lag dominant response. Liquid flow has a self-regulating response but the process time constant is less than the total loop dead time. A signal filter used to reduce noise can make the process lag dominant. A large automation system lag can make a response lag dominant that would have been dead time dominant based on the process time constant.

If the open loop time constant is much larger than the loop dead time, the process response seen by the PID is a ramp in the control region of 4 dead times. A PID tuned for disturbance rejection will have done nearly all of its corrections in 4 dead times. So far as what the PID sees, the process response is integrating. Such processes are termed “near-integrating” and can use integrating process tuning rules. Temperature and composition control loops on continuous processes are near-integrating.

## **Integrating Response**

Processes that have no negative feedback process time constant and hence no steady state are termed integrating processes. The process response integrates any unbalance in the process. Since the unbalance is nearly never exactly zero, the process will always ramp when the PID is in manual. For most integrating processes the ramp rate is extremely slow (e.g. 0.000001 %/sec). The extreme exception is gas pressure where the ramp rate can be a million to a billion times faster. An electrical

furnace had an integrating process gain of 1 %/sec/%. A waste heat incinerator gas pressure had an integrating process gain of 1000 %/sec/% (too fast for any digital controller).

The most common example of an integrating process is level. Gas pressure has an integrating response when changes in operating pressure are a small fraction of the pressure drop across inlet or outlet valves and consequently do not appreciably affect the flow into or out of the system. Batch operations cause an integrating response because there is no discharge flow until the operation is complete. Batch temperature is integrating unless the heat loss changes appreciably with temperature. Batch composition and batch pH has an integrating response unless the reactant or reagent is rapidly consumed in a reaction.

## Runaway Response

Processes with a positive feedback time constant can accelerate. These processes are termed open loop unstable. If the PID is put in manual, the process can accelerate and run off-scale. Exothermic reactions where the reaction rate greatly increases with temperature can have runaway response if the heat release exceeds the cooling capability. A large secondary time constant (e.g. heat transfer surface or thermowell lags) or dead time approaching the positive feedback time constant can cause the process to be unstable regardless of PID tuning.

## Process Objectives

The main task of the PID is to determine the transfer of variability from a controlled variable such as temperature to the manipulated variable such as coolant flow. The process input most often manipulated is a process or utility flow. For maximum disturbance rejection, the transfer of variability is maximized. This is generally the case for vessel and column temperature, composition, gas pressure, and pH control. An insufficiently aggressive transfer of variability can cause product degradation or an emission, uncontrollable reactions, a relief valve or rupture disc to blow, and a safety instrumented system (SIS) to activate. Safety, equipment and environmental protection, on-stream time, and yield depend upon disturbance rejection.

To meet these objectives the integrated and peak error must be minimized. The ultimate limit to how small these errors are depend upon the total loop dead time ( $\theta_o$ ) and the open loop time constant ( $\tau_o$ ). Figure 2 shows that the ultimate limit to the peak error is proportional to the ratio of dead time to the 63% response time which is the dead time plus the open loop time constant. The peak error is the altitude of the right triangle whose base is the dead time since no correction can be enacted and seen till one dead time. A mirror image of this right triangle placed back to back with the first triangle shows the total recovery from the disturbance. The ultimate limit to the integrated error is the area of the two triangles which ends up being the dead time to 63% response time multiplied by the dead time.

In some applications the transfer must be minimized because changes in the manipulated flow affect other loops and operations. The most common application is surge tank level control where the level PID manipulates the discharge flow to the degree only needed to keep the tank level within low and high limits. The objective is to minimize the disruption of feed changes to downstream operations. This objective is also effective for a distillate receiver level PID manipulating distillate flow. This goal is often termed “maximization of the absorption of variability”.

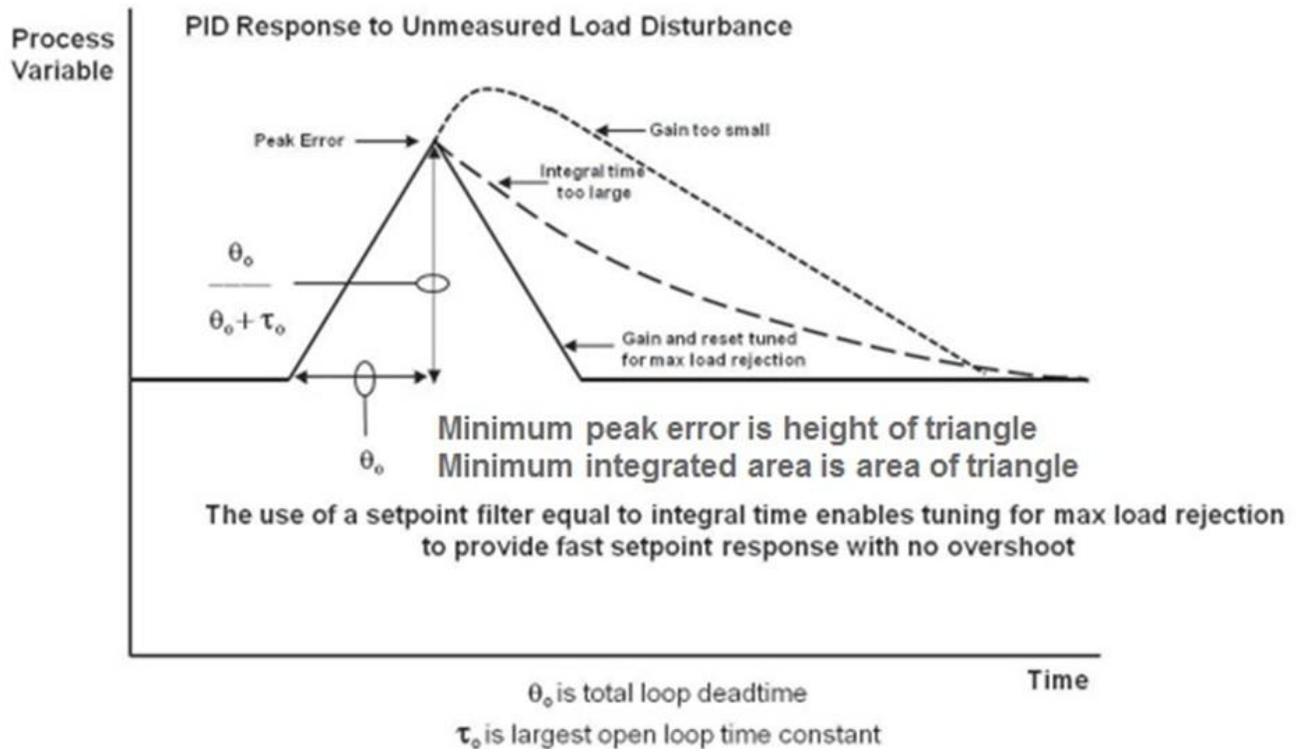


Figure 2 – The ultimate limit to the peak and integrated errors depends upon the total loop dead time and open loop time constant

Aggressive control action can also upset other loops. For interactions that cannot be reduced sufficiently by decoupling techniques, the transfer of variability for the least important loops is minimized to avoid upsetting the more important loops.

When loops are manipulating flows affecting a unit operation, the transfer of variability to these flows is coordinated so the stoichiometric ratio is not momentarily upset from one flow being faster or slower than another flow. In this case the transfer of variability in all loops is regulated to match the slowest loop. This coordination of manipulated flows is particularly important for operations without back mixing to smooth out transients. Coordination is consequently critical for blending and plug flow liquid and gas reactions (e.g. catalyst bed gas reactors).

PID loops must not increase variability. A PID loop can oscillate from overly aggressive action and interactions and in integrating and runaway processes from inadequate immediate action due to a gain setting that is too low. A PID can create variability when there are automation system difficulties.

### Automation System Difficulties

The response of the automation system can introduce excessive discontinuities, lag, noise, and dead time. Here we look at the components of the system for the sources of these difficulties. There is only room in this paper to note some of the more significant deficiencies.

## **Final Control Element**

A control valve can have significant stiction from excessive friction often described as a resolution limit. Actuators and positioners can also have a threshold sensitivity limit. One or more integrators in the process or PID in a loop with a resolution or threshold sensitivity limit will cause a limit cycle that cannot be eliminated by tuning. A limit cycle is a sustained equal amplitude oscillation.

A control valve or damper can have significant backlash from play in linkages and connections seen as a dead band upon reversal of direction of the signal. Two or more integrators in the loop will cause a limit cycle than cannot be eliminated by tuning. Thus, an integrating process and a PID controller with integral action can cause a limit cycle. A self-regulating process with a cascade loop where the primary and secondary PID each have integral action will exhibit a limit cycle.

A variable speed drive (VSD) with a resolution limit due to insufficient number of bits in the analog to digital signal input card will result in a limit cycle for one or more integrators. If a dead band is introduced in the VSD setup, limit cycle will develop for two or more integrators.

A loop with a slow final control element that cannot keep up with the rates of change of the controller output will burst into oscillations. The oscillations may not show up during loop tests where typically small changes are made. The burst of oscillations may only occur for large fast disturbances and large setpoint changes causing the rate of change of the PID output to be faster than the valve slewing rate or VSD rate limited response. Large valves and dampers and VSD with marginally sized motors can be too slow.

## **Measurement**

The sensor design, location, and operating conditions can result in excessive measurement noise. Differential head meters and vortex meters are especially sensitive to velocity profile and hence the upstream and downstream piping configuration. Most flow meters get noisy as you approach the low flow limit. pH electrodes are particularly vulnerable to noise from electromagnetic interference and concentration fluctuations.

The measurement can also introduce an excessive lag into the control loop. Fouled electrodes and thermowells can have an excessive lag from a decrease in the mass transfer and heat transfer coefficients, respectively. Pressure and differential pressure transmitters on compressor control can become too slow by simply increasing the transmitter damping setting.

If the measurement lag becomes by far the largest time constant in the loop an insidious situation develops where the trend charts may look smoother because the measurement is showing an extremely attenuated (filtered) view of the actual process oscillations. The controller gain may even be able to be increased further misleading the user into thinking the measurement lag is beneficial. The key is the loop period should show an increase due to the increase in loop dead time.

Wireless measurements can introduce excessive dead time in fast loops by a slow default update rate. At-line analyzers and especially off-line analyzers can cause excessive dead time in almost any loop from the sample transport and processing, analysis cycle time, and multiplex time.

## Controller

The controller can introduce an excessive dead time from too large of a module execution time or signal filter time. Even without considering the effect of the increase in dead time on the ultimate limit to errors and tuning, a signal filter and module execution time larger than 10% of the reset time will increase the integrated errors by more than 10% for an unmeasured disturbance.

Too slow of a secondary controller tuning can cause a cascade loop to burst into oscillations when the primary controller output is changing faster the secondary loop can respond. Insufficient gain action can cause slow rolling oscillations in an integrating process and instability in a runaway process.

## Unified Tuning

The key breakthrough in thinking is to use the lambda tuning rules for integrating processes for lag dominant self-regulating and runaway processes. The lambda tuning rules automatically prevent the product of the PID gain and reset time from being less than the twice the inverse of the integrating process gain that would result in slow rolling oscillations. The main decision is whether to maximize the transfer of variability or maximize the absorption of the variability. Key features are used rather than tuning to address the automation system difficulties and to meet different process objectives.

### Maximization of the Transfer of Variability

The maximum dead time for all possible operating conditions is used as the arrest time, which is the time to start to turn back an excursion from an unmeasured disturbance. The arrest time is lambda. This minimum lambda is the dead time as shown in Figure 2. The minimum lambda plus the maximum integrating process gain (fastest ramp rate) is used in the tuning equations. The lambda tuning rule then gives the same gain and reset time as the Ziegler Nichols Reaction Curve method. For temperature, composition, and pH loops, the rate time should be set equal to the secondary time constant. If this time constant cannot be identified, 1/2 of the total loop dead time can be used as an approximation. This gives a rate time that is about 1/6 of the reset time. Note that for the ISA Standard Form PID, the rate time must always be less than the reset time to avoid instability (see Appendix A).

If a lower controller gain is used. The equivalent lambda needs to be back calculated and used in the calculation of the reset time to prevent slow rolling oscillations. For runaway processes, the PID gain must always be greater than the inverse of the process gain to prevent a runaway.

### Maximization of the Absorption of Variability

When the absorption of variability must be maximized, lambda is chosen to be as large as possible. The maximum arrest time lambda depends upon the integrating process gain, the allowable change in the process variable and the available change in the manipulated flow. The following equations show how to calculate lambda for a generic application and then a surge tank level loop.

The maximum lambda ( $\lambda$ ) is the maximum allowable % excursion ( $\Delta\%PV_{\max}$ ) divided by the maximum possible PV ramp rate. The maximum possible ramp rate is the PV rate of change per percent output change multiplied by the maximum available percent output change ( $\Delta\%CO_{\max}$ ).

$$\lambda = \frac{|\Delta\% PV_{\max}|}{\left| \frac{\frac{\Delta\% PV}{\Delta t}}{\Delta\% CO} \right| * |\Delta\% CO_{\max}|} \quad (1)$$

Realizing that the integrating process gain is the PV ramp rate per percent output change:

$$\lambda = \frac{1}{K_i} * \frac{|\Delta\% PV_{\max}|}{|\Delta\% CO_{\max}|} \quad (2)$$

An equivalent setpoint rate limit on the controller output (e.g. flow controller setpoint):

$$\left| \frac{\Delta\% CO}{\Delta t} \right|_{\max} = \frac{|\Delta\% CO_{\max}|}{\lambda} = K_i * \frac{|\Delta\% CO_{\max}|^2}{|\Delta\% PV_{\max}|} \quad (3)$$

For a PV limit ( $\%PV_{Limit}$ ) and corresponding CO limit ( $\%CO_{Limit}$ ) we have:

$$\lambda = \frac{1}{K_i} * \frac{|\%PV_{Limit} - \%SP|}{|\%CO_{Limit} - \%CO|} \quad (4)$$

The above calculation would be done for high and low operating limits and various setpoints. The smallest of the arrest times would be used in tuning.

We can obtain the more detailed requirements for surge level tank level control by computing the integrating process gain for level. The integrating process gain is the product of the valve, process, and measurement gains:

$$K_i = K_v * K_p * K_m \quad (5)$$

The valve gain or variable speed drive gain for a linear installed characteristic or flow loop is:

$$K_v = \frac{\Delta F_v}{\Delta\% CO} = \frac{\Delta F_{\max}}{100\%} \quad (6)$$

The level process gain for mass flow is (omit density term for volumetric flow):

$$K_p = \frac{1}{A * \rho} \quad (7)$$

The level measurement gain is:

$$K_m = \frac{\Delta\% PV}{\Delta L} = \frac{100\%}{\Delta L_{\max}} \quad (8)$$

Substituting in the valve, process, and measurement gains, the integrating process gain is:

$$K_i = K_v * K_p * K_m = \frac{\Delta F_{\max}}{100\%} * \frac{1}{A * \rho} * \frac{100\%}{\Delta L_{\max}} = \frac{\Delta F_{\max}}{\Delta L_{\max}} * \frac{1}{A * \rho} \quad (9)$$

The consequential arrest time for a level loop is:

$$\lambda = \frac{A * \rho * \Delta L_{\max}}{\Delta F_{\max}} * \frac{\Delta\% PV_{\max}}{\Delta\% CO_{\max}} \quad (10)$$

An equivalent setpoint rate limit on the controller output (e.g. flow controller setpoint):

$$\left| \frac{\Delta\% CO}{\Delta t} \right|_{\max} = \frac{|\Delta\% CO_{\max}|}{\lambda} = \frac{\Delta F_{\max}}{A * \rho * \Delta L_{\max}} * \frac{|\Delta\% CO_{\max}|^2}{|\Delta\% PV_{\max}|} \quad (11)$$

Where:

$A$  = cross sectional area (m<sup>2</sup>)

$\Delta F_{\max}$  = maximum change in valve or variable speed drive flow (e.g. flow span) (kg/sec)

$\Delta L_{\max}$  = maximum change in level (e.g. level span) (m)

$K_i$  = integrating process gain (1/sec)

$K_m$  = measurement gain (%/m for level)

$K_p$  = process gain (m/kg for level)

$K_v$  = valve or variable speed drive gain (kg/sec/%)

$\lambda$  = arrest time (sec)

$\rho$  = fluid density (kg/m<sup>3</sup>)

% $CO$  = operating point controller output (%)

% $CO_{Limit}$  = controller output limit (%)

$\Delta\% CO_{\max}$  = maximum available change in controller output (%)

% $PV_{Limit}$  = process variable limit (%)

$\Delta\% PV_{\max}$  = maximum allowable change in process variable (%)

% $SP$  = operating point setpoint (%)

## Key PID Features

There are seven key related PID control features; structure, external-reset feedback, feedforward, setpoint filter and lead-lag, setpoint rate limits, output tracking, and an enhancement for wireless. These key features synergistically provide capabilities that have not been fully realized. See Appendix A for conversion of tuning settings for different PID Forms and tuning setting units.

### PID Structure

Most modern DCS offer a choice of structures shown in Table 1.

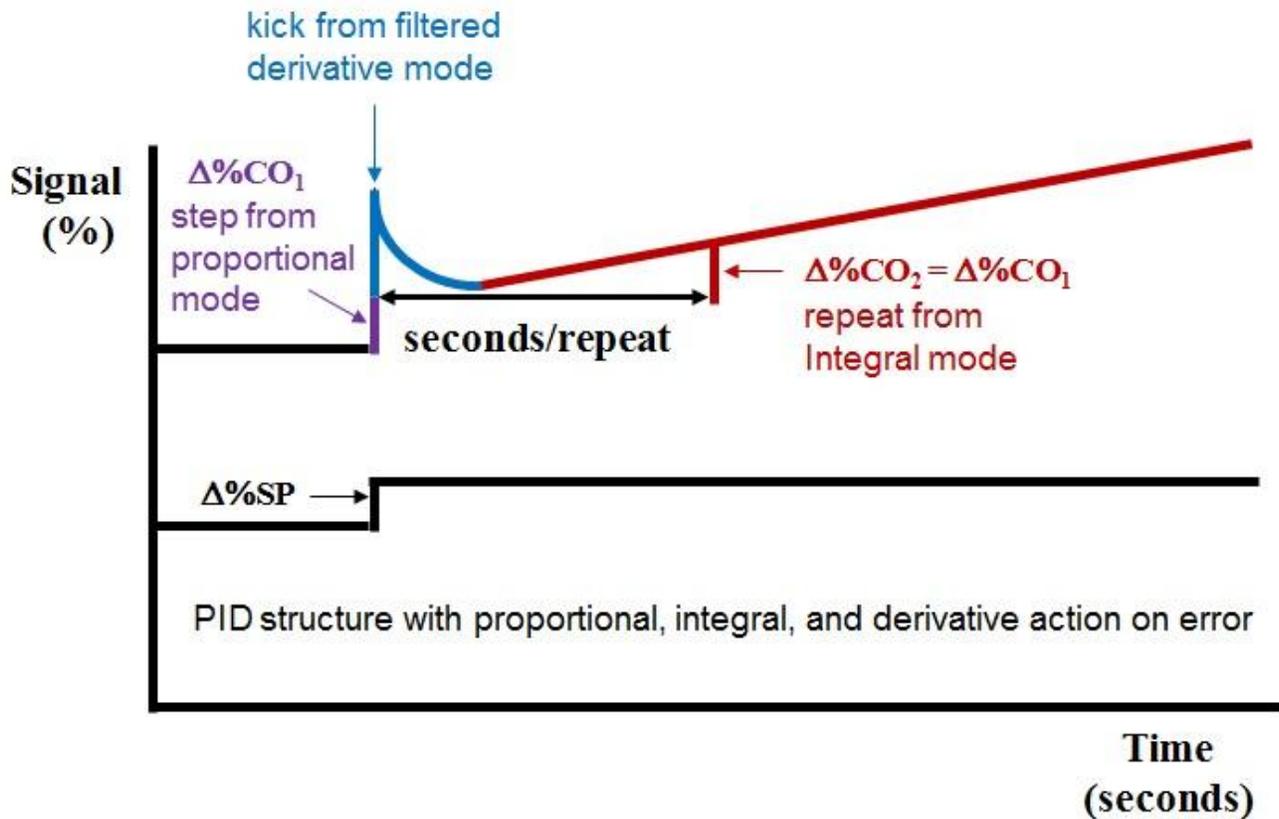
Table 1 - Major PID Structure Choices

- (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 ( $\beta$  and  $\gamma$  adjustable 0 to 1)

The structure choices use  $\beta$  and  $\gamma$  set point weighting factors. A controller that has both factors adjustable is called a “two degree freedom controller.” Other structures have the  $\beta$  and  $\gamma$  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.

In choosing the correct structure it is important to know the relative contribution of the proportional, integral, and derivative modes. Figure 3 shows the contribution of each PID mode for a setpoint change. There is no feedback from the process in this figure (e.g. block valve closed), so we can more clearly see the contribution of each mode.

Structure 1 (PID on error) provides the fastest approach to a new setpoint by virtue of a step and a kick in the controller output from the setpoint change as seen in the second response in Figure 3. For small setpoint changes and low controller gains, the step and kick can help get through significant valve backlash and stick-slip to get the valve moving. The kick 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 step and kick is more of a psychological than a process concern.



**Contribution of Each PID Mode for a Step Change in the Set Point  
Structure of PID on error ( $\beta=1$  and  $\gamma=1$ )**

Figure 3 – Contribution of each PID mode for a step change in setpoint for a PID on error structure

Structure 2 (PI on error, D on PV) is the structure most often used. Structure 2 eliminates the kick from derivative action for a setpoint change by setting gamma to zero ( $\gamma=0$ ). 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 for a high controller gain. Increases in process gain or dead time will increase the overshoot unless the controller gain is decreased accordingly. If the elimination of setpoint overshoot is much more important than rise time (e.g. bioreactor temperature and pH shifts), then structure 3 may be best.

Structure 3 (I on error, PD on PV) eliminates overshoot but with quite a sacrifice in speed of approach to the setpoint. For near and true integrating and runaway processes, structure 3 may have a rise time an order of magnitude larger than the rise time for structure 1 for the same tuning settings and process dynamics. A setpoint filter set equal to the reset time provides the same response.

Structure 4 (PD on error, no I) 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 also 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 on error, D on PV, no I) is used for many of the same reasons as structure 4.

Structure 6 (ID on error, no P) 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 Figure 4 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.

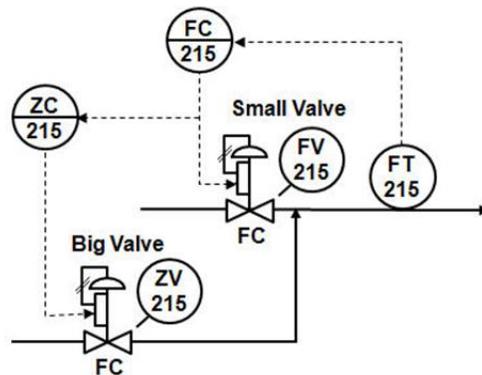


Figure 4 - Valve position control increases precision and rangeability

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. Tuning rules often cited suggest 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. In addition external-reset feedback, setpoint directional rate limits, and an enhanced PID developed for wireless can be used as a flexible and easy to tune solution to deal with unknowns. This solution is described in the Setpoint Rate Limits section.

Structure 7 (I on error, D on PV, no P) 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 (two degrees of freedom via adjustable  $\beta$  and  $\gamma$ ) is used to provide a balance between a fast rise time and overshoot. The same effect can be accomplished by the use of a setpoint lead-lag as discussed in the setpoint filter section. This structure can be used to get the structures with proportional or derivative action on PV by zeroing the respective  $\beta$  and  $\gamma$  set point factors.

### External-Reset Feedback

PID controllers with the positive feedback implementation of integral action as shown in Figure 5 enable the use of an external reset feedback. When the switch in Figure 5 is in the position to get an external reset feedback to the integral mode from the process variable (PV) of a setpoint being manipulated whether an Analog Output block or a secondary PID, the PID output is prevented from changing faster than the PV of the control valve, VSD, or secondary loop can respond. There is an increasingly larger scope of potential benefits discovered as described in subsequent features and the summary of the unified methodology. The only requirement is that the external reset feedback be properly connected to a fast update of the PV for the valve, VSD, or loop being manipulated. Updates of actual valve position, VSD speed, or secondary loop PV should be at least twice as fast as the the execution time of the controller with external reset feedback.

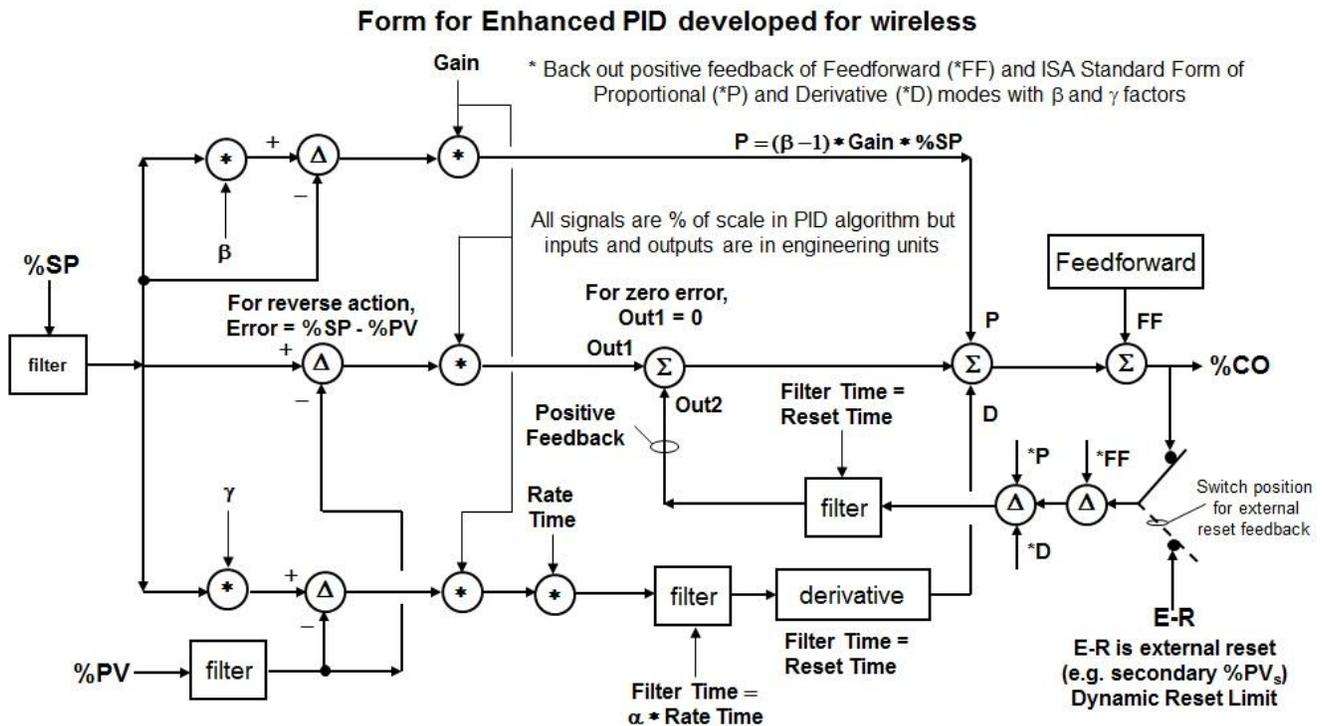


Figure 5 – ISA Standard Form PID with External Reset Feedback

## **Feedforward**

Feedforward control is particularly effective when a disturbance is large, fast, and precisely measured. While theoretically a feedforward multiplier is more effective for changes in the slope of a plot of the manipulated flow versus the measured disturbance, for many practical reasons such as scaling and nonlinearity problems, a feedforward summer is used for feedforward control especially on vessels and columns. The most common feedforward signal is flow because the changes in feed flows are faster than changes in composition and temperature and the flow is the most common process input. A flow feedforward summer is implemented via an external ratio and bias station so the operator can set the desired ratio and see the actual ratio. On startup, the operator may run on ratio control with the process controller in manual until the equipment reaches operating conditions. The initial flow ratios can be set based on the process flow diagram (PFD). The PFD can be put live online on the operator graphics to show the actual ratios used. The technique also enables a plant wide feedforward system to rapidly increase and decrease production rates.

## **Setpoint Filter and Lead-Lag**

A setpoint filter can provide the desired closed loop time constant important for the coordination of loops. The use of secondary PID flow setpoint enables stoichiometric ratios to be enforced without a transient unbalance for production rate changes and setpoint changes and disturbances to the primary PID. External reset feedback must be turned on in the primary PID to prevent it from trying to change the secondary PID set point faster than filter will allow the secondary loop to respond. Turning external reset on eliminates the need to retune the primary PID when the secondary PID filter is added or tuned. If coordination of loops is not necessary, a setpoint filter should not be added to a secondary loop because it will slow down the rejection of primary loop disturbances.

The rise time (time to reach setpoint) and overshoot for a setpoint change can be optimized by the use of a setpoint lead-lag. In this case the setpoint lag time is set equal to the reset time and the setpoint lead time is set equal to about 20% of the lag time. If preventing overshoot is more important than reducing rise time for a setpoint change, the lead time can be decreased.

## **Setpoint Rate Limit**

The setpoint rate limit option in analog output (AO) and PID function blocks in a modern DCS can provide move suppression, a tuning parameter found to be very useful in model predictive control. Since there is often an up and down rate limit, directional move suppression can be used provided the ability needed to enable a slow approach to an optimum and a fast getaway to prevent an abnormal condition. This technique is used to provide a fast opening and slow closing anti-surge control valve. There are many other applications as we will see in the unified methodology. If external reset feedback is turned on in the PID manipulating the setpoint, the PID does not have to be retuned when the rate limits are added or tuned in the manipulated AO or PID. These rate limits can be written to online.

## **Output Tracking**

Disturbances can be too fast and the consequences too severe to rely just on PID control. Here output tracking is used to get the loop out of trouble for equipment and environmental protection. The most common example is compressor surge prevention, where the PID output will track a surge valve

position large enough to prevent surge when a crossing of the surge curve is detected or a precipitous drop in compressor suction flow occurs indicating the start of surge. The PID stays in output tracking holding the protective output for several seconds to ensure the system has stabilized.

The fastest possible setpoint response is obtained from a full throttle approach (bang-bang control method) where the PID is put in output tracking at the output limit to drive the PV to the setpoint. The PID output stays at the limit until the PV one dead time into the future is predicted to reach setpoint. A dead time (DT) block with its parameter set equal to the loop dead time is used to create an old PV that is subtracted from the new PV to create a delta PV. The delta PV added to the new PV that is the DT block input gives the PV one dead time into the future. When the future PV is projected to reach setpoint, the PID tracks a final resting value (FRV) captured from previous batches or startups and holds this FRV for one dead time before releasing the PID back to feedback control.

### **Enhanced PID for Wireless**

The enhanced PID option developed for wireless enables the following features:

1. Positive feedback implementation of reset with external-reset feedback
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 depending upon PID structure divided by the time interval since the last update (instead of PID execution time) multiplied by the gain and rate time

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. For offline analyzers, late lab results do not cause a problem. The enhanced PID eliminates the need for retuning as discontinuous delays from wireless devices and analyzers are increased. For interacting loops and valve position control and final control elements with resolution limits and deadband, a threshold sensitivity setting enables the enhanced PID to eliminate reactions to insignificant changes that could cause fighting between loops or perpetuate a limit cycle. The threshold sensitivity is set just larger than the PV noise amplitude after filtering.

## **A Unified Methodology**

The following is a generalized step by step procedure to address a wide spectrum of automation system difficulties and application objectives. Note that PID features vary considerably from one DCS manufacturer to another. Any setting of options or tuning must be based on information from the supplier and should be thoroughly tested before and carefully observed after commissioning.

1. Set the output limits to keep the manipulated setpoints in the desired operating range. For variable speed drives set the process PID low output limit so the speed cannot cause the discharge head to approach the static head in order to prevent excessive sensitivity to pressure and to prevent reverse flow. In general, set the anti-reset windup limit to match the output limit. If the output scale is engineering units, the output limits and anti-reset windup must be based on the

output scale range and units.

2. Choose the best structure for your application. Generally the best choice is structure 2 with PI on error and D on PV. For a single direction response (e.g. batch heating or neutralization), use structure 4 or structure 5 so that there is no integral action. For a highly exothermic reaction, you might want structure 5 to help prevent a runaway from integral action.

3. Set the signal filter noise just large enough to keep the controller output fluctuations from exceeding the resolution limit or deadband of the final control element.

4. For near-integrating, true integrating, and runaway processes use the lambda integrating process tuning rules. To maximize the transfer of variability from the process variable to the manipulated variable, set the lambda (arrest time) equal to twice the maximum possible dead time and use the largest integrating process gain for all operating conditions in the tuning. To maximize the absorption of variability (e.g. surge tank level) use the minimum arrest time computed from Equations 1 through 10 for all possible operating conditions. If you decrease the PID gain, proportionally increase the PID reset time to prevent slow rolling oscillations. For runaway processes, the PID gain must not be less than the inverse of the open loop gain. When changing from the Series Form to the ISA Standard Form in newer DCS, convert the tuning settings based on the differences in Form. In the ISA Standard Form the rate time should not be greater than  $\frac{1}{4}$  the reset time. Be sure to realize and convert any difference in Form and tuning setting units between different PID vintages and suppliers (see Appendix A). You must take into account the units and inverse relationship between gain and proportional band and reset time in seconds and minutes per repeat. Not addressing these differences can result in settings off by orders of magnitude.

5. For self-regulating processes with the open loop time constant less than 4 times the dead time, use the lambda self-regulating tuning rules. To maximize the transfer of variability from the process variable to the manipulated variable set the lambda (closed loop time constant) equal to the twice the maximum possible dead time and use the largest process gain and smallest time constant for all operating conditions in the tuning. Schedule tuning settings preferably with an adaptive tuner with the specified lambda to deal with changes in the dynamics.

6. Turn on external reset feedback. Make sure the external reset feedback signal is correctly propagated back to the PID (e.g. BKCAL signal) especially if there are split range, signal characterizer, or signal selector blocks on the PID output.

7. For final control elements that are slow or that have deadband or resolution limit, use a fast readback of the valve position or variable frequency drive speed as the external reset feedback to prevent a burst of oscillations from the PID output changing faster than the final control element can respond.

8. For final control elements that create limit cycles from resolution limits and deadband, use a fast readback of the valve position or variable frequency drive speed to stop the limit cycles.

9. For cascade control, use the PV of the secondary loop as the external reset feedback to prevent a burst of oscillations from violation of the cascade rule where the secondary loop must be significantly faster than the primary loop.

10. For setpoint filters of secondary loops for coordination of flow loops, use the PV of the secondary loop as the external reset feedback to prevent the need to retune the PID.

11. For setpoint rate limits use the PV of the analog output block or secondary loop as the external reset feedback to prevent the need to retune the PID. Add setpoint rate limits to minimize the interaction between loops and in valve position control and to provide directional move suppression to enable a fast getaway for abnormal conditions and a slow approach to optimum. For valve position control, use an enhanced PID developed for wireless with a threshold sensitivity limit to ignore insignificant changes in the valve position to be optimized.

12. Add output tracking for equipment protection and a full throttle (bang-bang control) strategy for the fastest possible time to reach setpoint on startup and for batch operations.

13. Add output tracking logic to momentarily track an output that insures equipment and environmental protection. For compressor surge protection track a sufficiently large opening of the surge valves. To prevent a RCRA pH violation, track a rapidly incrementing reagent valve position to prevent an effluent excursion below 2 pH or above 12 pH.

14. Add feedforward control for large and fast measured disturbances. For flow feedforward, use a ratio and bias station so the operator can enter a desired flow ratio and see the actual flow ratio. Setup the PID to provide a bias correction to the manipulated flow. Add dynamic compensation (dead time and lead-lag blocks) to the feedforward so the manipulated flow arrives at the same point in the process at the same time as the measured disturbance.

15. For wireless devices or analyzers that introduce a large discontinuous PV update delay use an enhanced PID to eliminate the need to retune the controller to prevent oscillations. If the delay is much larger than the 63% process response time, the PID gain can be set as large as the inverse of the open loop gain for self-regulating processes.

16. For valve position control (VPC), make sure the key PID features in Table 3 are used. For examples of VPC for process optimization, see Table 4. The use of an enhanced PID enables the use of integral action in the VPC and simplifies the tuning of the VPC. While model predictive control can provide a more sophisticated optimization, the VPC can be implemented by a simple configuration change that involves setting up a PID as the VPC. The minimal implementation cost and time is attractive for small optimizations.

Table 3 – Key Feature Function and Advantages for Valve Position Control

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

Table 4 – Examples of Valve Position Control for Process Optimization

Optimization	VPC PID PV	VPC PID SP	VPC PID Out
Minimize Prime Mover Energy	Reactor Feed Flow PID Out	Max Throttle Position	Compressor or Pump Pressure SP
Minimize Boiler Fuel Cost	Steam Flow PID Out	Max Throttle Position	Boiler Pressure SP
Minimize Boiler Fuel Cost	Equipment Temperature PID Out	Max Throttle Position	Boiler Pressure SP
Minimize Chiller or CTW Energy	Equipment Temperature PID Out	Max Throttle Position	Chiller or CTW Temperature SP
Minimize Purchased Reagent or Fuel Cost	Purchased Reagent or Fuel Flow PID Out	Min Throttle Position	Waste Reagent Or Fuel Flow SP
Minimize Total Reagent Use	Final Neutralization Stage pH PID Out	Min Throttle Position	First Neutralization Stage pH PID SP
Maximize Reactor Production Rate	Reactor or Condenser Temperature PID Out	Max Throttle Position	Feed Flow or Reaction Temperature SP
Maximize Reactor Production Rate	Reactor Vent Pressure PID Out	Max Throttle Position	Feed Flow or Reaction Temperature SP
Maximize Column Production Rate	Reboiler or Condenser Flow PID Out	Max Throttle Position	Feed Flow or Column Pressure SP
Maximize Ratio or Feedforward Accuracy	Process Feedback Correction PID Out	50% (Zero Correction)	Flow Ratio or Feedforward Gain

Not mentioned is the benefit of using cascade control. Fast secondary loops can correct for many disturbances before they affect the primary loop. Secondary loops also reduce nonlinearities seen by the primary loop. A secondary flow loop isolates the nonlinearity of the installed characteristic of the control valve and enables flow feedforward. A secondary jacket or coil temperature loop isolates a crystallizer, evaporator, or reactor temperature loop from cooling or heating system nonlinearities.

## Conclusion

The use of key PID features, a unified PID tuning method, and an adaptive tuner enables an effective application of PID control for wide spectrum of process responses. Drastically different sources of variability, automation system difficulties and changing process objectives do not require PID retuning. PID features can be used to tailor the PID response to different application requirements on a much more understandable basis by the user. The role of the PID can be expanded to include process optimization besides regulation.

## Resources

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## Appendix A – PID Controller Forms

### Conversion of Settings to Correct Units and Form

The methodology in this paper is based on the ISA Standard Form with specific tuning setting units where the controller gain is dimensionless, the reset time is seconds (seconds per repeat), and the rate time is seconds. PID tuning settings must first be checked for units and if necessary undergo a units conversion before being used.

- (1) Convert proportional band (%) to controller gain (%/%) (dimensionless)

$$\text{Gain} = 100 \% / \text{Proportional Band}$$

- (2) Convert reset setting in repeats per minute to reset time in seconds per repeat

$$\text{Seconds per repeat} = 60 / \text{repeats per minute}$$

- (3) Convert rate time in minutes to rate time in seconds

$$\text{Seconds} = 60 * \text{minutes}$$

- (4) After the tuning setting units are verified, convert the Series or Parallel Form tuning settings to the ISA Standard Form tuning settings per equations below. The primed tuning settings are for the Series Form, the double primed tuning settings are for the Parallel Form, and the unprimed tuning settings are for the ISA Standard Form.

To convert from Series to ISA Standard Form controller gain:

$$K_c = \frac{T_i' + T_d'}{T_i'} * K_c' \quad (\text{A.1})$$

To convert from Series to ISA Standard Form reset (integral) time:

$$T_i = \frac{T_i' + T_d'}{T_i'} * T_i' = T_i' + T_d' \quad (\text{A.2})$$

To convert from Series to ISA Standard Form rate time:

$$T_d = \frac{T_i'}{T_i' + T_d'} * T_d' \quad (\text{A.3})$$

Note that if the rate time is zero, the ISA Standard and Series Form settings are identical. When using the ISA Standard Form, if the rate time is greater than ¼ the reset time the response can become

oscillatory. If the rate time exceeds the reset time, the response can become unstable from a reversal of action from these modes. The Series Form inherently prevents this instability by increasing the effective reset time as the rate time is increased.

We can convert from ISA Standard Form to the Series Form using the following equations if the reset time is equal to or greater than 4 times the rate time ( $T_i \geq 4 * T_d$ ).

$$K_c' = \frac{K_c}{2} * \left[ 1 + \left( 1 - 4 * \frac{T_d}{T_i} \right)^{0.5} \right] \quad (\text{A.4})$$

$$T_i' = \frac{T_i}{2} * \left[ 1 + \left( 1 - 4 * \frac{T_d}{T_i} \right)^{0.5} \right] \quad (\text{A.5})$$

$$T_d' = \frac{T_i}{2} * \left[ 1 - \left( 1 - 4 * \frac{T_d}{T_i} \right)^{0.5} \right] \quad (\text{A.6})$$

The Parallel form is used in some older DCS and PLC to isolate the proportional mode tuning setting from the other the modes. The gain setting does not affect the contribution from the integral and derivative modes and is sometimes called non-interacting. We can convert from the Parallel to the ISA Standard Form using the following equations.

$$K_c = K_c'' \quad (\text{A.7})$$

$$T_i = K_c'' * T_i'' \quad (\text{A.8})$$

$$T_d = \frac{T_d''}{K_c''} \quad (\text{A.9})$$

We can convert from the ISA Standard Form to the Parallel Form by the following equations:

$$K_c'' = K_c \quad (\text{A.10})$$

$$T_i'' = \frac{T_i}{K_c} \quad (\text{A.11})$$

$$T_d'' = K_c * T_d \quad (\text{A.12})$$

Where:

$K_c$  = controller gain for ISA Standard Form (%/%) (dimensionless)

$K_c'$  = controller gain for Series Form (%/%) (dimensionless)

$K_c''$  = controller gain for Parallel Form (%/%) (dimensionless)

$T_i$  = integral time (reset time) for ISA Standard Form (seconds)

$T_i'$  = integral time (reset time) for Series Form (seconds)

$T_i''$  = integral time (reset time) for Parallel Form (seconds)

$T_d$  = derivative time (rate time) for ISA Standard Form (seconds)

$T_d'$  = derivative time (rate time) for Series Form (seconds)

$T_d''$  = derivative time (rate time) for Parallel Form (seconds)

### Positive Feedback Implementation of Integral Mode

The positive feedback implementation of the integral mode effectively yields the equations above as seen in literature for a PI controller using Laplace transforms. Instead of an integrator, a filter whose input is the controller output and whose output is added to the contribution from the proportional mode in the positive feedback implementation. The filter time is the integral time setting. When external reset feedback is enabled, the input to the filter is switched from the controller output to the external reset feedback signal.

$$O(s) = K_c * E(s) + \frac{1}{1 + T_i * s} * O(s) \quad (\text{A.13})$$

$$O(s) * \left( 1 - \frac{1}{1 + T_i * s} \right) = K_c * E(s) \quad (\text{A.14})$$

$$O(s) * \left( \frac{T_i * s}{1 + T_i * s} \right) = K_c * E(s) \quad (\text{A.15})$$

$$\frac{O(s)}{E(s)} = K_c * \left( \frac{1 + T_i * s}{T_i * s} \right) = K_c + \frac{K_c}{T_i * s} \quad (\text{A.16})$$

Where:

$E(s)$  = Laplace transform of controller error (%)

$O(s)$  = Laplace transform of controller output (%)