Chapter 14
Industrial Applications of PID Control

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Nomenclature

a	PID process variable amplitude for relay method of auto tuning (%)
B_v	valve backlash (deadband) (%)
d	PID output step size for relay method of auto tuning (%)
n	PID process variable noise band for relay method of auto tuning (%)
A_o	amplitude of limit cycle (%)
CO_{t1}	controller output at time $t1$ before correction for load disturbance (%)
CO_{t2}	controller output at time $t2$ after correction for load disturbance (%)
E_i	integrated error (% s)
E_L	open loop error corrected for load disturbance time constant (%)
E_o	open loop error for unmeasured step disturbance (%)
E_x	peak error (%)
K_c	PID gain (dimensionless)
K_i	integrating process gain (% per second per %)
K_o	open loop gain for self-regulating processes (dimensionless)
K_u	ultimate gain (dimensionless)
K_v	valve gain (Flow e.u./CO %)
K_p	process gain (PV e.u./Flow e.u.)
K_m	measurement gain (PV %/PV e.u.)
N_m	measurement noise (%)
$PV_{100\%}$	process variable at 100% of measurement scale (PV e.u.)
$PV_{0\%}$	process variable at 0% of measurement scale (PV e.u.)
ΔCO	change in controller output (%)
ΔCO_{max}	maximum available change in controller output to output limit (%)
ΔPV	change in controller process variable (%)
ΔSP	change in controller setpoint (%)

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S_{1x}	1st split ranged span (PV e.u.)
S_{2x}	2nd split ranged span (PV e.u.)
S_v	valve stick-slip (resolution-threshold sensitivity) (%)
S_m	wireless trigger level (measurement threshold sensitivity) (%)
T_i	PID integral time (s/repeat)
T_o	period of limit cycle (s)
T_r	rise time of setpoint response (s)
T_u	ultimate period (s)
λ	closed loop time constant for setpoint change (s)
θ_i	implied total loop deadtime (s)
θ_o	original total loop deadtime (s)
θ_s	effective deadtime from threshold sensitivity setting (s)
$\theta_{\Delta T}$	effective deadtime from update time interval (s)
θ_v	effective deadtime from valve resolution-sensitivity (stick-slip) (s)
θ_w	effective deadtime from wireless settings (s)
τ_f	signal filter or volume attenuating time constant (s)
τ_L	load disturbance time constant (s)
τ_o	open loop time constant (s)
τ_p	process time constant (s)
ΔT_x	controller execution time (s)

14.1 Introduction

The PID controller is an essential part of the control loop in the process industry [1]. Studies have shown that the PID provides an optimal solution of the regulator problem (rejection of disturbances) and with simple enhancements, provides an optimum servo response (setpoint response) [3]. Tests show that the PID performs better than Model Predictive Control (MPC) for unmeasured disturbances in terms of peak error, integrated error, or robustness [8]. The PID controller in the modern Distributed Control System (DCS) has an extensive set of features. However, primarily due to the lack of understanding of the functionality and applicability of the PID, the full power of the PID is rarely utilized [22]. This section explores key PID features and provides examples of their importance for addressing challenging applications and control objectives for common unit operation applications in the process industry.

Industrial processes are characterized by unmeasured disturbances, nonlinear process dynamics, noise, measurement delays and lags, resolution and sensitivity limits, and valve nonlinearities and non-idealities. It will be shown that the total PID loop deadtime in industrial processes determines the ultimate limit to loop performance. The total loop deadtime has many sources most of which are variable. The process deadtimes and time constants are rarely constant. In a first order plus deadtime approximation, all of the time constants smaller than the largest open loop time constant (τ_o) become an equivalent deadtime (θ_o). The fraction of the small time constant converted to deadtime approaches 1 as the ratio of small to largest time constant approaches 0 [6]. Examples of small time constants are valve actuator

lags, process heat transfer and mixing lags, thermowell and sensor lags, transmitter damping settings, and signal filters. The deadtime from these lags are summed with the pure delays from valve pre-stroke delay, valve backlash and stiction, process and sample transportation delays, analyzer and wireless measurement update times, and PID execution time [2, 4, 6, 22, 24]. Except for damping settings, signal filters, analyzer and wireless update times, and PID execution times, these lags and delays are generally unknown and variable. The key features in a PID offer the flexibility and capability to achieve the ultimate limit to loop performance despite the challenging characteristics of industrial processes [21, 23].

14.2 Challenges and Solutions

A myriad of options and techniques have been used to address industrial automation system limitations and process objectives. Before we look at specific solutions used in industry, we need to understand the practical and ultimate limits to PID performance for unmeasured load disturbances in industrial processes. The first subsection provides practical equations developed over the years to detail the important relationships between load performance and dynamics and tuning. This subsection also offers a new equation to show how an important metric for setpoint performance also depends upon dynamics and tuning. Since setpoint changes are exactly known unlike unmeasured disturbances many methods exist to circumvent the limitations imposed by tuning. Subsequent subsections discuss methods such as setpoint feedforward and smart bang-bang logic.

14.2.1 Practical and Ultimate Limits to PID Performance

Special algorithms can be designed to deal with measured load disturbances at the process input, setpoint changes, and disturbances at the process output (e.g., noise). Often neglected is the overriding requirement that controllers in industrial applications must be able to deal with unmeasured and unknown load disturbances at the process input. Fortunately, the PID controller excels at this load disturbance rejection. An estimate of the current and best possible load rejection as a function of the process and automation system dynamics and controller tuning provides the information on what can be done to improve plant design and tuning. A simple set of equations can be developed that estimates the integrated error and peak error for a step change in a load disturbance. The value is more in helping guide decisions on improvements rather than predicting actual errors because of the uncertainty of the size and speed of load disturbances and the nonlinear and non-stationary nature of industrial processes. The equations are simple enough to provide key insights as the relative effects of the controller gain and integral time and the first order plus deadtime approximation (FOPDT) of the process and automation system dynamics. In the FOPDT model, a fraction of each of the time constants smaller than the largest

time constant is taken as equivalent deadtime and summed with the pure deadtimes to become the total loop deadtime (θ_o) termed a process deadtime (θ_p) in the literature. The fraction of the small time constants not taken as deadtime is summed with the largest time constant to become the open loop time constant (τ_o). While the equations for tuning and estimation of errors is based on the open loop time constant, we will assume the largest time constant is in the process so we have the more common term of process time constant (τ_p) seen in the literature. In reality fast loops, such as liquid flow and pressure, have a time constant in the FOPDT model much larger than the flow response deadtime due to a transmitter damping setting and signal filter time constant. Similarly, the equations seen in the literature use a process gain (K_p) rather than the open loop gain (K_o) that is the product of the final control element, process, and measurement gain. For improving dynamics, a distinction of the location of nonlinearities, deadtime, and the largest time constant are important. By avoiding the categorization of dynamics as being solely in the process, a better understanding of the effect of the final control element size, installed characteristic, stick-slip, and backlash, the effect of measurement noise, lag, delay, calibration span, and the effect of PID filter and execution time is possible. The nomenclature used in the quantification of these effects is defined at the beginning of the chapter.

Since a controller cannot compensate for an unmeasured load disturbance before the loop deadtime, the peak error (E_x) (maximum error for a disturbance) is the excursion of the first order response to the step disturbance (E_o) based on the open loop time constant for a time duration of the loop deadtime (14.1) [2]. The open loop error is the final error seen at the PID from an unmeasured load disturbance if the PID was in manual. The terms "open loop" and "closed loop" are used for a response without and with feedback correction, respectively.

$$E_x = \left[1 - e^{-\frac{\theta_o}{\tau_o}}\right] E_o. \tag{14.1}$$

If the total loop deadtime is much larger the open loop time constant, then the peak error is basically the open loop error. If the deadtime was less than the time constant, then (14.1) can be simplified to (14.2), eliminating the exponential term [2, 4]

$$E_x = \frac{\theta_o}{(\theta_o + \tau_o)} E_o. \tag{14.2}$$

The minimum integrated error (E_i) can be approximated as the area of two right triangles with the altitude equal to the peak error and the base equal to the deadtime. Taking the area of each triangle as $\frac{1}{2}$ the base multiplied by the altitude we obtain (14.3) where the integrated error is simply the peak error multiplied by the deadtime and consequently proportional to the deadtime squared [2, 4]

$$E_i = \frac{\theta_o^2}{(\theta_o + \tau_o)} E_o. \tag{14.3}$$

Equations (14.2) and (14.3) are for the minimum possible errors determined by the open loop process and system automation system dynamics. It is not possible

to do better than what is permitted by the dynamics. Thus, these are the ultimate limits to loop performance for unmeasured load disturbances. What is achieved in feedback control depends upon the tuning. In practice, controllers are not tuned aggressively enough to achieve the ultimate limit because the response tends to be too oscillatory especially for large setpoint changes and the controller lacks robustness. A 25% increase in loop deadtime or open loop gain or 25% decrease in the open loop time constant can result in oscillations that do not sufficiently decay. We can develop the equations that set the practical limit in terms of controller tuning settings from the equations for the ultimate limit based on open loop dynamics. We will also see that we can independently arrive at the same equation for the integrated error from the response of the PI algorithm to a step disturbance.

If we divide through by the deadtime term in (14.2), we have (14.4) where the peak error depends upon the ratio of the open loop time constant to total loop deadtime

$$E_x = \frac{1}{(1 + \tau_o/\theta_o)} E_o. \tag{14.4}$$

Most tuning methods for maximum disturbance rejection use a controller gain (K_c) that is proportional to the ratio of the open loop time constant to total loop deadtime and inversely proportional to the open loop gain (14.5) [2, 4, 6]

$$K_c = \frac{\tau_o}{\theta_o K_o}. \tag{14.5}$$

If we solve for the open loop time constant to total deadtime ratio, we see that this ratio is simply the product of the controller gain and open loop gain ($K_c K_o$). If we substitute the product for the ratio in (14.4), we have (14.6), which is the practical limit to the peak error [2, 4]. Peter Harriott developed the same form of the equation but with a numerator of 1.5 for the peak error from a proportional only controller tuned for quarter amplitude decaying response [5]

$$E_x = \frac{1}{(1 + K_c K_o)} E_o. \tag{14.6}$$

For time constant-to-deadtime ratios that are much larger than one, which is the case for pressure and temperature control of vessels and columns, the product of the controller gain and open loop gain is much greater than one leading to the peak error being simply inversely proportional to the product. Since the controller gain used in practice is about half of the gain for maximum disturbance rejection, we end up with (14.7) for the peak error

$$E_x = \frac{2}{K_c K_o} E_o. \tag{14.7}$$

Equation (14.7) corresponds to a peak error reached in about two deadtimes. If we approximate the integrated error as the area of two right triangles each with a

base equal to two deadtimes and consider the integral time (T_i) setting as being 4 deadtimes, we end up with (14.8) for the integrated error [2, 4, 6]

$$E_i = \frac{T_i}{K_c K_o} E_o. \tag{14.8}$$

We can derive (14.8) from the equation for a PI controller's response to an un-measured load disturbance. The change in controller output from time $t1$ to time $t2$ is the sum of the contribution from the proportional mode and the integral mode (14.9a). The module execution time (Δt) is added to the reset or integral time (T_i) to show the effect of how the integral mode is implemented in digital controllers. An integral time of zero ends up as a minimum integral time equal to the execution time so there is not a zero in the denominator for the integral mode. For analog controllers, the execution time is effectively zero [4, 6]

$$\mathrm{CO}_{t2} - \mathrm{CO}_{t1} = K_c(E_{t2} - E_{t1}) + \left[\frac{K_c}{(T_i + \Delta t)}\right]\int_{t1}^{t2} E_t \Delta t. \tag{14.9a}$$

The errors before the disturbance (E_{t1}) and after the controller has completely compensated for the disturbance (E_{t2}) are zero ($E_{t1} = E_{t2} = 0$). Therefore, the long term effect of the proportional mode, which is first term in (14.9a), is zero. Equation (14.9a) reduces to (14.9b) [4, 6]

$$\Delta \mathrm{CO} = \left[\frac{K_c}{(T_i + \Delta t)}\right]\int_{t1}^{t2} E_t \Delta t. \tag{14.9b}$$

The integrated error is the integral term in (14.9b), giving (14.9c). For over-damped response the integrated error and the integrated absolute error (IAE) are identical

$$E_i = \int_{t1}^{t2} E_t \Delta t. \tag{14.9c}$$

If we substitute (14.9c) into (14.9b), we have (14.9d)

$$\Delta \mathrm{CO} = \left[\frac{K_c}{(T_i + \Delta t)}\right] E_i. \tag{14.9d}$$

The change in controller ($\Delta \mathrm{CO}$) multiplied by the open loop gain (K_o) must equal the open loop error (E_o) for the effect of the disturbance to be eliminated. We can express this requirement as the change in output being equal to the open loop error divided by the open loop gain (14.9e) [4, 6]

$$\Delta \mathrm{CO} = \frac{E_o}{K_o}. \tag{14.9e}$$

If we substitute (14.9e) into (14.9d) and solve for the integrated error, we end up with (14.9f), which is the same as (14.8) except for the addition of the execution

time interval for the digital implementation of the PI algorithm [6]

$$E_i = \left[\frac{(T_i + \Delta t)}{K_o K_c} \right] E_o. \tag{14.9f}$$

Recently, Greg Shinskey added a term to the numerator to include the effect of a signal filter time constant on the integrated error (14.10) [24, 30–32]. In Shinskey's presentation of the equation, the change in controller output rather than the open loop error is used, which eliminates the open loop gain in the denominator. Equation (14.10) is applicable regardless of tuning settings. The additional equivalent deadtime from the filter time and execution time interval may necessitate a decrease in controller gain and increase in integral time further degrading performance [4, 24]

$$E_i = \left[\frac{(T_i + \Delta t + \tau_f)}{K_o K_c} \right] E_o. \tag{14.10}$$

To summarize, in the process industry, automation system and process dynamics, and in particular the loop deadtime, set the ultimate limit to loop performance but controller tuning sets the practical limit for unmeasured disturbances. For example, a loop with a small deadtime will perform as badly as a loop with a large deadtime if the PID has sluggish tuning. On the other hand, a PID with fast tuning may have an excessive oscillatory response for increases in the loop deadtime or process gain. Equation (14.6) shows the practical limit to the peak error (E_x) is inversely proportional to 1 plus the product of the PID gain (K_c) and the open loop gain (K_o) [2, 4, 6, 22, 24]. Equation (14.9f) indicates the integrated error (E_i) is proportional to the ratio of the PID integral time to gain (T_i/K_c) [2, 4, 6, 22, 24, 30–32, 34]. For small filters (τ_f) and PID execution time (ΔT_x), the controller gain is decreased and the integral time is increased based on the increase in loop deadtime. The filter and execution time can be added to the integral time for the integrated error to show the increase in the practical limit (14.10) [10]. For a deadtime much less than the open loop time constant, (14.2) reveals that the ultimate limit to the peak error depends upon the ratio of the total loop deadtime (θ_o) to the open loop time constant (τ_o) [2, 4, 6, 22, 24]. Equation (14.3) indicate that the integrated error depends upon the ratio of the loop deadtime squared to open loop time constant. A PID controller tuned for maximum disturbance rejection has a controller gain proportional to the ratio of the largest open loop time constant to loop deadtime (τ_o/θ_o), and an integral time proportional to the loop deadtime [2, 4, 6, 22, 24, 30–32, 34]. Note that the controller tuning depends upon the largest open loop time constant and not the process time constant. If the largest time constant is in the measurement path, the observed peak error in the measurement predicted by (14.2) will be smaller than the actual peak error in the process because of the signal filtering effect of the measurement time constant.

The peak error is important for preventing: shutdowns from reaching trip settings of safety instrumentation systems (SIS), environmental emissions and process losses from reaching the relief settings of rupture discs and relief valves, off-spec paper sheet and plastic web from exceeding permissible variation in thickness and

clarity, compressor shutdowns from crossing surge curve, and recordable incidents by exceeding environmental limits [22].

The integrated error is a good indicator of the quantity of liquid product off-spec in equipment with back mixing. In these volumes, positive and negative fluctuations in concentration are averaged out unless irreversible reactions are occurring [22].

An important emerging consideration is the realization that initial open loop response in the FOPDT approximation of a self-regulating is a ramp seen in the response of an integrating process such as level and batch temperature [2, 4, 6, 9, 24]. The ramp is more persistent in a self-regulating process with a large open loop time constant. The process is termed "near integrating" or "pseudo integrating". An equivalent integrating process gain (K_i) can be approximated as the open loop gain divided by the open loop time constant (14.11). For processes where the open loop time constant is more than ten times larger than the deadtime, the identification of this near integrator gain in 3 deadtimes can reduce the time required for process identification by more than 90% compared to those techniques that go to the 98% response time. The process variable (PV) is passed through a deadtime block to create an old PV that is subtracted from the new PV to create a ΔPV and then an integrating gain by dividing by the deadtime and the change in controller output. The maximum of a continuous train of these "near integrating" process gains updated every execution of the PID module can be used for tuning controllers on all types of processes

$$K_i = \frac{K_o}{\tau_o}. \tag{14.11}$$

If we substitute the near integrating gain for the open loop gain to time constant ratio in (14.5), we have (14.12). Recently, this method was found to even work on processes where the deadtime was greater than the time constant. To provide a smoother response, less setpoint overshoot, and more robust settings, the controller gain in both (14.5) and (14.12) is cut in half

$$K_c = \frac{1}{\theta_o K_i}. \tag{14.12}$$

The optimum integral time depends upon the type of process. The integral time ranges from about 4 times the deadtime for integrating and "near integrating" processes to one half the deadtime for severely deadtime dominant process ($\theta_o \gg \tau_o$). Equation (14.13) provides a reasonable curve fit to the required relationship for self-regulating processes. For a deadtime less much less than the time constant ($\theta_o < 0.1\tau_o$), the ultimate period is about 4 times the deadtime and the denominator is about 1, giving an integral time that is about 4 times the deadtime. For a deadtime much greater than the time constant ($\theta_o > 10\tau_o$), the ultimate period is about 2 times the deadtime and the denominator is about 4, giving an integral time that is $\frac{1}{2}$ of the deadtime.

For self-regulating processes:

$$T_i = \frac{T_u}{\min(4, 3(\frac{4\theta_o}{T_u} - 1)^2 + 1)}. \tag{14.13}$$

For a deadtime dominant process, the combination of (14.13) for integral time and (14.5) for controller gain results in almost an integral-only controller. Since the controller gain is so low, this process is a candidate for setpoint feedforward to reduce the setpoint response rise time.

For an integrating process, the product of the controller gain and integral time must be greater than twice the inverse of the integrating process gain to prevent slowly decaying oscillations from the integral mode dominating the proportional mode [9, 24]. If the user is confident in the knowledge of the integrating process gain, this relationship can be used to find the integral time (14.14a). Since the maximum controller gain allowable on many level and batch temperature loops is greater than 100 and the actual controller gain used is often less than 10, the integral time must be increased to prevent the slow rolling oscillations. Consequently, while an integral time of 4 deadtimes is possible for an integrating process, in practice an integral time of 40 deadtimes is more appropriate because the maximum controller gain is beyond the user's comfort level (14.14b).

To prevent slowly decaying oscillations integrating processes from excessive integral action, the following is used:

$$T_i > \frac{2}{K_c K_i}.$$ (14.14a)

The positive feedback in the runaway processes necessitates an integral time ten times larger than the integral time for a "near integrating" self-regulating process. The integral time should be 40 deadtimes or larger for a runaway process (14.14b). Some highly exothermic polymerization batch reactors have gone to proportional plus derivative control to avoid the problem of a user setting too small of an integral time.

For integrating processes with controller gains less than 10 times the maximum permissible controller gain and for runaway processes, one takes:

$$T_i = 40\theta_o.$$ (14.14b)

Too small of a controller gain or too large of a controller gain can cause a runaway reaction. There is a window of allowable controller gains for positive feedback processes [6, 24]. Any changes in tuning settings particularly for runaway reactions must be closely monitored.

Common metrics for a setpoint response are rise time (time to reach setpoint), overshoot (maximum error after crossing setpoint), and settling time (time settle out within a specified band around the setpoint). The ultimate limit for rise time is proportional to the loop deadtime. The ultimate limit for overshoot and settling time is theoretically zero. The practical limit to rise time is similar to the practical limit for peak error for fast tuning settings but degrades to the relationship for the integrated error for sluggish tuning settings. Fortunately, there are many features that can be used to readily help achieve the ultimate limit to the rise time. The practical limits for overshoot and settling time depend upon a balance between the contributions from the integral and proportional modes. In general, the controller

gain for maximum disturbance rejection can be used to minimize rise time and the integral time can be increased to minimize overshoot and settling time [22, 24].

The minimum rise time (T_r) can be approximated as the change in setpoint (ΔSP) divided by the maximum rate of change of the process variable. For an integrating or "near integrating" process, the maximum PV ramp rate is the integrating process (K_i) gain multiplied by the change in controller output as detailed in the denominator of (14.15a). If the step change in controller output from the proportional mode for a structure of proportional action on error is less than the maximum available output change (difference between current output and output limit), (14.15a) simplifies to (14.15b) for feedback control. The output change must be corrected for methods used to make the setpoint response faster. For setpoint feedforward, the step change in output is a combination of the feedforward and feedback action. For smart bang-bang logic, the step output change is the maximum available output change

$$T_r = \frac{\Delta\text{SP}}{K_i \min(|\Delta\text{CO}_{\text{max}}|, (K_c + K_{ff})\Delta\text{SP})} + \theta_o. \quad (14.15a)$$

For a maximum available output change larger than the step from the proportional mode ($|\Delta\text{CO}_{\text{max}}| > K_c\Delta\text{SP}$), the change in setpoint in the numerator and denominator cancel out, yielding a simpler equation without feedforward:

$$T_r = \frac{1}{(K_i K_c)} + \theta_o. \quad (14.15b)$$

For the "near integrating" process response seen in vessel and column temperature loops where the process time constant is significantly larger than the total loop deadtime, the integrating gain is the open loop gain (K_o) divided by the open loop time constant (τ_o) and (14.15b) becomes (14.15c) [6, 11, 24]

$$T_r = \frac{\tau_o}{(K_o K_c)} + \theta_o. \quad (14.15c)$$

The practical and ultimate limit to loop performance can be reconciled by realizing that there is an implied deadtime (θ_i) from the tuning. Equation (14.16) shows the implied deadtime that can be approximated as the original deadtime (θ_o) plus Lambda (λ) multiplied by a factor that is 0.5 [15, 17, 21, 23]. Lambda is the closed loop time constant for a setpoint change. For a PID tuned for maximum disturbance rejection, Lambda is set equal to the original deadtime. The implied deadtime is then equal to the original deadtime [15, 17, 22, 24]

$$\theta_i = 0.5(\lambda + \theta_o). \quad (14.16)$$

The peak and integrated errors for unmeasured step disturbances represent the worst case. Step disturbances originate from manual actions, safety, switches, and sequential operations. If discrete actions (e.g., the opening and closing of on-off valves and the starting and stopping of pumps) are replaced by control loops with modulated final control elements (throttling valves and variable speed drives) or are

attenuated by intervening volumes, the step disturbances are smoothed. The attenuated load disturbance has a time constant (τ_L) that is the residence time of the volume or closed loop time constant of the upstream control loop. To include the effect of a load time constant, the process excursion in the first deadtime, which is the key time for determining minimum peak error, can be computed by (14.17). The open loop error (E_o) in the equations for peak and integrated error can be replaced with the load disturbance (E_L), that is, the open loop error multiplied by the exponential response of the disturbance in one deadtime The effect is mitigated by a reset time that is slow relative to the disturbance time constant [22, 24]

$$E_L = \left(1 - e^{-\theta_o/\tau_L}\right)E_o. \tag{14.17}$$

PID controllers tuned too fast can introduce process variability from an oscillatory response, PID controllers tuned too slow can make a loop with good dynamics perform as badly as a loop with poor dynamics. In other words, money invested to reduce process deadtime or to get faster measurements and valves is wasted unless the PID controller tuning is commensurate with the speed of the process so that the practical limit approaches the ultimate limit to loop performance.

In some cases, slower tuning, longer wireless update times, and a PID enhanced for wireless will reduce the oscillations from feedforward timing errors and interaction between loops. Also, in cases of blend control, all of the flow loops may be forced through tuning to be as slow as the flow loop with the largest deadtime to provide a coordination of flows that leads to greater product consistency.

Since industrial processes have valve, process, and measurement dynamics that vary with time, operating point, and step size, it is important to have automated methods of tuning.

14.2.2 On-demand and Adaptive Tuning

On-Demand and Adaptive Tuning integrated into the PID function block in a DCS enables the use of PID tuning that achieves the ultimate performance limit. The relay method by Karl Åström provides a straightforward On-Demand Tuner [1, 2, 4, 6, 11]. A user-selected step change is injected into the PID output initially and any time the process variable reverses direction and crosses the setpoint and the corresponding noise band. The controller action is used to determine if the reversal in the process variable is in the correct direction to drive the process variable back to setpoint. The ultimate period (T_u) is the oscillation period. Equation (14.18) is used to compute the ultimate gain (K_u) from the PID output step size (d) and the process variable amplitude (a) corrected for the noise band (n). Figure 14.1 shows the relay oscillation method with a large change in the process variable (PV) for illustrative purposes. For processes with large time constants, the PV amplitude (a) is so small, the oscillation is barely perceptible and the oscillation period is about 4 deadtimes. Since the more important PID loops, such as temperature, have a large process time constant, the auto tuner provides a test that is less disruptive and faster

Fig. 14.1 Relay oscillation method offers fast tuning test [23]
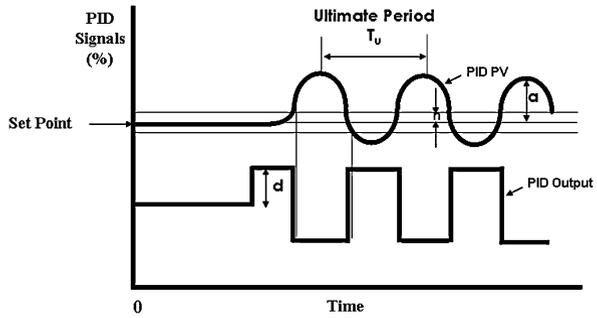
than an open loop test that is waiting to reach a new steady state to identify the process time constant. The time constant identified in relay oscillation method is not very accurate. Thus, when the relay oscillation method, tuning settings based on the ultimate period and ultimate gain are more accurate than those that require knowledge of the process time constant

$$K_u = \frac{4d}{\pi + \sqrt{a^2 - n^2}}. \tag{14.18}$$

The PID gain is the ultimate gain multiplied by a 0.25 factor [22, 24]. The PID integral time is the ultimate period multiplied by 1.0 factor for self-regulating and 10.0 for non-self-regulating processes [11, 22, 24]. The PID rate time is the ultimate period multiplied by 0.1 when derivative action is beneficial [11, 22, 24]. If the ultimate period is less than 3 times the deadtime, the rate time should be 0 since the loop is deadtime dominant (deadtime is significantly greater than the largest time constant in the loop) [11]. If the ultimate period is greater than 4 times the deadtime, rate time should be used to prevent a runaway since the process may have positive feedback and an unstable open loop response. These factors are generally in the direction to provide a non oscillatory PID response that is more robust (more resistance to excessive oscillations from changes in process dynamics). The Ziegler–Nichols factors were designed to provide a quarter amplitude response (amplitude of each succeeding oscillation is $\frac{1}{4}$ the amplitude of the last oscillation). Most publications on tuning based on the ultimate period and ultimate gain use the Ziegler–Nichols factors leading to improper conclusions on smoothness and robustness of the tuning method [11].

Adaptive tuners use a more advanced method to identify process dynamics without relay oscillations. Significant manual and remote output changes and setpoint changes trigger the search for the dynamic parameters for a first order plus deadtime approximation (process gain, deadtime, and time constant) that provides a model's response that matches the process response. A particular adaptive tuner computes the integrated squared error (ISE) between the model and the process output for changes in each of three model parameters from the last best value. Exploring all combinations of three values (low, middle, and high) for three parameters, results in 27 models. The correction in each model parameter is interpolated by the application of weighting factors that are based on the ISE for each model normalized
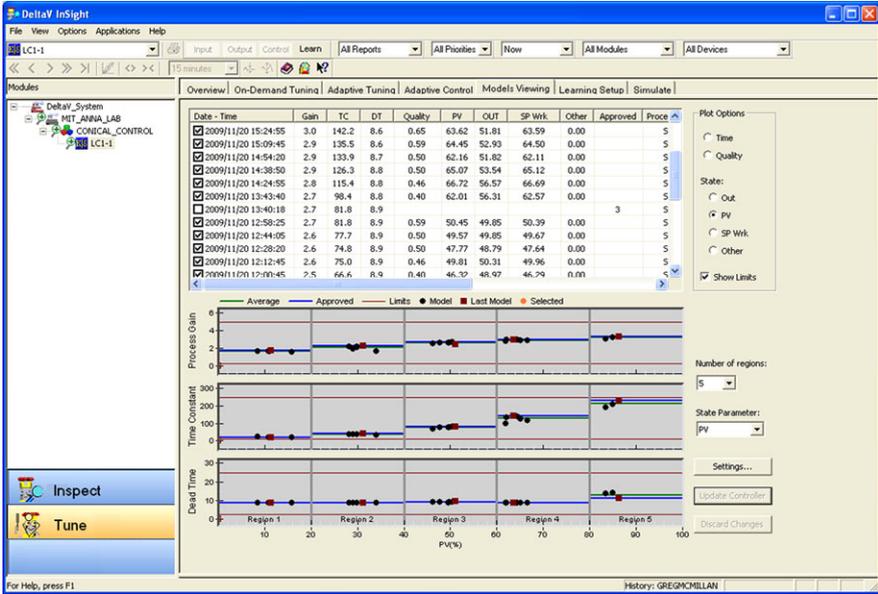
Fig. 14.2 Models enable adaptive level control of conical tank [19]

to a total ISE for all the models over the period of interest. After the best values are computed for each parameter, they are assigned as the middle values for the next iteration. This model switching with interpolation and re-centering has been proven mathematically by the University of California, Santa Barbara to be equivalent to a least square identification that provides an optimum approach to the correct model [10, 36].

Adaptive tuners schedule tuning settings identified for regions defined by a user-selected variable. For valves with nonlinear characteristics such as equal percentage, the variable for scheduling is the PID's output. For nonlinear processes, such as pH, the variable for scheduling is the PID's process variable. The scheduling provides preemptive correction of the tuning settings eliminating the delay in performance associated with the re-identification of settings as the PID moves into another region [10, 24, 29, 36]. For a gravity discharge conical tank, adaptive tuning made the level setpoint response fast with a consistent settling time over the entire range of operation by increasing the process time constant as the cross-section area increased from bottom to top [29]. In this example, the gravity discharge flow makes the process self-regulating rather than integrating. Consequently, the nonlinearity of the change in cross-sectional area predominantly affects the process time constant rather than the process gain. Figure 14.2 shows the models automatically identified in five regions for scheduling tuning settings to account for the changes in cross-section with level.

Since an adaptive tuner uses current tuning settings to compute process dynamics as the starting point for its search, the number of tests required to get an adaptive

Fig. 14.3 Positive feedback integral mode enables key PID features (external reset, wireless enhancement, and deadtime compensation) [4, 23]

model with a high fidelity rating can be minimized by first running the On-Demand tuner with a requirement of just 2 or 3 cycles. Since the cycle period is on the average the ultimate period, the test is usually faster than an Adaptive Tuning test, especially for the overly conservative (sluggish) tuning commonly found in industrial PID controllers that have not been tuned by an automated method.

The step size in the output for On-Demand and Adaptive Tuning should be at least: 5 times the noise band, the trigger level of a wireless device, and the dead band and resolution-threshold sensitivity of the control valve [2]. Note that these step changes will not show the deadtime from wireless update times and valve backlash and stick-slip. For wireless devices, about half of the default update rate should be added to the identified deadtime [15, 22, 24, 30–32, 36].

14.2.3 Positive Feedback Implementation of Integral Mode

Instead of integrating the error, the feeding back of the controller output or external reset signal through a filter block and adding it to the contribution of the proportional and derivative modes creates an integral mode action where the filter time constant is the integral time setting [4, 24]. When the error is zero, the output of the filter block is simply the controller output or external reset signal and integral action stops. The positive feedback implementation illustrated in Fig. 14.3 enables several important PID options, such as dynamic reset limit, enhancement for wireless, and deadtime compensation. Figure 14.3 is for the ISA standard form for the PID controller. The eight structures commonly used in industrial processes are obtained by setting the

setpoint weight factor β for the proportional and the setpoint weight factor γ for the integral mode in Fig. 14.3. If the factor is zero, a setpoint change does not affect the contribution to the output from respective mode (action is on PV only). If the factor is one, the full effect of a setpoint change is included (full action is on error). A factor between zero and one provides the ability to include but moderate the effect of a setpoint change (balanced action on setpoint change and PV change). In this figure, the multiplication symbol "*" in a circle is used to denote the multiplication by the β or γ weight factor.

The eight PID structures commonly used in industrial processes are:

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

β and γ are setpoint multiplication factors for the proportional and derivative modes, respectively, to determine how much proportional and derivative action occurs on setpoint changes. These factors do not affect the ability of the PID to reject disturbances. For the fastest possible setpoint response, structures 1 and 2 are used. If preventing overshoot is more important than minimizing rise time, structure 3 is used. If the ability to customize the balance between fast rise time and minimum overshoot for a setpoint response is needed, structure 8 is used. This structure also offers the ability to achieve both good load and setpoint responses.

14.2.4 Dynamic Reset Limit (External Reset)

When an external signal is used as the input to a "Filter" block in the positive feedback implementation of the integral mode, the integral action will not drive the controller output faster than the external reset signal is changing. This capability is particularly important for slow final control elements (large valves and variable frequency drives), cascade control, and override control.

If the external reset signal is the actual valve position or variable frequency drive (VFD) speed, the PID controller output will not ramp faster than the valve or VFD can respond [24]. Control valves and dampers have a slewing rate that increases with actuator size and stroke length. Damper slewing rate is particularly slow due to the need to prevent positive feedback from negative torque requirement. VFDs have velocity limiting of the command signal to prevent overloading the motor. If the external reset signal is the secondary loop process variable (PV) for cascade control, the primary PID cannot ramp the setpoint of the secondary PID faster than the secondary PID PV can respond. This capability is important for inherently preventing severe oscillations from breaking out for large setpoint changes or large

disturbances [23, 33]. The use of the selected PID output as an external reset signal for override control also inherently prevents the unselected PID controllers from ramping off-scale. PID algorithms without the positive feedback implementation of integral action, add a "Filter" block to the external reset signal with a filter time equal to the PID reset time to prevent the ramping off-scale of the unselected PID output. The dynamic reset limit is a key feature that enables the development of an enhancement of the PID for wireless measurements that also has the ability to eliminate oscillations from threshold sensitivity and resolution limits and feedforward timing errors [20, 21, 24].

The dynamic reset limit can open opportunities important for sustainable manufacturing and in particular abnormal situation management and optimization. If a setpoint velocity limit is set in the analog output block, the dynamic reset limit prevents the PID from going faster than the velocity limit. The PID can achieve a slow approach to an optimum and a fast recovery upon encroachment of a constraint such as encountered in the prevention of compressor surge, exothermic reactor runaway, RCRA pH violation, and Bioreactor biomass starvation. Previously, an open loop back-up (kicker) has been used for these applications because the tuning of the controller for drastically different speeds of actuation is problematic. The dynamic reset limit option eliminates the need to tune the controller based on direction and the concern about the exact value of the velocity limit. The tuning is set for the fastest recovery. The velocity limit is adjusted for the slowest approach to the optimum.

There are many more examples where an intelligent adaptation of the speed of actuation of the final control element or secondary loop could be beneficial. In general, you want to approach optimums slowly to minimize disruption but as you operate close to the edge, you depend upon a fast recovery to prevent going over the edge. With compressor surge control the edge is literally a cliff. While other applications might not be as dramatic, the technique opens a wide spectrum of PID techniques for sustainable manufacturing, which in its broadest definition includes efficiency, flexibility, operability, maintainability, safety, and profitability [35].

14.2.5 Enhancements for Wireless

Wireless measurement devices have a "default update rate" (time interval for periodic reporting) and a "trigger level" (threshold sensitivity limit for exception reporting) set as large as possible to conserve battery life. The integral mode in the traditional PID will continue to ramp while the PID is waiting for an updated measurement from a wireless device. Also, when an update is received, the traditional PID considers the entire change to have occurred within the PID execution time interval (ΔT_x). If derivative mode is used, the rate of change of the measurement is the difference between the new and old measurement divided by the PID execution time interval. The result is a spike in the controller output.

The non-continuous update scenario occurs for many applications besides wireless devices. During the time when a measurement is not updated due to a failure,
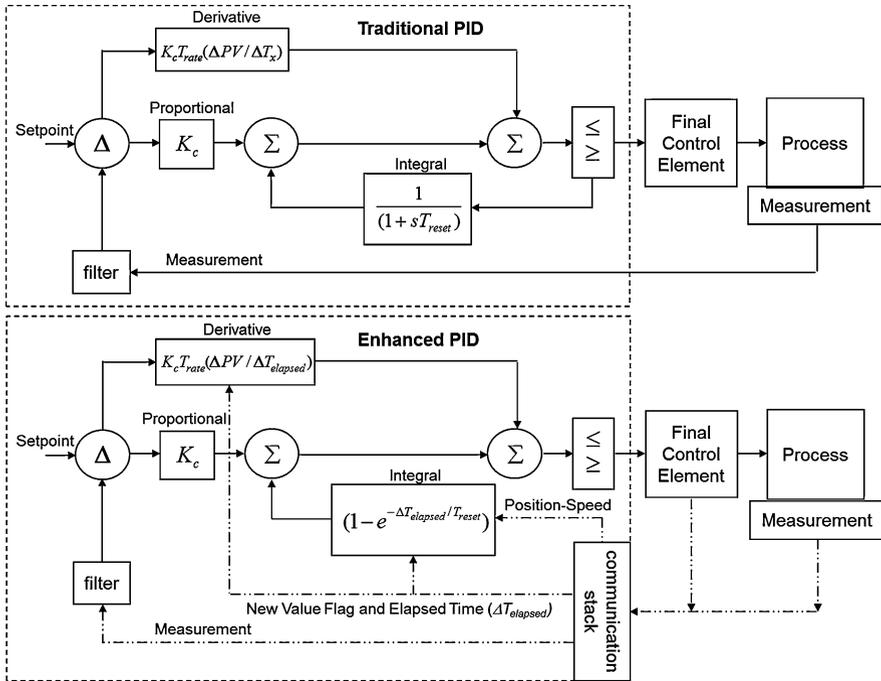
Fig. 14.4 Enhancements of PID for wireless prevent the ramping from integral action and the spikes from derivative action for discontinuous updates [21]

resolution limit, threshold sensitivity limit, or backlash, the PID output continues to ramp from the integral mode. Failures, resolution limits, and threshold sensitivity limits can originate in an analyzer, sensor, transmitter, communication system, or control valve. Analyzers also have a time interval between updates determined by the sample time and cycle time.

The enhanced PID for wireless executes the PID algorithm as fast as wired devices. A change in setpoint, feedforward signal, and remote output translates immediately (within PID execution time interval) to a change in PID output. However, integral action does not make a change in the output until there is an update. When an update occurs, the elapsed time between the updates is used in an exponential calculation that mimics the action of the filter block in the positive feedback implementation of integral action. If derivative action is used, the elapsed time rather than the PID execution time interval is used to calculate the rate of change of the process variable. The integral and derivative calculations are executed only once upon a change in setpoint or measurement [21, 23, 24]. A threshold sensitivity setting is used to prevent an update from noise. Figure 14.4 compares a simplified block diagram of the traditional PID to the enhanced PID.

A traditional PID will have to be detuned to prevent instability for a large increase in the time between updates. The enhanced PID will continue to be stable for even the longest update time interval. For a measurement update time interval larger

than the process response time, the enhanced PID controller gain can be set equal to the inverse of the open loop gain (product of valve, process, and measurement gain) to provide a complete correction for setpoint change or update. Subsequent sections show the enhanced PID can suppress oscillations from a wide variety of sources. This reduction in variability results from the suspension of integral action and the wait in feedback correction till there is a more complete response [26]. To achieve these benefits, the user simply enables the enhanced PID option in the PID block, which automatically enables the dynamic reset limit option. No retuning is necessary to achieve a smooth response but if the update time is larger than the process response time the enhanced PID can be tuned with a much higher gain.

14.2.6 Deadtime Compensation

Adding a "Deadtime" block to the external reset of a positive feedback implementation of the integral mode can provide deadtime compensation equivalent to a Smith Predictor but with the advantage that the setup is simpler and the process variable is intact. In the positive feedback implementation of deadtime compensation, the user simply needs to set the deadtime parameter in the "Deadtime" block equal to the total loop deadtime. The dynamic reset limit option for the PID must be enabled so the external reset signal is used. The process gain and process time constant parameters used in a Smith Predictor are not necessary for this implementation of deadtime compensation. To get the benefit from the PID knowing the effect of deadtime, the integral time needs to be decreased and the controller gain needs to be increased [23]. Like the Smith Predictor, this deadtime compensator is more sensitive to an overestimate rather than an underestimate of the total loop deadtime. Normally, a PID will just become sluggish if an overestimate of the deadtime is used for the tuning settings. For PID controllers with deadtime compensation, high frequency oscillations will rapidly start for overestimates of the loop deadtime [11, 23, 24]. Thus, for robustness it is better to use a deadtime that is always less than the minimum loop deadtime often associated with high production rates.

In tests, the following myths about deadtime compensators were exposed [23]:

1. **Deadtime is eliminated from the loop.** The Smith predictor, which created a PV without deadtime, fools the controller into thinking there is no deadtime. However, for an unmeasured disturbance, the loop deadtime still causes a delay in terms of when the loop can see the disturbance and when the loop can enact a correction that arrives in the process at the same point as the disturbance. The ultimate limit to the peak error and integrated error for an unmeasured disturbance are still proportional to the deadtime, and deadtime squared, respectively.

2. **Control is faster for existing tuning settings.** The addition of deadtime compensation actually slows down the response for the existing tuning settings. Setpoint metrics, such as rise time, and load response metrics, such as peak error, will be adversely affected. Assuming the PID was tuned for a smooth stable response, the controller must be retuned for a faster response. For a PID already tuned for maximum disturbance rejection, the gain can be increased by 250%. For deadtime dominant systems where the total loop deadtime is much greater than the largest loop time constant (hopefully the process time constant), the reset time must also be decreased or there will be severe undershoot. If you decrease the reset time to its optimum, undershoot and overshoot are about equal. For a test case where the total loop deadtime to primary process time constant ratio was 10:1, the reset time could be decreased by a factor of 10. Further study is needed as to whether the ratio of the old to new reset time is comparable to the ratio of deadtime to time constant and whether half of the total loop deadtime or the module execution time (0.5 s) is the low limit to the reset time for an accurate deadtime estimate.

3. **Compensator works better for loops dominated by a large deadtime.** The reduction in rise time is greatest and the sensitivity to percent deadtime modeling error particularly for an overestimate of deadtime is least for the loop that was dominated by the process time constant. You could have a deadtime estimate that was 100% high before you would see a significant jagged response when the process time constant was much larger than the process deadtime. For a deadtime estimate that was 50% too low, some rounded oscillations developed for this loop. The loop simply degrades to the response that would occur from the high PID gain as the compensator deadtime is decreased to zero. While the magnitude of the error in deadtime seems small for the test case, you have to remember that for an industrial temperature control application, the loop deadtime and process time constant would be often at least 100 times larger. For a 400 second deadtime and 10,000 second process time constant, a compensator deadtime 200 seconds smaller or 400 seconds larger than actual would start to cause a problem. In contrast, the deadtime dominant loop developed a jagged response for a deadtime that was high or low by just 10%. This requirement is unreasonable in industrial processes. A small filter of 1 second on the input to the deadtime block can help.

4. **An underestimate of the deadtime leads to instability.** In tuning calculations for a conventional PID, a smaller than actual deadtime can cause an excessively oscillatory response. Contrary to the effect of deadtime on tuning calculations, a compensator deadtime smaller than actual deadtime will only cause instability if the controller is tuned aggressively after the deadtime compensator is added.

5. **An overestimate of the deadtime leads to sluggish response and greater stability.** In tuning calculations for a conventional PID, a larger than actual deadtime simply causes a slow smooth response. Contrary to the effect of deadtime on tuning calculations, a compensator deadtime greater than actual deadtime will cause jagged irregular oscillations.

14.2.7 Fast Setpoint Response

The rise time can be minimized by using the maximum possible controller gain and using a PID structure that has proportional and derivative action on setpoint changes. For slow loops, such as temperature and composition on vessels and columns, it is particularly important to make the approach to setpoint as fast as possible to minimize cycle time for batch operations and the startup and product grade transition time for continuous operations. Overdrive (driving the PID output beyond its final resting value) is essential for getting these slow loops to setpoint quickly. Fortunately, the deadtime to process time constant ratio for these loops is so small that large PID gains are permissible. Temperature loops often have significant secondary time constants from heat transfer surface and thermowell lags that would benefit from the use of rate action. A structure of PID action on error will provide a step in the PID output from the proportional mode and a bump in the PID output from the derivative mode for a step change in the setpoint. For this beneficial action to occur, the option SP track PV should be used and the controller must be in the auto mode when the setpoint change is made.

A setpoint feedforward signal added to the controller output can be useful if the change in PID output from the proportional mode is not sufficient. For small setpoint changes and for low controller gains, setpoint feedforward can reduce rise time. The feedforward action is the process action, which is the opposite of the control action, taking into account valve action. In other words, for a reverse control action, the feedforward action is direct, provided the valve action is inc-open or the analog output block, I/P, or positioner reverses the signal for an inc-close valve. For control loops where the loop deadtime is larger than the process time constant, the feedforward gain is approximately the inverse of the open loop gain minus the PID gain for enhanced PID structures with P action on error.

If the final resting value (FRV) is preloaded as an external reset signal during the rise time (dynamic reset limit option is enabled), the overshoot of the setpoint can be minimized [34]. If the FRV is not accurately known, the preload is prematurely disabled. Various versions of "batch controllers" since the 1960s have preloaded the integral mode.

The FRV can be captured from previous batches and is often the split range point. For changes in the setpoint of batch temperature and pH loops already in service that have an integrating response, the FRV is the PID output just before the setpoint change. For continuous process loops, the FRV is PID output just before the setpoint change plus the setpoint change multiplied by the inverse of the open loop gain for a self-regulating response [13, 24]. In all of these cases, the estimated FRV is based on an assumption that changes in process load and process disturbances are negligible during the setpoint change.

The fastest possible rise time with minimum overshoot and settling time is obtained by smart bang-bang logic that uses a simple prediction of when the process variable will reach setpoint. The PID output is stepped to its output limit to maximize the rate of approach to setpoint. When the projected PV equals the setpoint less a bias, the PID output is repositioned to the FRV. The PID output is held at the
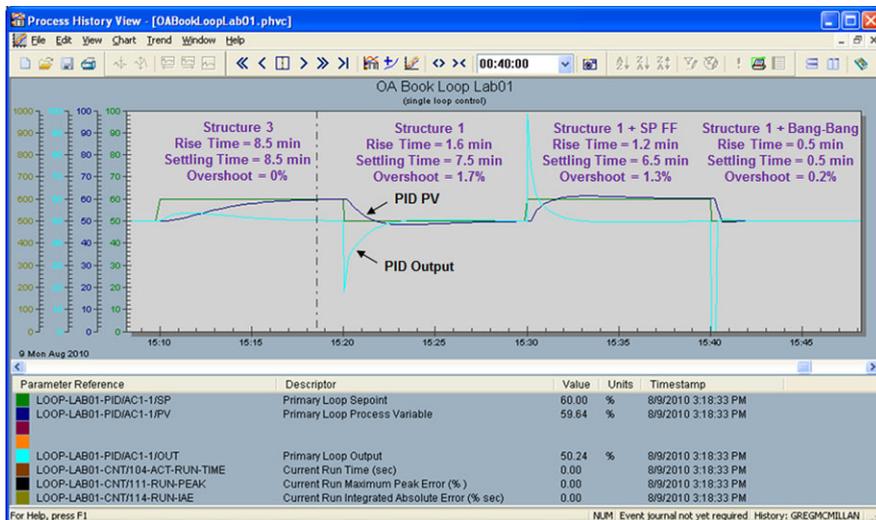
Fig. 14.5 Setpoint response shows rise time can be successively faster by a PID on error structure, setpoint feedforward, and a smart bang-bang logic [24]

FRV for one deadtime and then released for feedback control [13, 24]. A deadtime (DT) block must be used to compute the rate of change so that new values of the PV are seen immediately as a change in the rate of approach. If the total loop deadtime is used in the DT block, the projected PV is simply the current PV minus the output of the DT block (expected PV change over next deadtime) plus the current PV [24]. For self-regulating processes, such as flow with the loop deadtime approaching or greater than the largest process time constant, the logic is revised to step the PID output immediately to the FRV. Overdrive is not advisable in loops where the process time constant or the inverse of the integrating process gain is less than the deadtime because the potential for overshoot is too great for a slight underestimate of the deadtime. The total loop deadtime (without the effect of threshold sensitivity and resolution limits) can be identified at the start of the setpoint response.

Figure 14.5 shows that a PID structure of proportional and derivative action on error (structure 1) and setpoint feedforward can make the setpoint response faster, but the biggest improvement is achieved by smart bang-bang logic for an integrating process [24]. For self-regulating processes where the deadtime is larger than the process time constant, setpoint feedforward suffices.

14.2.8 Signal Linearization

Signal characterizers are used on the PID output to compensate for the nonlinear gain of control valves. The characterizer computes the percent stroke (X axis of the installed characteristic) from the percent flow (Y axis) based on the inherent trim

characteristic (e.g., equal percentage), a given system resistance curve, and the net static head [6]. A pressure drop measurement across the valve eliminates the need to know the resistance curve and static head, but is rarely available. While characterizers are available in positioners, the location of the characterizer in the configuration provides better visibility and accessibility and offers the opportunity to display the valve signal to the operator after linearization as well as before linearization so no arises confusion in checking valve positions in the field.

Signal characterizers are used on the PID process variable (PV) to compensate for operating point nonlinearities of the process gain. The most common example is pH where the signal characterizer computes the percent reagent demand (X axis of the titration curve) from pH (Y axis) for a given composition of acids and bases in the feed. While the X-axis in actually the ratio of reagent flow to feed flow, for purposes of signal characterization the X axis is simply scaled 0 to 100% reagent demand [7, 24, 28]. Since the operator typically wants to enter an SP and see and trend the PV in pH units, both the SP and PV as pH and reagent demand are on the operator interface. A signal characterizer is used to convert the operator entered SP from pH to percent reagent demand. Since ensuring stability with faster tuning settings for higher process gains is the primary objective of the signal characterizer, the fidelity of the titration curve slope in the steepest regions is most important [7, 24].

Signal characterizers could also be used to compensate for the nonlinear process gain for temperature control in distillation columns. For example, signal characterizer based on a plot of tray or packing temperature versus reflux to feed ratio could be used to compute a linear reflux demand signal from temperature. The PID SP would be percent reflux demand from an operator-entered temperature setpoint. Pressure compensation could be applied to either the PID setpoint or input but not to both.

The use of signal characterization provides a finer resolution of process gain corrections than adaptive tuning but does not eliminate the need for adaptive tuning. The curves on which the signal characterization is based will change with operating conditions. The use of characterization and adaptive tuning is a synergistic relationship. If the signal characterization was perfect, the process gain would be one. The degree of difference between the identified process gain and a unity gain is a measure of the fidelity of the curve in various operating regions that could lead to a better curve.

14.2.9 Open Loop Backup (Kicker)

Some excursions are too fast and the consequences too severe or unstable to rely on feedback control alone. In these cases, an open loop backup is used to rapidly change the PID output to get the loop out of danger. The most common example is preventing compressor surge but other examples including preventing environmental violations and runaway reactions [24].

Upon detection of an excursion towards an unstable condition or environment violation, the open loop backup puts the PID in remote output (ROUT) and increments or decrements the PID output until the excursion is stopped. The open loop backup then waits for at least one deadtime before switching the PID back to its last mode bumplessly. The detections cited in the literature have been based on a PV trigger point, a PV rate of change, or a predicted PV similar to what is used by smart bang-bang logic for a fast setpoint response [24]. The PID execution must be less than 10% of the total deadtime. There may be some overuse of energy or reagent in the process to ensure personnel, property, and environmental protection for worst case conditions. In general, the avoidance of loss of damage and downtime more than pays for the short term costs.

The use of direction velocity limits in the analog output block in conjunction with the dynamic reset option in a PID with the positive feedback implementation of integral action can replace the need for some open loop backups. The velocity limits are set to provide a slow approach to the optimum and a fast getaway. For compressor surge control, the oscillations may be too fast and too severe for recovery by just PID feedback control.

14.2.10 Final Element Resolution, Threshold Sensitivity, and Backlash

Final control elements use the PID output to manipulate a flow, which is the predominant input to an industrial process. The most common final control element used in production units is a control valve. For utility systems, the final control element is often a damper due to large sizes and ducts. For agitators and pumps with large variable dynamic loads, a variable frequency drive is used to manipulate the speed as the final control element. For mixing, the agitating rate creates mixing flow rate that is called the agitator pumping rate.

All control valves have some degree of stick-slip and backlash. For sliding stem control valves with diaphragm actuators and digital positioners, the stick-slip and backlash can be less than 0.2%. For rotary valves originally designed for on-off tight shutoff rather than throttling service, the stick-slip and backlash can be as large as 10%. Furthermore, smart digital positioners are most likely measuring actuator shaft position in these cases which is not indicative of the actual position of the internal closure element (plug, disc, or ball) [9, 14, 18, 23–25]. The control valve with the least stick-slip and backlash is the sliding stem valve with low friction packing, diaphragm actuator, and a digital positioner. Figure 14.6 shows a control valve with a large backlash and some stick-slip near the closed position.

Stick-slip can be approximated as either a resolution limit or a threshold sensitivity limit [25]. For a resolution limit, the valve moves in steps which correspond to a slip equal to a stick. For a threshold sensitivity limit, slip is variable and results in the position of the internal closure element (plug, disc, or ball) momentarily catching up with the desired position set by the output of the PID. The source of stick-slip
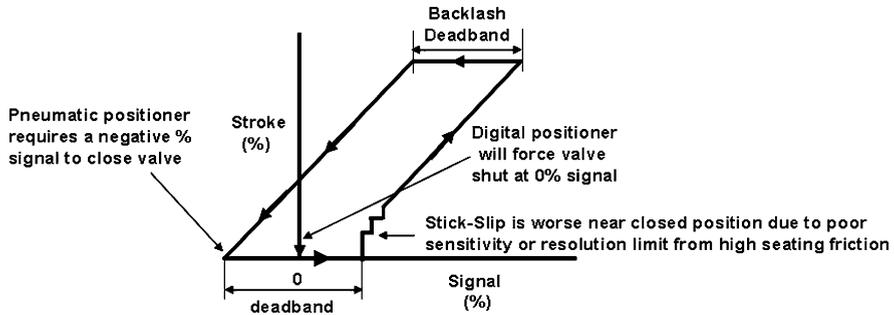
Fig. 14.6 Backlash causes a deadband on signal reversal and stick-slip causes a stair step response that is largest near the seat where the friction is greatest [24]

is friction of valve stem packing and friction of the seating and sealing of the internal closure element. The stick-slip is greatest near the closed position as the internal closure element approaches the closed position. Valves designed for tight shut-off tend to have the greatest stick-slip. For rotary valves, the sealing of the ball or disc can cause shaft windup where the shaft or stem twists but the ball or disc does not move. When the ball or disc breaks free, the valve jumps (slips) to a position that can exceed the desired position. This behavior can be approximated as a threshold sensitivity limit with overshoot [9, 14, 18].

Rack and pinion piston actuators and gear driven motor actuators have a resolution limit that corresponds to the teeth spacing. Double acting pistons have a threshold sensitivity limit that depends upon the pressure unbalance needed to overcome O-ring seal friction. Variable frequency drives (VFD) have a resolution limit set by the number of bits in the A/D input card for the command signal. Unfortunately, the standard VFD input card has only 8 bits which with 1 sign bit leaves only 7 bits for resolution of the signal (0.78% resolution) [9, 14, 18].

Backlash can be approximated by a deadband which is the amount of signal change required to reverse the direction of the valve stroke. Once the valve moves, the valve position catches up to the signal if there is no stick-slip. While technically valve deadband is defined for a full scale stroke, it can occur at any position for a signal reversal [9, 14, 18]. The primary sources of deadband are the translation of linear actuator to rotary motion and what might seem like insignificant gaps in linkages and looseness in actuator to valve stem connections [18]. Backlash is greatest for dampers and rotary valves with piston actuators originally designed for on-off action. Variable frequency drives introduce a backlash by a deadband setting used to prevent the drive from hunting or reacting to noise. Unfortunately, this deadband is often set too large because users are not aware of the detrimental impact of the backlash created.

Resolution and threshold sensitivity limits cause a limit cycle (constant amplitude sustained oscillation) in any process where the PID has integral action. Equation (14.19a) shows that the oscillation amplitude (A_o) is set by the open loop gain (K_o) and valve stick-slip (threshold sensitivity) (S_s) and is therefore independent of PID tuning. Equation (14.19b) shows the period depends upon the PID gain and integral

time besides the open loop gain. The open loop gain is the product of the valve, process, and measurement gains as detailed by (14.19c). For high process gains, such as that encountered on the steep slopes of titration curves for pH control, stick-slip can cause unacceptable variability from the PID in automatic. The limit cycle oscillation period (T_o) will increase as the PID integral time is increased. For small degrees of stick-slip, noise, frequent disturbances, or historian data compression the limit cycle pattern may not be discernable [9, 18]

$$A_o = S_v K_o, \tag{14.19a}$$

$$T_o = 4T_i\big[1/(K_o K_c) - 1\big], \tag{14.19b}$$

$$K_o = K_v K_p K_m, \tag{14.19c}$$

$$K_m = \frac{100\%}{\mathrm{abs}(\mathrm{PV}_{100\%} - \mathrm{PV}_{0\%})}. \tag{14.19d}$$

If there are two or more integrators in the process and control system that affect the final element, deadband from backlash (B_v) will cause a limit cycle. The integrators can be the result of a cascade loop that has integral action in both the primary and secondary controllers or the result of an integrating process with a single controller with integral action. For an integrating process, such as level, the controller output is a sinusoidal oscillation and the process variable ramps with some rounding of the peaks (smoothed sawtooth) from backlash. The flow is a clipped oscillation. If there are no disturbances, the result is a limit cycle. Equation (14.20a) indicates the limit cycle amplitude is the deadband divided by the controller gain per. Equation (14.20b) shows the limit cycle period is proportional to integral time and is inversely related to controller gain. Detuning the controller (decreasing the controller gain) increases both limit cycle amplitude and period. Detuning also increase the deadtime for disturbances [9, 18]

$$A_o = B_v/K_c, \tag{14.20a}$$

$$T_o = 5T_i\big[1 - 2/\big(K_c^{0.5}\big)\big]. \tag{14.20b}$$

Deadband (backlash) limit cycles can be eliminated by suspending integral action in controllers so that the total number of integrators in the control loop including the process is one or less. Thus backlash limit cycles can be eliminated in processes with a single integrator, such as level, by turning off integral action in all PID controllers. For a self-regulating process, integral action would be permitted in one PID. Backlash limit cycles cannot be eliminated in a loop with two integrating processes. While processes with two integrators are rare, a possible example is a cascade loop where the primary loop is batch temperature and the secondary loop is vessel pressure. A runaway process, such as a highly exothermic reaction, counts as an integrator [24].

Threshold sensitivity or resolution (stick-slip) limit cycles can be eliminated by suspending integral action in controllers so that the total number of integrators in the control loop including the process is zero. Thus stick-slip limit cycles can be

eliminated in self-regulating processes by turning off integral action in all PID controllers [20, 21, 24].

Many PID controllers have the option to turn off integral action when the process variable (PV) is within a band centered about the setpoint. The band is set equal to maximum amplitude of the limit cycle above and below setpoint. In some PID blocks, the parameter is called integral deadband or "IDEADBAND". Since the amount of backlash and stick-slip varies considerably with stroke and time due to changes in piston actuator O-rings, linkages, packing tightness, solids, pressure, and temperature, finding the right setting is problematic. Also, tuning algorithms generally do not include the effect of the integral deadband setting in tuning calculations. Finally, the response to load disturbances is delayed particularly for loops where the PID gain is low and hence the proportional action alone is insufficient.

The enhanced PID inherently suspends integral action when there is no change in the PV. Hence, the wireless PID can automatically eliminate limit cycles from backlash in a single integrating process and limit cycles from stick-slip in a self-regulating process. No settings are required and normal tuning procedures can be used. Furthermore, the integral reaction to disturbances is not delayed. The use of wireless transmitters is not required. The user simply needs to enable the enhanced PID [20, 21, 24]. Figure 14.7 shows the enhanced PID inherently stops the limit cycle without the need of adjustments regardless of deadband or stick-slip size.

While the application described is for limit cycles originating in the valve or variable speed drive response, limit cycles can also occur due to threshold sensitivity or resolution limits in measurements. Pneumatic, mechanical, wireless, and analytical instruments have precision limits that can create appreciable limit cycles. Again, the enhanced PID inherently eliminates the limit cycles [25, 26].

14.2.11 Slow Final Control Element Response

Control valves have a velocity-limited exponential response. The velocity limit known as the slewing rate can be quite slow for large valves and dampers. Dampers are particularly slow with slewing rates of less than 1% per second because of large duct sizes and the need to prevent dynamic instability of the wafer or blade. Variable frequency drives also have a velocity limit or ramp rate to prevent motor overload. This ramp rate is often set too conservatively because of a lack of knowledge of the impact on process control.

If the PID controller output changes faster than the final control element can respond, the loop will break out into oscillations. The problem may only occur for large changes in setpoint or disturbances. The fact that the loop is fine for small changes and intermittently develops stability problems is confusing to operations. The solution is to use positive feedback implementation of integral action with the external reset signal for the dynamic reset limit being a readback of valve stroke or variable frequency drive speed [24, 33]. The use of auxiliary variables (second, third, or fourth process variable) in HART communication is generally not fast enough.
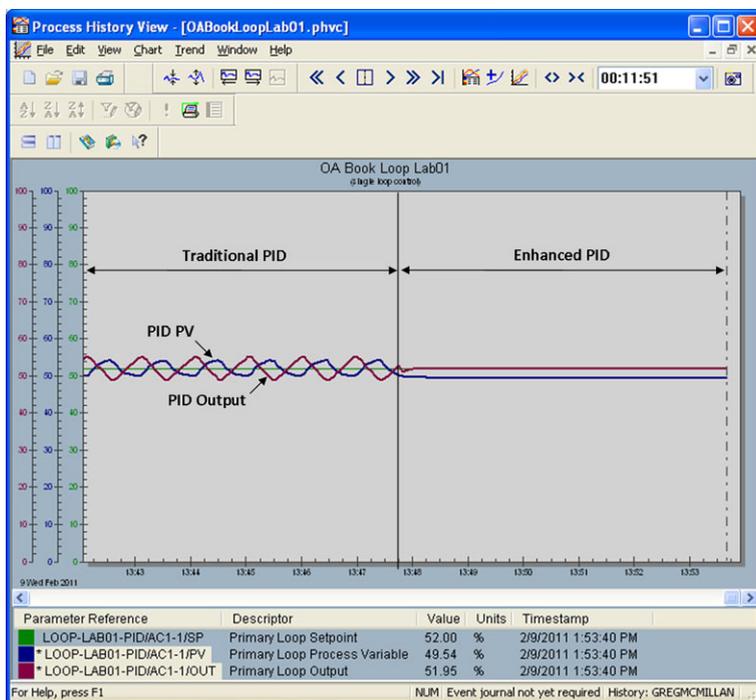
Fig. 14.7 Enhanced PID eliminates limit cycles from deadband and threshold sensitivity or resolution by suspending integral action when there is no update

The stroke or speed for external reset needs to be a primary process variable. Ideally, the valve should be speeded up by the use of volume boosters on the positioner output [18, 19, 24]. The constraints on VFD ramp rate must be more intelligently set. For most pumps and fans with properly sized motors, the speed ramp rate can be quite fast (e.g., 10% per second) without motor overload. Figure 14.8 shows that the burst of oscillations for large setpoint changes in a loop with a slow final control element are eliminated by dynamic reset limit option.

14.2.12 Slow Secondary Loop Response

If the secondary loop is slower than the primary loop, we have a problem similar to that discussed in the section on slow final control elements. For large changes in the setpoint or disturbance of the primary loop, oscillations develop. The solution is the same: dynamic reset limit should be used. In this case, the external reset is the process variable of the secondary loop.

Ideally, the secondary loop should be 4 times faster than the primary loop for maximum secondary disturbance rejection and minimal oscillations from interactions between the loops. If the secondary loop cannot be made faster, the primary
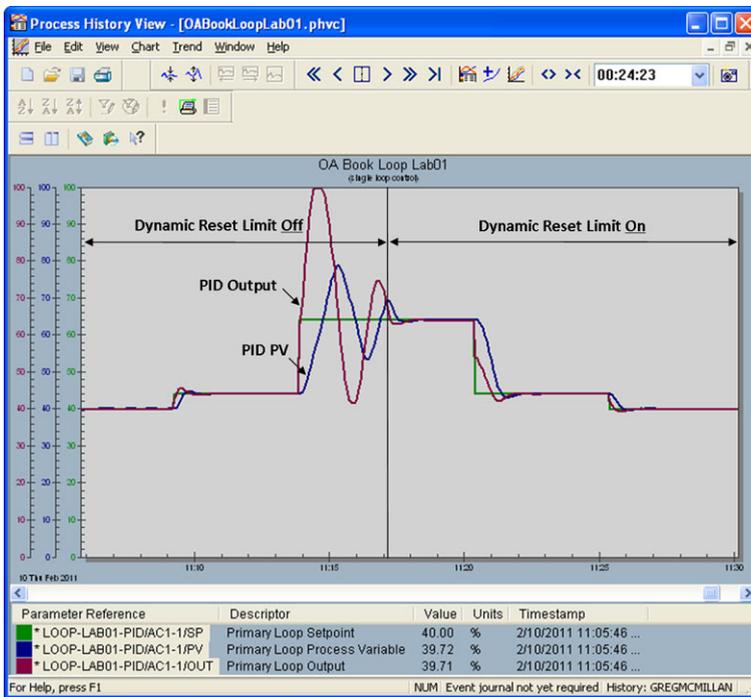
Fig. 14.8 Dynamic reset limit option eliminates the oscillations for a slow valve or VFD that occur for large changes in a PID setpoint or load

loop may need to be slowed down. If wireless measurements with update rates larger than the 63% response time are used, the default update rate of the secondary measurement should be 4 times faster than the default update rate of the primary measurement per the cascade rule that the secondary loop is sufficiently faster than the primary loop to prevent interaction between the loops. For a cascade control of static mixer pH to reagent flow using the enhanced PID, a default update time of 60 seconds for pH and 16 seconds for flow provided exceptional pH setpoint control. The PID gains were set equal to the inverse of the open loop gain [20, 21]. Figure 14.9 shows that a cascade loop of pH to reagent flow on a static mixer with the enhanced PID can provide nearly perfect setpoint control.

14.2.13 Large Wireless Update Times

If the wireless update time is larger than the 63% process response time without the update delay, the use of the enhanced PID provides excellent setpoint control by allowing the use of a PID gain that is the inverse of the open loop gain. The enhanced PID is stable without retuning as the wireless default update time is increased to
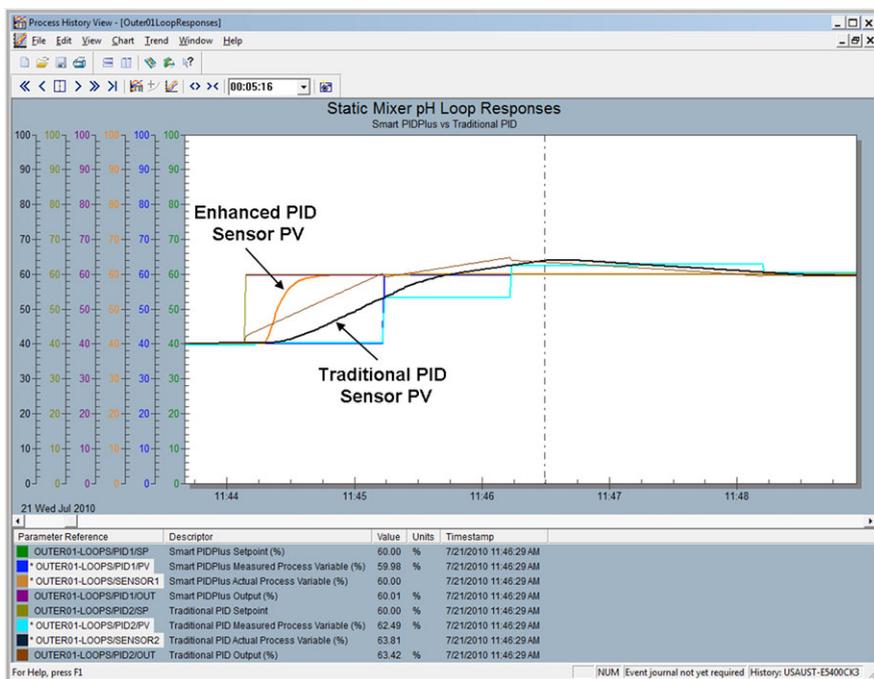
Fig. 14.9 Wireless PID with a default update rate of 60 seconds for pH and 16 seconds for flow provided exceptionally tight static mixer setpoint control

prolong battery life and valve packing life by increasing the time interval between communications. Larger wireless update time also reduces feedforward timing errors [26]. The deadtime introduced into the loop is on the average about half of the update time since the measurement result is at the beginning of the time interval [6, 30–32, 34]. The additional deadtime increases the ultimate limit to loop performance for unmeasured disturbance per (14.3) and (14.4).

14.2.14 Large Analyzer Cycle Times

A large analyzer cycle time offers a similar opportunity for the use of a wireless PID. The primary difference is that the analyzer cycle time is typically much larger than the wireless update time, which means the wireless PID is beneficial for much slower processes. For an analyzer cycle time larger than the 63% process response time (without the analyzer delay), the use of the enhanced PID provides excellent setpoint control by allowing the use of a PID gain that is the inverse of the open loop gain. The enhanced PID is stable without retuning when the cycle time is increased due to the addition of sample points, decrease in sample flow rate, or an increase in chromatograph column length. The deadtime introduced into the loop is on the

average about 1.5 times the cycle time since the measurement result is at the end of the time interval [6, 32]. The additional deadtime increases the ultimate limit to performance for unmeasured disturbance per (14.3) and (14.4). In the enhanced PID setpoint response, the measured PV has the analyzer delay but the actual process does not have the delay. Thus, if the controller gain is the inverse of open loop gain, the rise time does not depend upon the additional delay introduced by the analyzer.

14.2.15 High Process Nonlinearity

The process gain and deadtime for inline blend composition, heat exchanger, vessel coil temperature, and jacket temperature control from the material and energy balance is inversely proportional to flow. For pH and conductivity, there is also a nonlinearity seen as the slope in the plot of pH and conductivity versus acid or base concentration. For column temperature, there is a nonlinearity that is the slope in the plot of temperature versus the manipulated flow to feed flow ratio. Finally, there is the gain nonlinearity from installed characteristic of the control valve. If the curves are known, the nonlinearity of pH, conductivity, and column temperature are best handled by a PID input signal characterization and the nonlinearity of the valve characteristic by PID output signal characterization. An adaptive controller is still needed to correct for the changes in the curves with operating conditions such as feed composition for pH and conductivity and pressure for the valve. For inline control, the tuning settings would be scheduled based on total flow. For split ranged control valves, there can be a huge difference in process dynamics if different valves and different types of streams are manipulated. In all cases, an adaptive controller that identifies the process dynamics and the scheduling of tuning as a function of a user selected variable is critical. The scheduling provides preemptive adjustment of the tuning based on recent or best test results.

14.2.16 High Process Deadtime

The total loop deadtime sets the ultimate limit to performance as shown in (14.3) and (14.4). The sources of process deadtime are transportation delays, mixing delays, dissolution times, reaction times, thermal lags, and volumes in series (e.g., column trays). The sources of automation system deadtime are threshold sensitivity and resolution limits and backlash for a ramping signal, actuator lags, sensor lags, transmitter damping, analyzer cycle times, signal filters, module execution time, and digital update time. Except for digital delays and signal lags, the deadtime is variable. A simple addition of a deadtime block in the external reset limit path and the enabling of the dynamic reset limit option provide deadtime compensation with performance equivalent to a Smith Predictor. The improvement is only seen if the controller tuning is made faster. The integral time can be decreased to half of the total loop deadtime if the total loop deadtime is accurately set. The improvement and

robustness is greatest for a loop where the process time constant is greater than the total loop deadtime, which is counter to common belief that deadtime compensators are most useful for deadtime dominant loops [23, 24]; see Sect. 14.2.6 on deadtime compensators for more detail.

The best option for dealing with high process deadtime is feedforward control with an accurate measurement of the disturbance and an accurate gain and timing for the feedforward correction.

14.2.17 Feedforward Timing Errors

If the feedforward correction does not arrive at the same time and the same place in the process as the disturbance, an irregular, oscillatory, and confusing response can occur. If the feedforward signal arrives too soon, an inverse response can occur. If the feedforward signal arrives too late, a second disturbance is created. A wireless measurement can help prevent the inverse response and second disturbance if the elapsed time of the wireless is reset upon activation of the feedforward and the default update rate is longer than the time to steady state. However, the additional delay of the wireless update rate could cause instability unless the traditional PID is detuned. An enhanced PID for wireless will prevent the instability without retuning regardless of how much the default update rate is increased to extend wireless battery life or ride out feedforward timing errors.

14.2.18 Split Range Nonlinearities and Discontinuities

The stick-slip is greatest, the change in the slope of the installed characteristic is greatest, and the change in process dynamics is most abrupt at the split range point. The split range can be between multiple secondary loops in cascade control or between multiple final control elements for single loop control. The split ranged final control elements are typically control valves but can be variable speed drives and heaters. When the demand for either final control element is minimal, there tends to be continual oscillation across the split range point. The oscillations can be a limit cycle from resolution or threshold sensitivity limits or from tuning settings that are too fast for the drastic change in the valve gain and in process dynamics. These oscillations introduce variability into the process and wear out valve packing from excessive movement and internal flow elements (plugs, discs, seals, and seating surfaces) from high velocities. When opposing media are manipulated, such as heating and cooling for temperature control and acid and base reagents for pH control, the oscillations cause a loss in process efficiency from an increase in energy and raw material use. The selection of the best split range point, adaptive tuning, and the enhanced PID algorithm can eliminate these oscillations.

The split range point is traditionally set at 50%. To compensate for the secondary loop or final control element and process gain nonlinearity from split range control,

the split range point should be set per (14.21a) instead of 50% so that the ratio of the controller output spans for each split ranged secondary loop or final control element is equal to the ratio of the product of the process and secondary or element gains. If you consider the useable range of the PID output to be 0–100% and a gap (G) could be set at the split range point, you also have (14.21b) for the first span (S_1). If you solve (14.21b) for the second span (S_2) and substitute the result into (14.21a), you end up with (14.21c) for the first span in terms of element and process gains [7]. Since these gains change with operating conditions and PID tuning also depends upon the open loop time constant and deadtime, adaptive tuning is still beneficial. The subscript "v" for the final control element gain is used because most split ranged outputs go to valves.

$$S_{x1} = \frac{K_{v1}K_{p1}}{K_{v2}K_{p2}} S_{x2}, \tag{14.21a}$$

$$S_{x1} = 100 - G - S_{x2}, \tag{14.21b}$$

$$S_{x1} = \frac{K_{v1}K_{p1}}{K_{v1}K_{p1} + K_{v2}K_{p2}}(100 - G). \tag{14.21c}$$

Adaptive tuning can identify the changes in the final element and process dynamics. The tuning settings can be scheduled based on controller output with the regions more closely spaced around the split range point to deal with the discontinuities. If the best split range point is selected, the identification task is easier and the time to get a high quality model is shorter.

The gap at the split range point is a method to prevent temporary inadvertent manipulation of the other final control element. However, integral action in the PID or in the process will eventually cause a transition to the other element. Consequently, the benefit is mostly in terms of preventing noise from causing an unnecessary crossing of the split range point. The consequences of noise are better minimized by transmitter damping or signal filtering. For wireless measurements, attenuation of noise by transmitter damping is preferred to prevent unnecessary updates decreasing battery life and reducing the enhanced PID performance.

Integral deadband is used to help prevent limit cycles and to slow down the transition between split ranged valves. However, integral deadband slows down the response to an upset and complicates the tuning. User-selection of the integral deadband is difficult because high variability and nonlinearity of dynamics and stick-slip is the norm at the split range point. The enhanced PID requires no adjustment to prevent limit cycles and unnecessary transitions.

The PID enhanced with a wireless measurement has the patience to wait out short term transitions and discontinuities. If the damping in the transmitter and the threshold sensitivity for an update are set to prevent transmission of noise, the enhanced PID accomplishes the objective of the split range gap and integral deadband. The ability to deal with discontinuities is especially significant since there are no preformulated PID parameters or options. For example, a measurement default update rate of 30 seconds for jacket temperature can reduce unnecessary excursions across the split range point and overreaction to steam shock in addition to improving the

loop's setpoint response for vessel temperature cascade control. If the default update rate is larger than the process response time, the PID tuning becomes independent of process deadtime and time constant. The PID gain can be simply set equal to the inverse of the open loop gain (K_o).

Split range control has often been used to achieve greater control valve range-ability by throttling a small valve in parallel with a large valve when the demand for flow is low. For a low flows, the small valve starts to open and the big valve freezes at minimum position to prevent riding the seat of the big valve which is the point of greatest stick-slip and wear. However, unless the big valve threshold sensitivity, resolution, and backlash in percent of stroke is much less than the small valve, the limit cycles at high flow demand are much larger. A better solution is to eliminate split range control and use instead valve position control where the process PID manipulates the small valve and a valve position control (VPC) PID manipulates the big valve. Traditionally, integral only control is used in the VPC with an integral time that is $10\times$ the product of the process PID gain and reset time to eliminate interaction. However, this action is sometimes too slow for large disturbances. For measured load upsets, feedforward control should be added. Here the feedforward signals for the process PID and VPC PID would be intelligently limited to keep the small valve in a good a throttle range and avoid unnecessarily positioning of the big valve with its larger stick-slip.

The enhanced PID with a large update trigger level for the small valve position can be used for VPC instead of an integral only controller. The enhanced PID for VPC suppresses limit cycles from stick-slip and backlash and reduces interaction with the process PID while offering more aggressive action for large disturbances.

The crossing back and forth and discontinuity of the split range control can also be reduced for opposing media by the temporary transition to VPC instead of completely closing the valve not being throttled by the process PID. In this case, the VPC provides a smooth transition by a small opening of the valve normally shut by split range control in order to keep the valve manipulated by the process PID from dropping below some minimum throttle position. There is a small temporary waste of energy or raw materials while the opposing media valves are both open but this may be less than the loss of efficiency from the crossing back and forth and discontinuity of the split range point. The enhanced PID can be used for this VPC to reduce interactions and provide more aggressive tuning.

Valve position control is also used for the prioritization of the manipulation of multiple flows (multiple PID outputs) and for decreasing energy use and increasing production rate. Here again, the enhanced PID is beneficial.

14.2.19 Multiple Inputs and Outputs

When there are several process variables (PID inputs) with limits that need to be honored, a PID for each constraint is setup for override control. Each override PID has a process variable that needs to be constrained as its controlled variable. The

output of the multiple PID goes to a signal selector that determines which output is used to manipulate a secondary loop setpoint or final control element. While we think of override control as a multiple input single output application, the output could be split ranged between multiple secondary loops or elements. The split range block can provide simultaneous besides the customary sequential manipulation of secondary loops or elements.

The principle problem with override control is the tuning, windup, and timely transition of the PID controllers. Each PID must be tuned to account for the different process and measurement dynamics of the controlled variables. The PID algorithm must not windup when not selected and must smoothly takeover control when controlled variable is above and below its setpoint for a high and low limit respectively. Implementation of the integral mode by literally integrating the error between the setpoint and controlled variable has caused the output of unselected controllers to walk off to low and high outputs despite the use of special options such as integral tracking. The positive feedback implementation of the integral mode inherently prevents this problem. Also, the use of the process variable of the secondary loop or the readback of valve position or speed for the final control element provides a smooth transition and prevents the override controller output from changing faster than the secondary loop or final control element can respond. If the proper external reset signal is used and the back calculate output of each function block is properly connected to the back calculate input of the upstream block, the most common cause of a disruptive transitions is the loss of status and continuity in the back calculate path. The loss is typically caused by the introduction of special logic by the user to try and accomplish what is automatically achieved by the proper setup of the override control system. Any special application requirements should be implemented via standard options in the standard PID, "Signal Selector", and "Splitter" blocks rather than by the introduction of calculations or sequences.

Users may desire that the override PID be selected when the process variables crosses the override PID setpoint. Since there is deadtime in all loops just by virtue of the delays and lags in the automation system (final control element, measurement, and controller), the process variable will overshoot the setpoint if correction is delayed to crossing the setpoint. The override setpoint in theses cases will need to be further away from the safety instrumentation system (SIS) trip or relief device setting. A way to eliminate the uncertainty of the setpoint bias needed to prevent activation if the SIS or relief device is to tune the override PID for a negligible overshoot by making the proportional mode contribution greater than the integral mode contribution in the PV approach to the override setpoint.

There may be multiple PID outputs because of alternate resources of fuel, utilities, raw materials, and reagent. Often, one of the resources is more abundant or less expensive. A common example is waste or recycled streams versus purchased fuel, heat, steam, reagent, and reactants. Valve position control (VPC) is used to maximize the use of the more abundant or lower cost resource by increasing its flow until the purchased resource manipulated by the process PID is at a minimum throttle position for minimum stick-slip and maximum valve life. As with other VPC applications, an enhanced PID eliminates limit cycles and reduces interaction and valve wear.

14.2.20 Energy Use and Production Rate Optimization

Valve position control (VPC) is used optimize energy use by decreasing the pressure of boilers and compressors and increasing the temperature of refrigeration units and cooling towers by working the furthest open user valve towards its maximum throttle position. The controlled variable for the VPC is the high signal selector output of user valve positions. The VPC PID output is the pressure or temperature setpoint of the utility.

Similarly VPC can be used to optimize production rate by increasing feed rate to columns, crystallizers, evaporators, and reactors by working the furthest open utility and feed valve towards its maximum throttle position. The controlled variable for the VPC is the high signal selector output of utility or feed valve positions. The VPC PID output is the feed flow setpoint.

To minimize interactions an enhanced PID can be used with a threshold sensitivity setting large enough to ignore insignificant changes in user, utility, or feed valve position that is the output of the high signal selector.

14.2.21 Process Interactions

The two predominant methods of reducing PID interaction are separation of the dynamics and decoupling. The separation of dynamics is accomplished by tuning the PID so that the closed loop time constant (Lambda) of the fighting PID loops are a factor of 10 different. The faster loop is made faster by a higher PID gain (smaller Lambda) and, if necessary, the slower loop is made slower by a lower PID gain (larger Lambda). As the Lambda is successively increased, integral only control is approached. If the loops have similar speeds, the most important loop is made faster. The enhanced PID and wireless measurements can be used to reduce interaction. The wireless trigger level setting is larger for the less important PID.

Decoupling can be achieved by adding a feedforward signal of one PID output to the other PID output. Half decoupling, the use of a feedforward for just one of two interacting PID, is sufficient to break interaction. The PID output from the least important loop is applied as feedforward to the more important loop. If there is no significant difference in the relative importance of the PID, then the feedforward is applied to the PID loop with the greatest deadtime.

14.2.22 Communication and Component Failures

Measurements or final control elements can fail to update because there is a loss of communication or a component failure to last value. Communication failures can occur in bus systems due to link failures and in wireless systems due to low battery power. Sensor failures to last value occur for cracked or coated pH glass

electrodes and plugged DP impulse lines. Final element failure to the last value occurs when a control valve does not move due to excessive stiction from high temperature, binding or occlusion of the internal control element (e.g., seat and plug) from the build up of solids and coatings, or the last position air failure of piston actuators.

When a failure to update occurs, a conventional PID will continue to ramp its output based on the inevitable error between the controlled variable and the setpoint. If the PID error is large or the duration of the failure is long, the output will hit its output limit. When the failure is corrected and an update occurs, the traditional PID considers entire change in measurement to have occurred in the last execution of the PID. The result is a large step in the output from proportional action and a bump from derivative action.

The enhanced PID suspends integral action when there is no update. As a result the PID output stays at the last value before the failure which is the least disruptive action to the process. When the measurement or valve responds or the communication is restored, the enhanced PID makes a small proportional and derivative mode and gradual integral mode correction based on difference between the current and the last known value. The enhanced PID is able to ride out update failures without overreaction.

14.2.23 Batch Profile Optimization

Batch columns, crystallizers, evaporator, and reactors have a key concentration and or temperature endpoint. Temperature endpoints are measured online but concentration endpoints are often based on lab analysis and thus depend upon repeatable conditions. The path of the key measurement from start to finish of the batch crystallization, reaction, or separation process is a profile. As with most optimization opportunities, the first objective in batch control is to make the profile repeatable so that the batch cycle time and endpoint are consistent and differences can be tracked down to changes in raw material inputs, utility systems, instrumentation, or operator actions. Since the achievement of a repeatable batch is an important step that should not be bypassed, the use of the best batch, "Golden Batch", as the objective instead of matching the average of a representative set of batches is counter productive [4]. If the batch profile is controlled, variability is transferred from the profile to manipulated process inputs (flows). The result is a repeatable profile than can be optimized and a narrowing of the problem to analysis of the changes in flow. The process knowledge gained from data analytics can be used to optimize the profile.

The measurement, control and optimization of the profile by a PID can provide advanced batch control. However, since the actual and desired profile response is only in one direction, typically an increase in production concentration, the standard PID algorithm cannot make corrections in both directions. A flattening of the profile and a possible windup from integral action can occur if the setpoint profile outpaces the actual profile. If the rate of change of temperature or concentration is used as

the controlled variable, the PID response is in both directions and smooth. The set point is now the slope of the optimum profile [4, 16].

For batch profile slope control, there is a translation of controller tuning settings from the use of measurement rate of change as the controlled variable. Proportional action is now derivative action, integral action is now proportional action, and derivative action is acceleration or second derivative action. Often the optimization of the profile simply consists of maximizing the slope [4, 16]. The maximum slope corresponds to the maximum biological product formation rate, crystallization rate, evaporation rate, chemical reaction rate, and separation rate.

14.2.24 Plantwide Feedforward Control

A plant can go to a new production rate within minutes by plantwide feedforward control. If all of the key flows (feed, utility, recycle, and vent flows) are set to the values on the Process Flow Diagram (PFD), the plant can move to a new production rate without waiting on composition, level, pressure, pH, and temperature loops to respond.

The PFD values may not be a simple ratio of the new to old production rate although this could be a good starting point. Better would be sets of flows captured at different operating rates and ambient conditions (e.g., summer versus winter operation) for the flow feedforwards. Process PID loops would as usual trim through feedback correction the feedforwards. The concept can be expanded to product grade changes and fed-batch operation for faster and more adaptive transitions and batches.

If a plant can respond quickly to changes in raw material supply and market demands, the inventories can be reduced. If a plant can increase rates when utility costs are low (e.g., off peak time rates) and ambient temperatures are low for cooling water, the energy efficiency can be increased.

14.2.25 Impact of Threshold Sensitivity and Noise on Performance

The equations presented in Sect. 14.2.1 on loop performance can be expanded to include the effect of wireless and valve threshold sensitivity and measurement noise.

For a step change in an unmeasured disturbance that would cause an open loop error of (E_o), the minimum peak error (E_x) (14.22) is proportional to the ratio of the total loop deadtime to the 63% response time (T_{63}). The minimum integrated error (E_i) (14.23) is approximately the peak error multiplied by the total loop deadtime that is the original deadtime (θ_o) plus the additional deadtime from a wireless measurement (θ_w) and a control valve (θ_v). The 63% response time (14.24) is simply the sum of the original loop deadtime plus the wireless measurement deadtime and the open loop time constant (τ_o), which is hopefully the primary process time constant.

Note that the valve deadtime from stiction and backlash does not appear in the 63% open loop step response time because the step change in controller output must be larger than the valve resolution, threshold sensitivity, and deadband to create a response. Consequently, on open loop step response trend charts, the deadtime from valve stiction and backlash is not seen. However, in closed loop operation, the controller output is ramping for an unmeasured disturbance and the time the controller takes to ramp through the valve resolution, threshold sensitivity, and deadband is additional deadtime

$$E_x = \frac{\theta_o + \theta_w + \theta_v}{T_{63}} E_o, \tag{14.22}$$

$$E_i = \frac{(\theta_o + \theta_w + \theta_v)^2}{T_{63}} E_o, \tag{14.23}$$

$$T_{63} = \theta_o + \theta_w + \tau_o. \tag{14.24}$$

The additional deadtime from a wireless measurement (14.25) is the smallest of the deadtimes from the wireless "default update rate" (update time interval) for periodic reporting (ΔT_w), and the wireless "trigger level" (threshold sensitivity) for exception reporting (S_w). The deadtime from periodic reporting ($\theta_{\Delta T}$) (14.26) is one half of the update time interval. The deadtime from exception reporting (θ_S) (14.27a) is one half of the threshold sensitivity setting divided by the maximum rate of change of the % process variable ($\Delta PV/\Delta t)_{max}$. Half of the interval and threshold sensitivity settings are used because on the average the disturbance starts halfway in the interval and halfway in the sensitivity.

The rate of change is a maximum during the beginning of the disturbance before the control loop has had any effect. This maximum rate of change is dictated by the size of the disturbance and process dynamics. The maximum ramp rate (14.27b) can be approximated by a near integrator gain multiplied by the equivalent change in controller for the disturbance. The equivalent change in controller output is the open loop error (E_o) divided by the open loop gain (K_o). The near integrator gain (14.27c) is the open loop gain divided by the open loop time constant (τ_o). The substitution of (14.27c) into (14.27b) cancels out the open loop gains giving the maximum ramp rate (14.27d) as simply the open loop error divided by the open loop time constant. The substitution of (14.27d) into (14.25) yields (14.27e) where we see that the deadtime from the threshold sensitivity setting decreases as the size of the disturbance increases and the open loop time constant decreases, both of which cause a faster rate of change

$$\theta_w = \min(\theta_{\Delta T}, \theta_S), \tag{14.25}$$

$$\theta_{\Delta T} = 0.5 \Delta T_w, \tag{14.26}$$

$$\theta_S = \frac{0.5 S_m}{(\Delta PV/\Delta t)_{max}}, \tag{14.27a}$$

$$(\Delta PV/\Delta t)_{max} = K_i (E_o/K_o), \tag{14.27b}$$

$$K_i = \frac{K_o}{\tau_o}, \tag{14.27c}$$

$$(\Delta PV/\Delta t)_{max} = \frac{E_o}{\tau_o}, \tag{14.27d}$$

$$\theta_S = \frac{0.5 S_m \tau_o}{E_o}. \tag{14.27e}$$

Similarly, the deadtime from valve threshold sensitivity (S_v) can be estimated (14.28a) as half of the sensitivity divided by the maximum rate of change of the controller output that is initially mostly due to the proportional mode. This rate of change of controller output (14.28b) is the rate of change of the process variable multiplied by the controller gain (K_c). The controller gain (14.28c) can be approximated as the Ziegler–Nichols gain multiplied by a detuning factor (K_x) to provide robustness and a smoother response. However, the gain should not inflict disturbances from fluctuations in the PID output from exceeding the control valve threshold sensitivity. An enhanced PID noise band expressed as measurement threshold sensitivity (S_m) can reduce this noise allowing a higher controller gain. The substitution of (14.28c) into (14.28b) cancels out the open loop time constant (14.28d). The substitution of (14.28d) into (14.28a) yields (14.28e) where the deadtime is proportional to the product of the open loop gain and original deadtime and inversely proportional to the product of the detuning factor and open loop error

$$\theta_v = \frac{0.5 S_v}{(\Delta CO/\Delta t)_{max}}, \tag{14.28a}$$

$$(\Delta CO/\Delta t)_{max} = K_c (\Delta PV/\Delta t)_{max}, \tag{14.28b}$$

$$K_c = \min\left[\frac{K_x \tau_o}{K_o \theta_o}, \frac{S_v}{\max[(N_m - S_m), 0.002]}\right], \tag{14.28c}$$

$$(\Delta CO/\Delta t)_{max} = \frac{K_x E_o}{K_o \theta_o}, \tag{14.28d}$$

$$\theta_v = \frac{0.5 S_v K_o \theta_o}{K_x E_o}. \tag{14.28e}$$

14.3 Unit Operation Examples

The process industry is quite diverse in terms of production methods. While each industry tends to see itself as unique, similarities emerge when viewed on a conceptual level. The solutions previously described offer a tool palette that addresses most of the industrial opportunities despite widely different processes. Examples of these solutions are offered for some of the most important unit operations in process industry. While not detailed, the use of the enhanced PID for valve position control can in general offer an easy, fast, and inexpensive increase in production rate of continuous and fed-batch unit operations.

14.3.1 Biological Reactors

Bioreactors are used for the production of most of new pharmaceuticals that are proteins too complex to be produced by chemical processes. Most of the older biological processes use genetically engineered fungal and bacterial cells. As biopharmaceutical proteins become more sophisticated, genetically engineered mammalian cells, such as Chinese hamster ovaries (CHO) cells are increasingly used. Biopharmaceutical batch processes are predominantly batch because of the concern for the buildup of toxins and genetically deficient cells with continuous processes and the need to get new drugs to market quickly.

The important loops for bioreactors are pH, temperature, and dissolved oxygen. In general, the allowable controller gain is high-limited only by the measurement noise since the ratio of deadtime to time constant is small for a well designed system. The high controller gain in combination with the extremely slow disturbances translates to exceptionally tight PID control. Most users do not see much of an advantage from improved tuning for well designed systems. The main concern with these systems is getting to temperature setpoint as fast as possible with essentially no overshoot at the start of the batch. When this fast start is important, the fast setpoint response techniques such as smart bang-bang control can be used [13]. If there is no concern about batch cycle time and low temperatures do not cause seed cell degradation, a PID structure of integral mode on error and proportional and derivative modes on PV is used to eliminate overshoot [24].

Often the automation and bioreactor system design is less than ideal. If the piping and sparger for the injection of oxygen or air for dissolved oxygen control and addition of carbon dioxide for pH control are not completely separated, the dissolved oxygen and pH loops interact. In this case, a half decoupler and an enhanced PID can reduce the interaction. Often variable frequency drives create electromagnetic interference and ground potentials that cause spikes in pH measurements that disrupt the dissolved oxygen and pH loops. Wireless pH transmitters have been found to eliminate the spikes commonly seen in bioreactor loops [28].

Many bioreactor loops do not use sliding stem (globe) control stem valves because they are not suitable for sanitary and sterilization-in-place (SIP) service. The alternative final control elements used may have a poor resolution or sensitivity limit and backlash. The enhanced PID can eliminate the associated limit cycles.

There remains a largely unrealized opportunity for PID control of substrates, such as glucose. Online analyzers can provide glucose concentration measurements every 4 to 6 hours depending upon the number of bioreactors serviced by the analyzer and the auto sampler system design. For a bioreactor batch that takes 1 to 2 weeks, this sample time is fast enough to do closed loop control if an enhanced PID is used to deal with the analyzer cycle time and resolution limits.

Total cell concentration can be measured by turbidity sensors. The same analyzer used for glucose may be able to provide cell count and diameter and the concentration of precursors to cell death. Smart dielectric measurements can indicate what fraction of the cells still have their membranes intact and thus are viable cells for product formation. Consequently, the viable cell concentration profile can be controlled and optimized by the use of an enhanced PID with a controlled variable that

is the rate of change of viable cell concentration and whose output is the glucose concentration and pH setpoints. The manipulation of the pH setpoint presents an additional challenge because of the narrow peak in the plot of cell growth and death rate versus pH and the reversal of process gain. The pH optimization can be limited to one side of the peak by the online computation of an inferential measurement of the change of viable cells per change in pH. It is conceivable that this inferential measurement (Δ cells/Δ pH) could be used as the controlled variable in a loop to eliminate the reversal of process sign problem. Here, the theoretical optimum PID setpoint of zero at the peak could be trimmed by the cell profile PID.

The online measurement of product concentration is more problematic, but given an inferential measurement is developed, online batch product concentration profile control is possible by the same type of PID used for viable cell concentration profile control. The optimum glucose and pH for product formation rate can be different than for optimum cell growth rate. The cell concentration profile PID may be used early in the batch during the pre-exponential and early exponential growth phases. The product concentration profile control would take over in the last half of the batch when product formation rate becomes significant.

14.3.2 Chemical Reactors

Chemical reactors set the stage for the production of bulk chemicals, intermediates, petrochemicals, polymers, pharmaceutical chemicals, and specialty chemicals. If the reaction is not producing the right product, not much else matters. Pharmaceutical and specialty chemicals use batch reactors while the other products predominantly use continuous reactors.

The Temperature PID is the most important controller since the reaction rate is often an exponential function of temperature via the Arrhenius Equation. The deadtime to time constant ratio for well mixed continuous reactors and the product of deadtime and integrating process gain for batch reactors are incredibly small (<0.001). The result is a permissible PID gain much larger than users are accustomed to (>50) and exceptionally tight temperature control. Since the most important task of the utility system is to satisfy the demands of the temperature PID, the transfer of variability by the PID from the reactor temperature to its utility system is maximized. The exception is where the utility system is used by a more important reactor in a more critical portion of its operation. The main limit to performance is the accuracy of the temperature sensor and the threshold sensitivity and resolution of the final control elements. Premium Resistant Temperature Detectors (RTD) and sliding stem valves with digital positioners should be used to allow the full capability of the PID to be realized [24–26].

Cascade control where the reactor temperature is the primary PID whose output is the setpoint of a jacket or coil temperature PID is used to linearize the reactor temperature loop and to provide fast compensation of disturbances to the heat transfer media [12]. Since the reactor temperature PID gain is exceptionally high there is a

significant risk the output of this primary PID will change faster than the secondary PID can respond. For large disturbances or large setpoint changes, this cascade control system can burst into oscillations. The use of the dynamic reset limit in the positive feedback implementation of integral action will prevent this problem. The external reset signal is the controlled variable of the secondary PID (jacket or coil temperature). For final control elements that are slower than the secondary loop PID, which can occur for large control valves or variable frequency drives with speed rate of change limits, the dynamic reset limit is extended to the secondary loop with an external reset signal of valve position or speed.

The reactant feed PID controllers are important to keep the reactants in the optimum ratio since the reaction rate and completeness depends on the concentration of reactants. Coriolis meters with their exceptionally accurate mass flow and density measurement can be used to provide a component mass balance and inferential measurement of feed composition, respectively [24–26]. A PID that has the optimum ratio of component mass as it controlled variable can slowly trim the ratio of reactants. Alternately, for exothermic reactions, a PID whose controlled variable is cooling rate and thus reaction rate can be used to trim the feed ratio.

If the utility systems are a constraint to increasing production rate, override controllers with integral only action or the enhanced PID algorithm can be used as valve position controllers (VPC). The controlled variable of the VPC is the utility system PID output, and the VPC setpoint is the maximum throttle position of the utility valve or maximum speed of a variable frequency pump. The output of each VPC goes to a low signal high signal selector whose output becomes the ratioed setpoint for the feed PID. These feed PID should be tuned with the same closed loop time constant (Lambda) so that changes in feed rate are coordinated and the total component masses in the reactor are maintained in the desired ratio.

The temperature and concentration profile of a batch reactor can be controlled by the use of a PID whose controlled variable is the rate of change of temperature or concentration with time as discussed for biological reactors [16]. However, the prevention of too fast of an increase in reaction rate for extremely exothermic reactors is more critical than in biological reactors because the exponential increase in reaction rate can through positive feedback create a runaway response. For runaway reactors, there is a window of PID gains where too low of a gain can result in a runaway besides the more familiar case of too high of a gain causing a growing oscillation. Since the permissible gain is much higher than most users realize, the more common mistake is a PID gain dangerously close to the low gain limit. Maximizing derivative action is also important to prevent a runaway. The PID rate time should be set equal to the sum of the thermowell and heat transfer surface lags. If the controlled variable is rate of change of temperature, then the integral mode is effectively the proportional mode and the proportional mode is effectively the derivative mode in terms of the temperature response. Thus, there is no integral mode to force the temperature to a specific temperature value, which is consistent with the objective of the progression of the reaction rate in the batch being most important. However, there is an important consideration in the setting of the integral time for runaway reactions. If temperature rather temperature rate of change is

used as the controlled variable, the reset time should be increased by a factor of 10 to prevent a runaway because of the extreme sensitivity of a process with positive feedback to the integral mode. Some severely exothermic polymerization reactors use proportional-derivative PD controllers (structure 4) for reactor temperature control because integral action has no sense of direction and promotes overshoot by delaying cooling until the temperature has crossed setpoint [24].

14.3.3 Crystallizers

Batch crystallizers are an important unit operation in pharmaceutical and specialty chemical processes. To minimize the formation of small seed crystals that can result in excessive fines and coating (frosting) in the beginning of the batch and to promote crystal growth and size toward the end of the batch, an optimum cooling curve and consequently temperature profile is used [27]. In the early part of the batch, the batch temperature is slowly lowered. The decrease in batch temperature is accelerated as the batch progresses to provide an exponential increase in cooling that is maximum at the near the end of the batch. Since crystallization releases heat, the PID can increase besides decrease the temperature and a temperature profile can be the setpoint for a PID whose controlled variable is crystallizer temperature. The use of temperature rate of change with time as the controlled variable is not necessary but might reduce batch time and prevent flattened sections or bumps in the temperature profile by emphasizing a continual decrease in temperature.

14.3.4 Distillation Columns

Column temperature at a given pressure provides an inferential measurement of column composition. The location in the column chosen for temperature control is the point where the change in temperature is the largest and most symmetrical for an increase and decrease in the flow manipulated by the temperature PID. The most effective scheme is the manipulation of distillate flow if this flow is not too small and the overhead receiver cross-sectional area is not too large. By manipulation of distillate flow, the reflux flow manipulated by receiver level control provides a degree of internal reflux control by changes in overhead vapor and hence reflux for changes in column loading and wall temperature. However, this scheme depends upon a change in distillate flow immediately translating to an appreciable change in reflux flow. If the overhead receiver area is too large, the change in level may be within the noise band or below the threshold sensitivity limit of the level measurement. The receiver level controller must have a very high gain. Normally, the product of deadtime and integrating process gain is so small the only limit to setting the overhead receiver level controller gain is measurement noise. However, users may not realize this and be close to the low gain limit that causes slow rolling oscillations besides poor temperature control. The oscillations may be incorrectly attributed to high of a PID gain

and the PID gain may be decreased making the problem worse. If the distillate flow is too small (the case where a small amount of impurity is driven off the top), the temperature manipulates the steam flow or the bottoms flow. If the bottoms flow is manipulated, a sump level PID manipulates the steam flow. In this scheme, level control is poor from an inverse response in the sump level due to shrink from the collapse of bubbles and the swell from the formation of bubbles [27].

The deadtime to time constant ratio for column temperature loops is typically about 0.1 because the largest source of deadtime is the equivalent delay that results from a large number of interacting time constants from trays in series. The maximum temperature controller gain is about 4 times the inverse of the open loop gain. Typical temperature controller gains range from 2 to 20 for narrow and wide span temperature ranges, respectively, if the control point exhibits reasonable sensitivity of temperature to a change in composition. While the deadtime to time constant ratio is good for control, the absolute magnitude of the deadtime is large for tall columns (columns with a large number of actual or equivalent trays) [27].

The use of feedforward control for temperature and level control of columns has proven to be beneficial because of the large deadtime value and the moderate controllers gains used. The most frequent disturbance is a feed change. Flow feedforward control is used where the reflux, distillate, steam, and bottoms flow setpoints are ratioed to the feed flow and corrected by a feedforward summer in the temperature and level PID. The timing is complex. The corrective action of the steam or reflux flow must arrive at the temperature control point at the same time as the feed disturbance. Dynamic compensation is normally used to improve the timing of the feedforward where a delay is added to a feedforward correction that arrives too soon and a lead is added to compensate for a lag in the feedforward correction that is greater than the lag in the disturbance. An enhanced PID could help prevent oscillations from improper feedforward timing and eliminate limit cycles from measurement and valve threshold sensitivity and resolution limits and backlash. Since the rate of change of temperature is extremely slow, the signal-to-noise ratio could be improved by the use of wireless measurements with a default update rate chosen to be slow enough to insure the change in signal is more reflective of a change in true temperature than resolution limits or noise. The use of portable wireless integral mounted temperature transmitters would also help finding the optimum temperature control point. Similarly, portable wireless integral mounted pressure transmitters would enable finding the trays that are flooding as column rates as pushed. As in the reactor feed maximization, valve position controllers for each limiting valve capacity could be used as override controllers whose largest output is used to push the column feed rate higher.

"Ratio" blocks are used to multiply the feed flow by the desired ratio. The "Ratio" block also has the actual ratio. The feedback correction by the temperature or level PID is done by a "Bias/Gain" block that acts a feedforward summer. The desired and actual ratio are displayed for the operator. For startup until column temperature and traffic has reached the normal operating range, the temperature PID is put in manual and the operator sets the desired ratio for pure flow ratio control.

14.3.5 Evaporators

In an evaporator, often a pressure PID manipulates the vapor flow out the top and a level controller manipulates the liquid discharge flow out the bottom. The steam flow sets the evaporation rate. The concentration of the product can be controlled by the use of a Coriolis meter to measure product density in a recirculation line or in the discharge line if there is always a discharge flow. The product density controller manipulates the steam flow. A "Ratio" block can be used to multiply the steam flow by the desired ratio of steam to feed flow. If a Coriolis meter is also used on the feed flow, the desired ratio of steam to feed flow can be computed from the actual feed density and the desired product density. The product density controller adds a feedforward correction to the output of the ratio block by means of a "Bias/Gain" block. For startup until evaporator temperature and vapor flow has reached the normal operating range, the density PID is put in manual and the operator sets the desired steam to feed ratio for pure flow ratio control.

14.3.6 Neutralizers

The pH control of neutralizers can be particularly challenging due to the extreme nonlinearity and sensitivity of the pH measurement as a result of the exponential relationship between pH and hydrogen ion activity. The changes in controller gain and the rangeability and threshold sensitivity of the final element needed are extraordinary. PID gain changes of 1000 to 1 and a final element rangeability requirement of 10,000 to 1 are possible with strong acid and base systems. The precision of the final element may determine the number of neutralization stages necessary to keep waste streams in compliance with environmental regulations. The limit cycle amplitude from threshold sensitivity and resolution limits can be extremely large due to amplification by the steep slope of the titration curve. The 7 pH value for a broken electrode or wire and the failure to last value of a coated electrode are insidious [7].

Signal linearization, adaptive tuning, split range control, valve position control (VPC), and the enhanced PID are solutions to be considered. Signal linearization can translate the controlled variable from pH to reagent demand per the titration curve. Adaptive tuning can correct for changes in the titration curve and process dynamics. Split ranged control can enable neutralization with both acids and bases. A VPC can adjust a large (coarse) valve to keep a small (fine) valve in a good throttle range for pH control. A VPC can also maximize the use of waste and low cost reagents. Portable wireless pH transmitters can optimize the control location and eliminate spikes from EMI. Finally, the enhanced PID can eliminate oscillations from split range point discontinuities and from valve backlash, threshold sensitivity, and resolution, reduce interactions between the small and large valves, prevent overreaction to pH electrode failures, and extend wireless battery life [21, 24, 28].

14.4 Conclusion

The PID has been the predominant method of feedback and feedforward control for the process industry since the inception of automation systems. The PID provides a consistent and convenient interface for the operator. The algorithm is optimal for unmeasured disturbances and unknown dynamics, the common case in manufacturing. Intelligence can be added to expand the capability through adaptive tuning, deadtime compensation, feedforward control, signal linearization, and smart bang-bang logic. Key capabilities, such as the "dynamic reset limit", "measurement threshold sensitivity setting", and an enhanced PID developed for wireless, can eliminate oscillations and improve efficiency (reduce raw materials, utilities, and waste), flexibility, operability, maintainability, profitability, and safety, that determines compliance and competitiveness. The innovative use of the developing and expanding capability of the PID is the key to sustainable manufacturing.

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