Consider IMC Tuning to Improve Controller Performance

Internal-model-control tuning rules have a number of advantages for maximizing PID controller performance. They minimize disturbance propagation and use only one tuning parameter. Plus, needed process models can be developed easily.

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THE VAST MAJORITY OF controllers in the chemical industry are PID controllers. There are many reasons for this, including their long history of proven operation, the fact that they are well understood by many industrial operational, technical, and maintenance individuals, and, in many applications, the fact that a properly designed and well-tuned PID controller meets or exceeds the control objectives. In addition, an industrial PID controller has many extensions that make it a practical tool for operating a chemical process. For example, it has automatic and manual switching, set-point tracking, external reset feedback, and emergency manual modes.

Many controller tuning methods have been proposed in the literature. Perhaps the most well known are the Ziegler–Nichols and the Cohen–Coon rules. Ziegler–Nichols uses critical gain and frequency information in order to determine the *P, I,* and/or *D* parameters based on quarter-decay criterion. Cohen–Coon utilizes a first-order-plus-deadtime model with quarter-decay criterion to determine the tuning parameters. Later extensions were made to higher-order models with a variety of performance criteria such as integral of the absolute value of the error, integral of the time-weighted absolute error, integral of the square error, maximum peak, and phase margin.

Rivera and co-workers *(1)* introduced a PID controller design method based on internal model control (IMC) in 1986 that is attractive to industrial users because it has only one tuning parameter. The parameter relates directly to the closed-loop speed of response and to the robustness of the control loop; the larger the tuning parameter, the greater the robustness. Moreover, the closed-loop load response exhibits no oscillation or overshoot. Experience indicates that this minimizes controller interactions and enhances overall process disturbance rejection.

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In this article, the IMC–PID tuning rules will be discussed in the context of industrial applications.

IMC–PID tuning method

The IMC–PID tuning method was first introduced by Rivera *et al.* *(1)* and was later extended to cover a wider range of process models by Chien *(2)*. The idea is to use straightforward, two-step IMC controller design to obtain conventional PID controller tuning constants. Figure 1 shows the typical IMC control structure. The simple G_{IMC} controller design procedure is as follows *(3)*.

Step 1. Factor the model

$$G_M = G_{M+}G_{M-} \quad (1)$$

such that G_{M+} contains all of the deadtimes and right half-plane zeros; consequently, G_{M-} is stable and does not involve predictors.

Step 2. Define the IMC controller as

$$G_{IMC} = G_{M-}^{-1}f \quad (2)$$

where f is a user-specified low-pass filter.

Figures 2 and 3 show how the IMC structure can be put into the conventional feedback control structure by canceling the two dashed-line paths involving G_M. The relationship between the feedback controller (G_C) and the IMC controller (G_{IMC}) is

$$G_C = G_{IMC}/(1 - G_{IMC}G_M) \quad (3)$$

Thus, by knowing the process model, the feedback controller (G_C) can be calculated by using Eqs. 1–3. The process model need not exactly match the process. The controller performs quite well even under severe model mismatch conditions *(2)*.

The IMC–PID tuning rules are derived for the common vendor implementations of the PID algorithm. The Laplace transforms of the algorithms are as follows. For PID(1),

$$G_C = K_C\left(1 + \frac{1}{\tau_I S}\right)\left(1 + \frac{\tau_D S}{(\tau_D/DG)S + 1}\right) \quad (4)$$

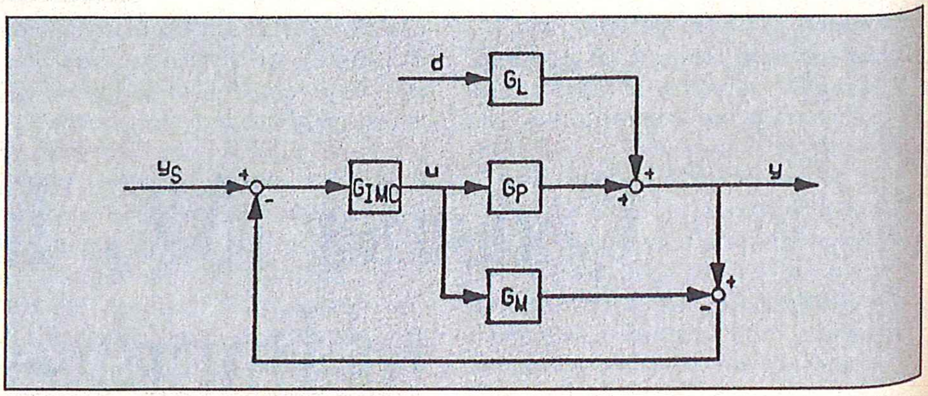

Figure 1. Internal model control structure.

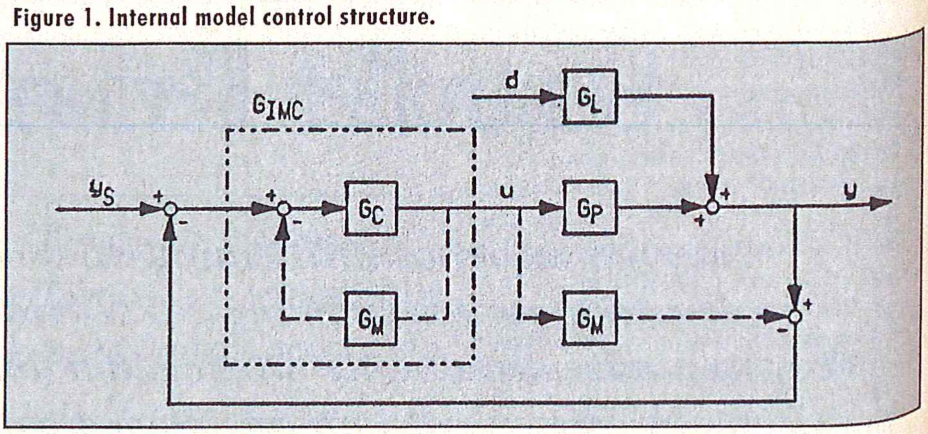

Figure 2. Relationship between IMC structure and feedback control structure.

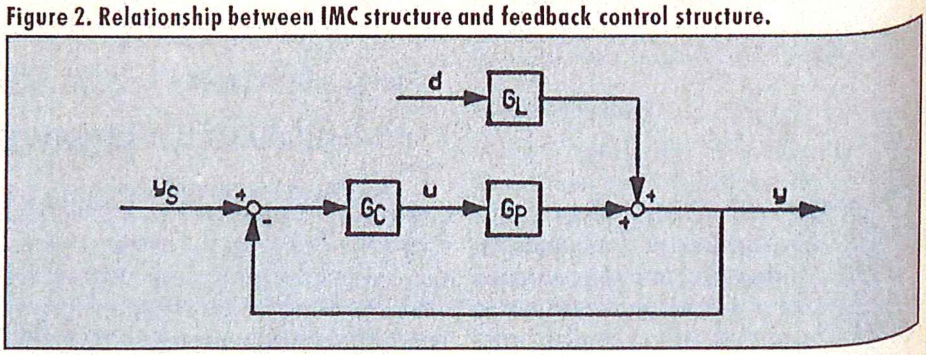

Figure 3. Feedback control structure.

For PID(2),

$$G_C = K_C\left(1 + \frac{1}{\tau_I S}\right)\left(\frac{1 + \tau_D S}{(\tau_D/DG)S + 1}\right) \quad (5)$$

For PID(3),

$$G_C = K_C\left(1 + \frac{1}{\tau_I S} + \frac{\tau_D S}{(\tau_D/DG)S + 1}\right) \quad (6)$$

where K_C is the controller proportional gain (dimensionless), τ_I is the reset time, τ_D is the derivative time, and DG is the derivative gain.

The procedure for obtaining PID tuning rules from the IMC controller design is as follows. First, approximate the deadtime by either a first-order Padé or a first-order Taylor series. Second, select f as

$$f = 1/(\tau_{CL}S + 1) \quad (7)$$

for first- and second-order processes or as

$$f = \frac{[2\tau_{CL} - G'_M + (0)]\; S + 1}{(\tau_{CL}\; S + 1)^2} \quad (8)$$

for processes with an integrator. Third, use Eqs. 1–3 to find the feedback controller, G_C. Finally, compare with Eqs. 4–6 and equate the coefficients to find the relationship between the PID parameters and model parameters.

Notice that, for these choices for f in the second step, there will be no

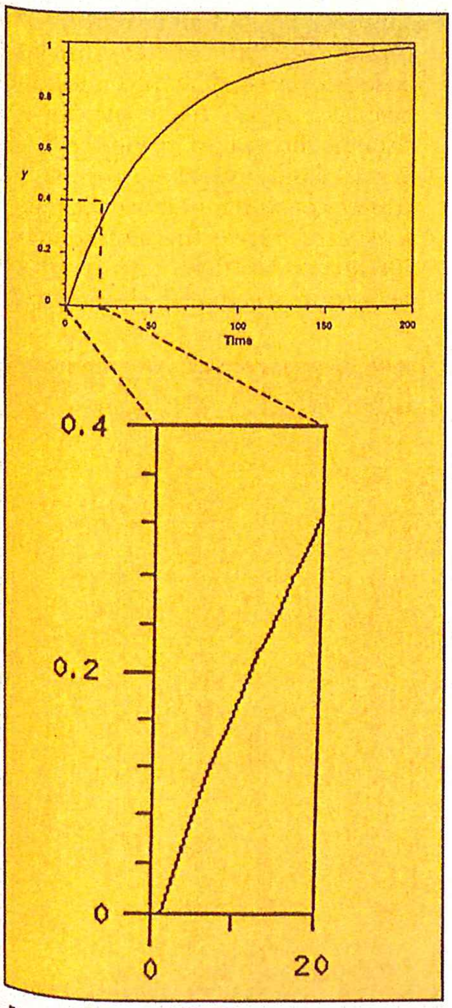

Figure 4. Open-loop unit step response of the process in Eq. 9.

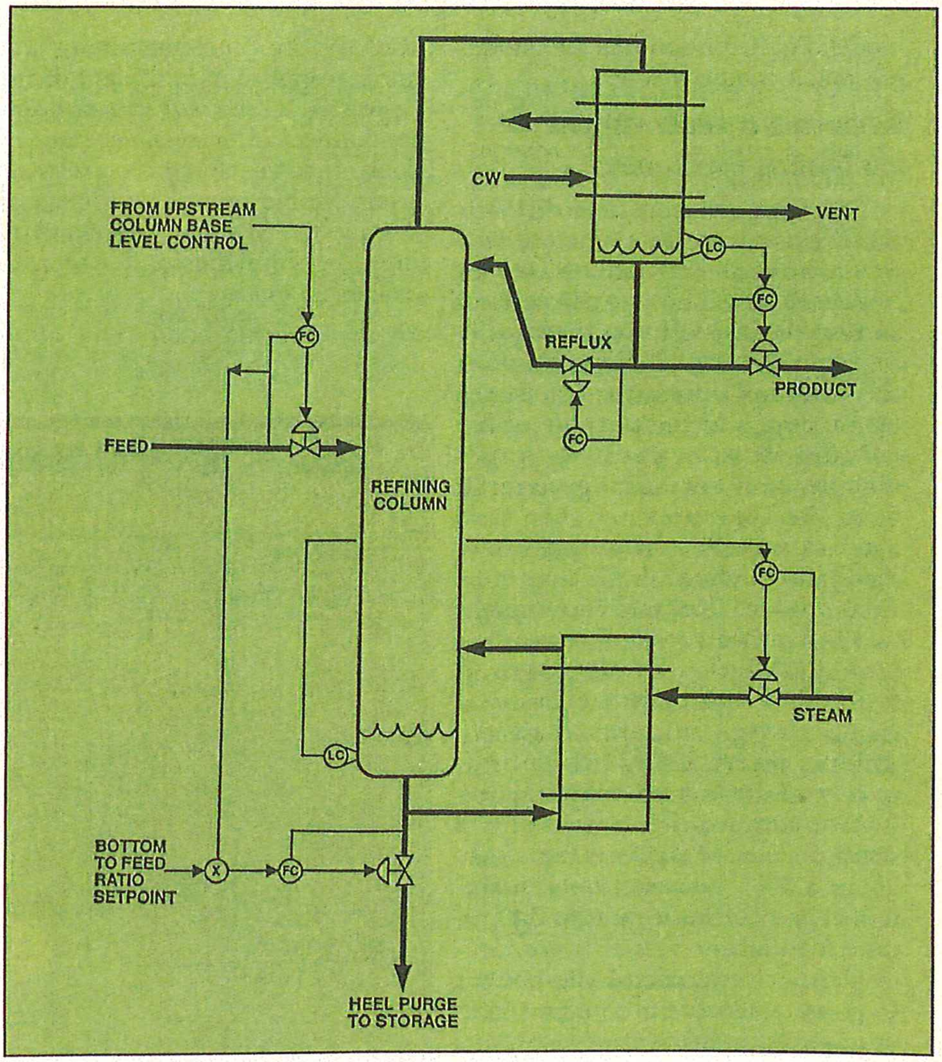

Figure 5. Strategy for distillation column bottom level control.

oscillation and overshoot for closed-loop load response. This characteristic makes this tuning method very appealing to industrial users. The PID tuning rules for commonly used process models in the chemical industry are given in Table 1. In Eqs. 7 and 8, τ_{CL} is a user-specified parameter that corresponds to closed-loop speed of response. A smaller τ_{CL} provides faster closed-loop response but the manipulated variable is moved more vigorously, while a larger τ_{CL} provides a slower but smoother response. A larger τ_{CL} is also less sensitive to model mismatch. Our experience shows that between the open-loop dominant time constant and the process deadtime is a good region from which to select τ_{CL}. Further refinement of τ_{CL} is trivial based on the properties of τ_{CL} described previously. For more detail on the IMC–PID tuning rules, see Chien *(2)*.

Time-constant-dominant processes

For processes with large time constants vs. deadtime, the IMC–PID tuning rules give slow load response because IMC design retains the open-loop load response into the closed loop. Thus, if the open-loop load response is slow, the closed-loop response will be slow no matter how small a τ_{CL} is chosen. One way to improve the tuning rules in this situation is to approximate the model as an integrator plus deadtime.

A numerical example will be used here to illustrate the idea. The Laplace transfer function of the example is

$$G_P = G_L = \frac{e^{-S}}{50S + 1} \quad (9)$$

Notice that the process and the disturbance dynamic are assumed to be the same, and the time constant is dominant with a deadtime-to-time-constant ratio of 0.02. Figure 4 shows the open-loop unit step response of this process. The bottom part of the figure represents the first 20 min of the overall dynamic response. The process dynamic initial response is very similar to the response of a process with an integrator and deadtime. For this example, the approximate model emphasized on the initial response is

$$G_M = \frac{0.018e^{-S}}{S} \quad (10)$$

Controller design based on this approximate model is intuitively sound because the closed-loop operation will always keep the process variable near the initial response operating region.

Simulation results show that tuning rules based on the integrator-plus-deadtime model, Eq. 10, gives much superior closed-loop load responses than the tuning rules based on the first-order-plus-deadtime model, Eq. 9, for nominal and model mismatch conditions.

Removing a small amount of low boiling component

The first example is a difficult control loop that uses the steam flow rate to control a distillation column bottom level. The overall control strategy for this column is shown in Figure 5. The distillation column is the final unit operation in a chemical process, and the purpose of the column is to separate a small amount of a low boiling material from the final product. The feed rate is controlled by an upstream distillation column base level. The final product flow rate is controlled by the overhead condenser receiver level. Reflux flow to the column is under flow control with a conservative high flow setpoint to ensure product purity. Using bottom flow to control the bottom level is not effective because this flow is only a small portion of the feed flow (only about 1.3%). Bottom flow is maintained at a constant ratio to the column feed rate.

The performance of the bottom level control loop was so poor that it required constant operator attention during start-up and normal operation. One of the authors (Chien) was asked to improve the control performance. Upon arrival at the plant, the control performance was investigated and found to be very cyclic as shown in Figure 6.

The process model for the level system, which was derived from a material balance equation, is:

$$\frac{K_p e^{-\theta S}}{S}$$

where K_p is the steam flow span times the steam latent heat divided by the latent heat of process fluid times the level span times the density times the cross-sectional area and θ is the process deadtime.

K_p is readily available from process information (in this case, K_p = 0.2/min), and θ usually can be obtained by an open-loop pulse test. In this case, the deadtime is alternatively obtained by using closed-loop simulation. The computer simulation used the same controller tuning constants as the plant and adjusted the process deadtime to fit the dynamic response of real plant data in Figure 6. The resulting dynamic simulation can be seen in Figure 7. Notice that the closed-loop response is very similar to real plant response (Figure 6). Because the deadtime found by simulation is 7.4 min, the process model is

$$\frac{0.2e^{-7.4S}}{S}$$

From Table 1, the PI tuning constants for τ_{CL} = 8 min are K_C = 0.49 and τ_I = 23 min. A simulated closed-loop response to a load disturbance using these tuning constants is illustrated in Figure 8. The actual plant response using the new tuning constants is shown in Figure 9. Again, notice the similarities in the responses. The benefits of this technique are that it eliminates the

Table 1. IMC–PID tuning rules.

	PID(1)			PID(2)	
Process Model	K_cK_p	τ_I	τ_D	K_cK_p	τ_I
$\frac{K_pe^{-\theta S}}{\tau S+1}$	$\frac{\tau}{\tau_{cl}+\theta}$	τ	–	$\frac{\tau}{\tau_{cl}+\theta}$	τ
$\frac{K_pe^{-\theta S}}{\tau S+1}$	$\frac{\tau}{\tau_{cl}+\frac{\theta}{2}}$	τ	$\frac{\theta}{2}$	$\frac{\tau}{\tau_{cl}+\frac{\theta}{2}}$	τ
$\frac{K_pe^{-\theta S}}{\tau S+1}$	$\frac{\frac{\theta}{2}}{\tau_{cl}+\frac{\theta}{2}}$	$\frac{\theta}{2}$	τ	$\frac{\frac{\theta}{2}}{\tau_{cl}+\frac{\theta}{2}}$	$\frac{\theta}{2}$
$\frac{K_p(\tau_3S+1)e^{-\theta S}}{(\tau_1S+1)(\tau_2S+1)}$	$\frac{\tau_1}{\tau_{cl}+\theta}$	τ_1	$\tau_2-\tau_3$	$\frac{\tau_1}{\tau_{cl}+\theta}$	τ_1
$\frac{K_p(\tau_3S+1)e^{-\theta S}}{(\tau_1S+1)(\tau_2S+1)}$	$\frac{\tau_2}{\tau_{cl}+\theta}$	τ_2	$\tau_1-\tau_3$	$\frac{\tau_2}{\tau_{cl}+\theta}$	τ_2
$\frac{K_p(\tau_3S+1)e^{-\theta S}}{\tau^2S^2+2\zeta\tau S+1}$	–	–	–	–	–
$\frac{K_p(-\tau_3S+1)e^{-\theta S}}{(\tau_1S+1)(\tau_2S+1)}$	$\frac{\tau_1}{\tau_{cl}+\tau_3+\theta}$	τ_1	$\tau_2+\frac{\tau_3\theta}{\tau_{cl}+\tau_3+\theta}$	$\frac{\tau_1}{\tau_{cl}+\tau_3+\theta}$	τ_1
$\frac{K_p(-\tau_3S+1)e^{-\theta S}}{(\tau_1S+1)(\tau_2S+1)}$	$\frac{\tau_2}{\tau_{cl}+\tau_3+\theta}$	τ_2	$\tau_1+\frac{\tau_3\theta}{\tau_{cl}+\tau_3+\theta}$	$\frac{\tau_2}{\tau_{cl}+\tau_3+\theta}$	τ_2
$\frac{K_p(-\tau_3S+1)e^{-\theta S}}{\tau^2S^2+2\zeta\tau S+1}$	–	–	–	–	–
$\frac{K_pe^{-\theta S}}{S}$	$\frac{2\tau_{cl}+\theta}{(\tau_{cl}+\theta)^2}$	$2\tau_{cl}+\theta$	–	$\frac{2\tau_{cl}+\theta}{(\tau_{cl}+\theta)^2}$	$2\tau_{cl}+\theta$
$\frac{K_pe^{-\theta S}}{S}$	$\frac{2\tau_{cl}+\frac{\theta}{2}}{\left(\tau_{cl}+\frac{\theta}{2}\right)^2}$	$2\tau_{cl}+\frac{\theta}{2}$	$\frac{\theta}{2}$	$\frac{2\tau_{cl}+\frac{\theta}{2}}{\left(\tau_{cl}+\frac{\theta}{2}\right)^2}$	$2\tau_{cl}+\frac{\theta}{2}$
$\frac{K_pe^{-\theta S}}{S}$	$\frac{\frac{\theta}{2}}{\left(\tau_{cl}+\frac{\theta}{2}\right)^2}$	$\frac{\theta}{2}$	$2\tau_{cl}+\frac{\theta}{2}$	$\frac{\frac{\theta}{2}}{\left(\tau_{cl}+\frac{\theta}{2}\right)^2}$	$\frac{\theta}{2}$
$\frac{K_pe^{-\theta S}}{S(\tau S+1)}$	$\frac{2\tau_{cl}+\theta}{(\tau_{cl}+\theta)^2}$	$2\tau_{cl}+\theta$	τ	$\frac{2\tau_{cl}+\theta}{(\tau_{cl}+\theta)^2}$	$2\tau_{cl}+\theta$
$\frac{K_pe^{-\theta S}}{S(\tau S+1)}$	$\frac{\tau}{(\tau_{cl}+\theta)^2}$	τ	$2\tau_{cl}+\theta$	$\frac{\tau}{(\tau_{cl}+\theta)^2}$	τ

and-error tuning procedure and that a set of tuning constants can be easily developed in which we can have a great deal of confidence. The plant has had no problems with this control loop since this tuning change.

High-purity distillation column

IMC–PID tuning rules can be applied to the middle-column temperature control of a high-purity distillation column using the tuning method for time-constant-dominant processes. This work originated with the control strategy definition of a project to combine the refining of two different products into an existing process. The column of interest is 10 ft in diameter and 110 ft high. The materials being separated are a mixture of three isomers and a small amount of other heavy components. The feed to the column is the bottom product of a topping column that removes water and other lighter components. The separation requires a large number of theoretical stages because the relative volatilities are near one and a high-purity product is required. The column has about 60 theoretical stages. At design production rates, there is very little excess separation capability. In other words, the design has a very low safety margin. This makes tight control very important in maintaining product quality.

The control strategy is illustrated in Figure 10. The control objective is to keep the distillate composition nearly pure in the lightest isomer while maintaining it at a very low level in the tails stream. Good composition control is important because the distillate is a product sold to outside customers.

The heat to the column is fixed because the heat source is a dedicated paracymene (*i.e.,* a heat-transfer fluid) vapor boiler that runs best at a fixed rate. The feed is set to fix the overall production rate for this portion of the process. The column vacuum is controlled at a fixed value by manipulating the vapor flow to the vacuum jet. Base level is controlled by manipulating the tails flow rate. The overhead condensate tank level is controlled by manipulating reflux flow. Composition control is accomplished by controlling the middle-column temperature. This loop manipulates the distillate flow rate. Middle-column temperature is used for composition control rather than an analyzer on the distillate stream because the temperature measurment is more reliable, much cheaper, and correlates very well with top composition. Proper location of the temperature sensor is very important. We use the steady-state model to determine the optimum location. A feedforward control loop changes reflux flow in response to distillate flow rate changes. This effectively eliminates the condensate tank level loop dynamics from the middle-column temperature loop. There is no in-

PID(3)		
$K_c K_p$	τ_I	τ_D
$\dfrac{\tau}{\tau_a + \theta}$	τ	–
$\dfrac{\tau + \frac{\theta}{2}}{\tau_a + \frac{\theta}{2}}$	$\tau + \dfrac{\theta}{2}$	$\dfrac{\tau\theta}{2\tau + \theta}$
–	–	–
$\dfrac{\tau_1 + \tau_2 - \tau_3}{\tau_a + \theta}$	$\tau_1 + \tau_2 - \tau_3$	$\dfrac{\tau_1\tau_2 - (\tau_1 + \tau_2 - \tau_3)\tau_3}{\tau_1 + \tau_2 - \tau_3}$
–	–	–
$\dfrac{2\zeta\tau - \tau_3}{\tau_a + \theta}$	$2\zeta\tau - \tau_3$	$\dfrac{\tau^2 - (2\zeta\tau - \tau_3)\tau_3}{2\zeta\tau - \tau_3}$
$\dfrac{\tau_1 + \tau_2 + \frac{\tau_3\theta}{\tau_a + \tau_3 + \theta}}{\tau_a + \tau_3 + \theta}$	$\tau_1 + \tau_2 + \dfrac{\tau_3\theta}{\tau_a + \tau_3 + \theta}$	$\dfrac{\tau_3\theta}{\tau_a + \tau_3 + \theta} + \dfrac{\tau_1\tau_2}{\tau_1 + \tau_2 + \frac{\tau_3\theta}{\tau_a + \tau_3 + \theta}}$
–	–	–
$\dfrac{2\zeta\tau + \frac{\tau_3\theta}{\tau_a + \tau_3 + \theta}}{\tau_a + \tau_3 + \theta}$	$2\zeta\tau + \dfrac{\tau_3\theta}{\tau_a + \tau_3 + \theta}$	$\dfrac{\tau_3\theta}{\tau_a + \tau_3 + \theta} + \dfrac{\tau^2}{2\zeta\tau + \frac{\tau_3\theta}{\tau_a + \tau_3 + \theta}}$
$\dfrac{2\tau_a + \theta}{(\tau_a + \theta)^2}$	$2\tau_a + \theta$	–
$\dfrac{2\tau_a + \theta}{\left(\tau_a + \frac{\theta}{2}\right)^2}$	$2\tau_a + \theta$	$\dfrac{\tau_a\theta + \frac{\theta^2}{4}}{2\tau_a + \theta}$
–	–	–
$\dfrac{2\tau_a + \tau + \theta}{(\tau_a + \theta)^2}$	$2\tau_a + \tau + \theta$	$\dfrac{(2\tau_a + \theta)\tau}{2\tau_a + \tau + \theta}$
–	–	–

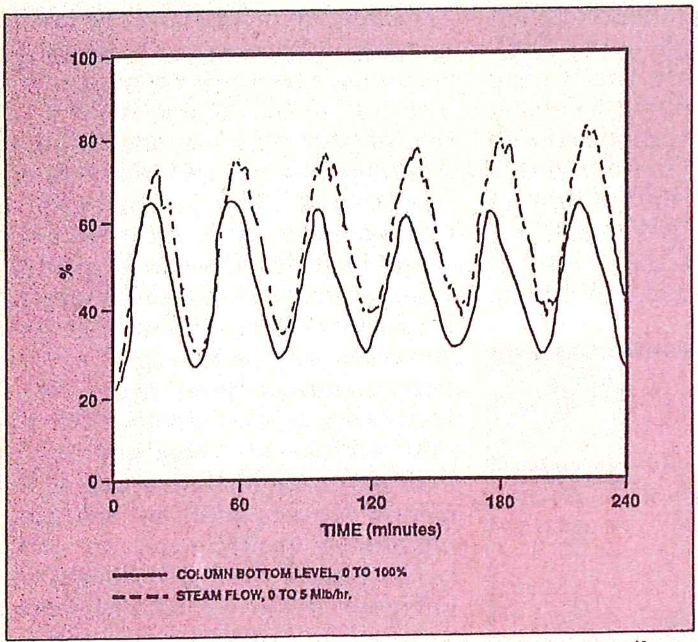

Figure 6. Column bottom level response with existing controller tuning parameters (real data).

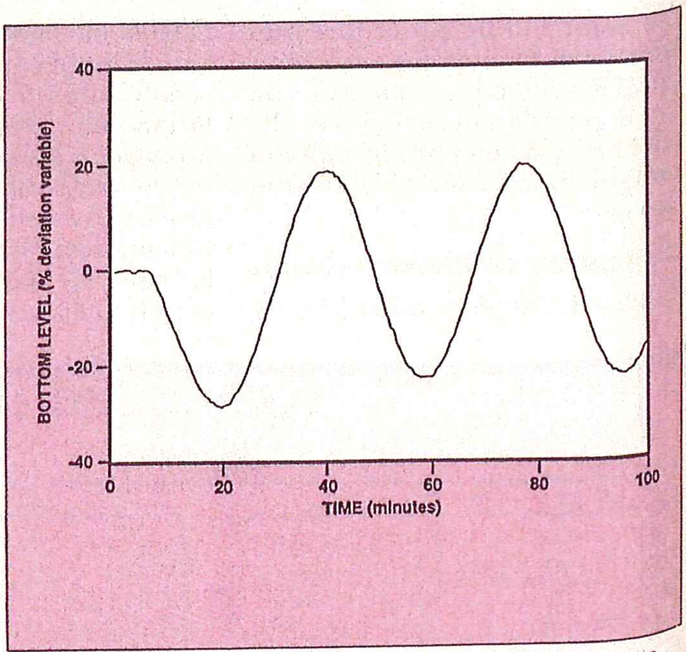

Figure 7. Simulation response with existing controller tuning parameters for a step load disturbance.

centive for dual composition control in this column; energy savings are not possible because of the need for a fixed heat input. In addition, the light isomer concentration in the column bottom can vary as long as the average stays below the specification.

For this example, the process model was developed using an open-loop step test on the process. The first step was to bring the process to a steady-state condition. After some trial and error, it was found that a sustained distillate rate of 7.9 thousand pounds/hr (MPPH), with everything else held constant, resulted in keeping the middle-column temperature very steady. In addition, 0.5 MPPH deviations above and below this equilibrium value would cause the temperature to move up or down at a reasonable and controlled rate. Of course, some deadtime between change and response was observed. After learning how the process responded, it was easy to bring the column to a steady state at the desired temperature. The open-loop tests were then conducted as shown in Figure 11. The distillate rate was varied up and down as shown. From this open-loop test and the earlier response, it was observed that the response most closely resembled that of an integrating process with deadtime. Although it was believed that the process would ultimately steady out, a deadtime first-order model form was not selected to describe the process because the closed-loop load distur-

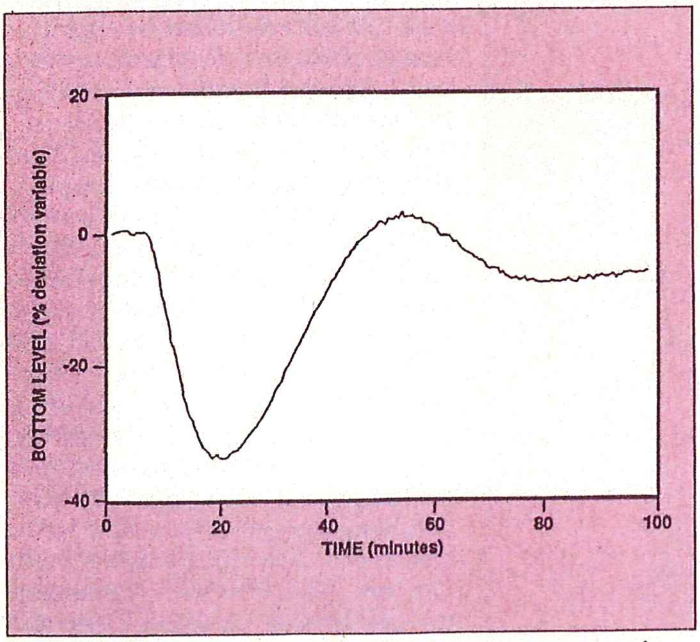

Figure 8. Simulation response with IMC–PID tuning parameters for a step load disturbance.

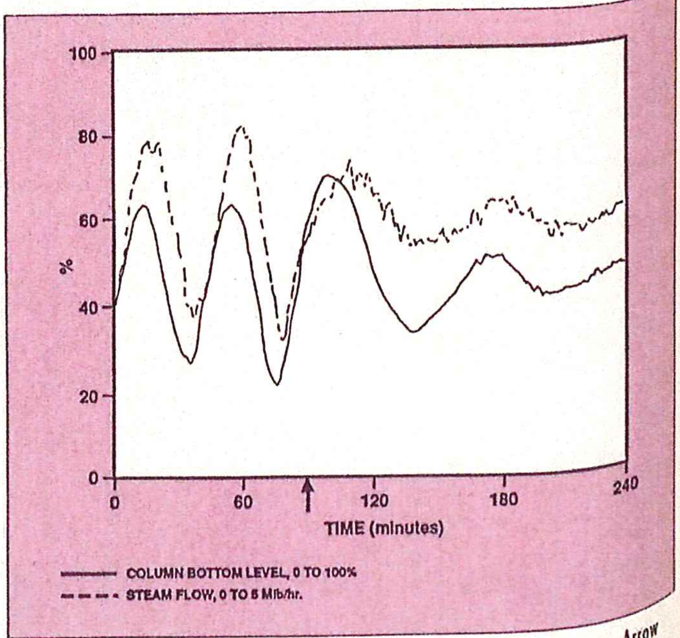

Figure 9. Column bottom level response with IMC-PID tuning parameters. Arrow indicates the point at which the tuning parameter changes are made.

Disturbance rejection can be significantly improved for long time-constant-to-deadtime processes by assuming they are deadtime-plus-integrator forms.

bance response would be very slow, the process did not follow a first-order form because of its nonlinear nature, and the controller always kept the process in the region where the initial response characterized the process dynamics.

The Laplace transfer function model and parameters calculated from the step tests are

$$\frac{\text{temperature}}{\text{distillate}} = \frac{0.01e^{-5.5S}}{S}$$

$$K_p = \frac{\Delta y/\Delta t}{\Delta u}$$

where t is the time (in minutes), Δy and Δu are in percentage of scale, the distillate span is 15 MPPH, and the temperature span is 30°C. The deadtime, which is the time from the first step change until the temperature begins to move, is 5.5 min. K_p, which is the change in temperature (Δy) per unit time divided by the change in distillate flow (Δu), works out to be 0.01.

The next step was to go to closed-loop temperature control. A conservative selection of τ_{CL} = 30 min was used following the guidelines outlined previously. PI controller tuning constants were calculated using the formulas in Table 1. The loop was then closed and stable control was observed. A 20% step reduction in the feed rate to the column was introduced. The closed-loop response is illustrated in Figure 12. As expected, the response exhibited no overshoot. The maximum deviation

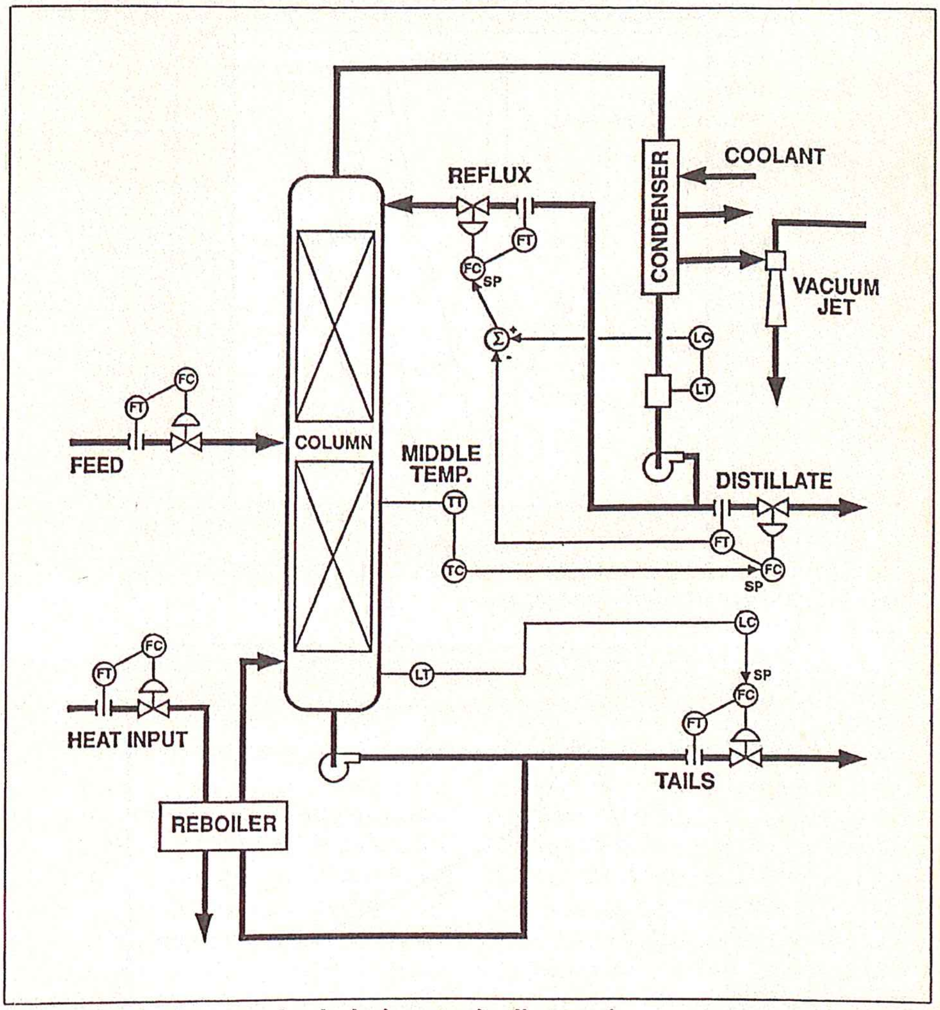

Figure 10. Control strategy for the high-purity distillation column.

Table 2. Column distillate composition.*

Date	Time	Light Component	Intermediate Component
1/18	12:30 A.M.	99.604	–
	3:00 A.M.	99.709	0.027
	5:00 A.M.	99.599	0.038
1/18	7:00 A.M.	99.57	0.04
	9:00 A.M.	99.60	0.035
	11:00 A.M.	99.601	0.031
	1:00 P.M.	99.591	0.026
	3:00 P.M.	99.631	0.031
	5:00 P.M.	99.630	0.031
	7:00 P.M.	99.600	0.033
	9:00 P.M.	99.620	0.033
	11:00 P.M.	99.609	0.038
1/19	1:00 A.M.	99.641	0.037
	3:00 A.M.	99.622	0.035
	5:00 A.M.	99.627	0.033
	7:00 A.M.	99.67	0.031
	9:00 A.M.	99.69	–
	11:00 A.M.	99.65	0.038
	1:00 P.M.	99.67	0.037
	3:00 P.M.	99.62	0.038
	5:00 P.M.	99.627	0.043
	7:00 P.M.	99.64	0.043

* Light component mean, 99.628; standard deviation, 0.0336. Heavy component composition is below detectable limits.

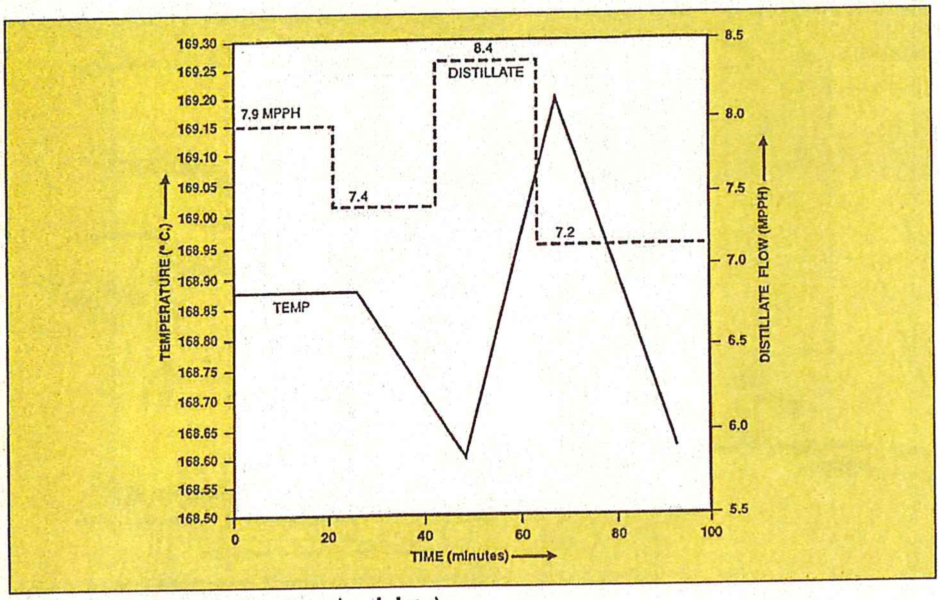

Figure 11. Open-loop step tests (real data).

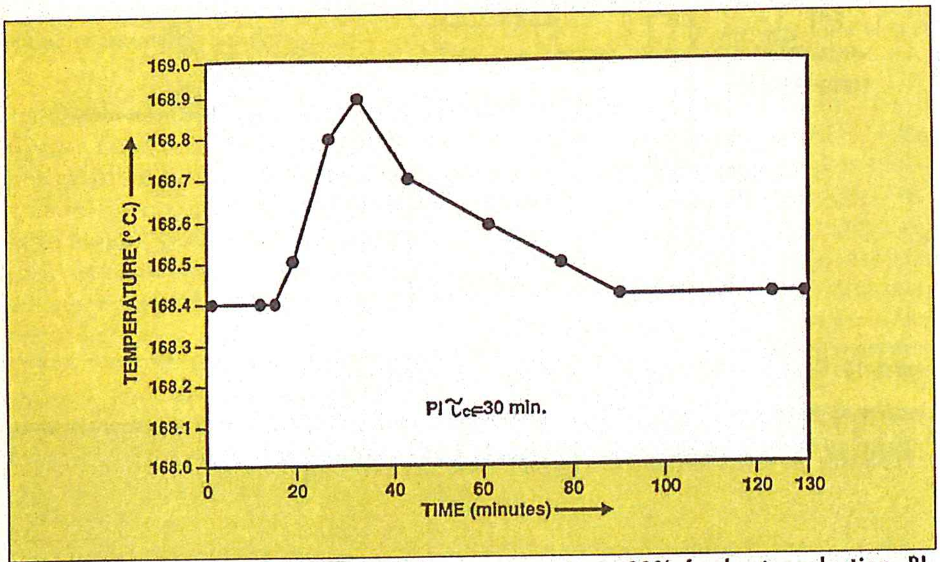

Figure 12. Middle temperature closed-loop response to a 20% feed rate reduction, PI, τ_{cl} = 30 min (real data).

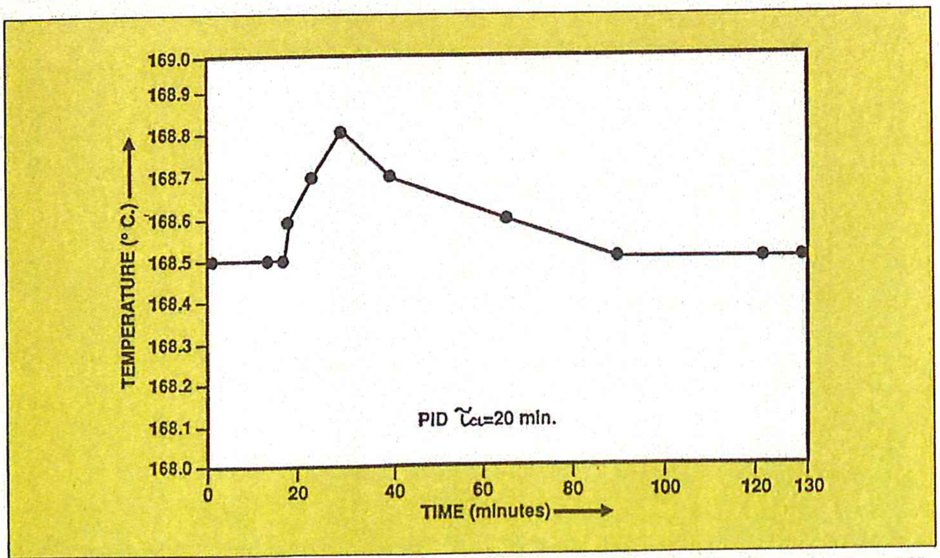

Figure 13. Middle temperature closed-loop response to a 20% feed rate reduction, PID, τ_{cl} = 20 min (real data).

in middle temperature was found to be only 0.5°C.

To achieve even better control and to further evaluate the IMC–PID technique, derivative action was added to this control loop. We felt that the additional control action would allow us to achieve better control, so we selected a smaller τ_{CL} = 20 min. The response to the same feed rate disturbance is shown in Figure 13. Again, no overshoot was observed in the response. The main difference is that the maximum temperature deviation was smaller (0.3°C). Because of the improved performance, the tuning was left at these values.

The manipulated variable (*i.e.*, distillate flow) response for PI and PID control is shown in Figure 14, which illustrates an important aspect of using derivative action. Note the much larger variability in the distillate flow when PID control is used. This shows that control with derivative action does not dampen the noise in the temperature measurement nearly as well as the PI controller. If the distillate was the feed to a downstream unit operation, PID control might not be desirable because of the added variability in the distillate flow. The magnitude of the noise in the manipulated variable is a function of the frequency and magnitude of the noise in the measurement and the tuning constants. This illustrates one aspect of the performance vs. disturbance dampening trade-off that one is often faced with in industry. We believe this is one reason why derivative action is not applied more often.

The last of the closed-loop tests to be reported is the response of the controls to a step change stopping of the column feed flow, which is a very severe disturbance. Figure 15 illustrates the middle-column temperature and distillate flow response to a 4-min loss of feed flow. The maximum temperature deviation is 1.8°C. The distillate flow stops completely and is then reestablished at the same rate as before the feed interruption. Again, little or no overshoot is observed. When the distillate is stopped, the column automatically goes into a total reflux mode of operation. The middle-column temperature control shuts off the distillate flow before the top composition goes out of specification. This example shows that the middle-column temperature control is so responsive and the overall control structure so flexible that the feed to the column can be stopped without operator intervention (*i.e.*, all controllers remain in automatic). At times, the feed to the column is frequently stopped because downstream equipment is plugged. This responsive and flexible control structure allows the column feed to be stopped and started with no delay.

As you recall, the control objective is to control composition of the distillate stream. Table 2 contains almost two days of laboratory samples from the actual operation of the column under PID control. The mean distillate composition is 99.628 with a standard deviation of 0.0336. Clearly, the control objective is being met.

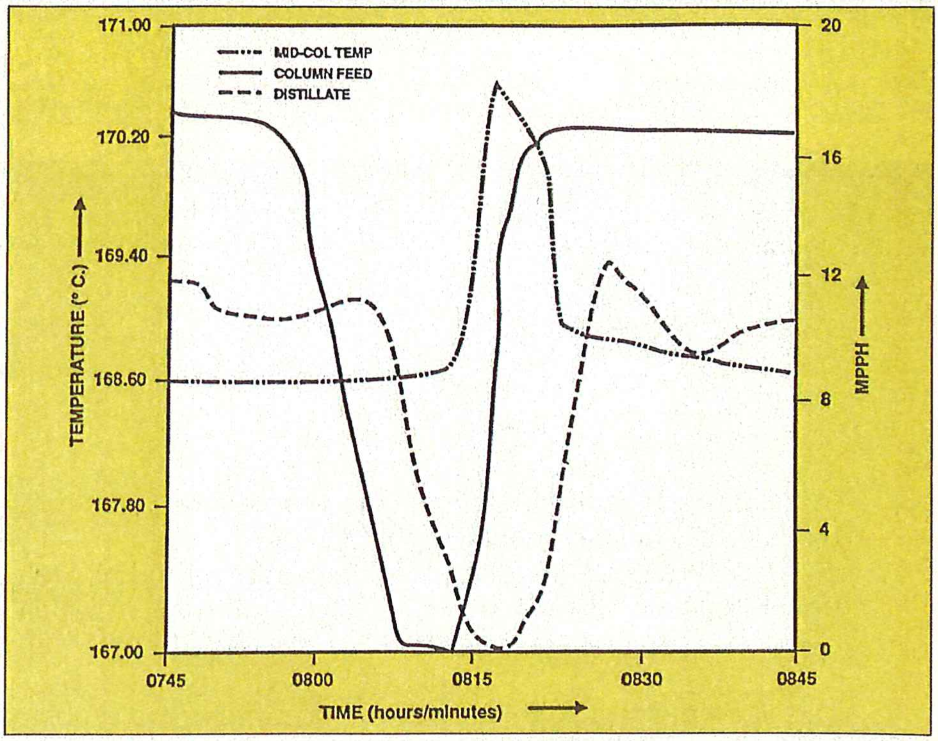

Figure 15. Closed-loop response to a 4-min loss of feed flow, PID, τ_{cl} = 20 min (real data).

In conclusion

The IMC–PID tuning rules are a superior industrial criteria. The no-overshoot character helps to minimize disturbance propagation. The rules reduce the tuning process to the selection of one tuning parameter as opposed to three. In addition, the single parameter is directly related to the closed-loop speed of response and robustness. Process models can be developed directly from open-loop tests or combinations of tests and first-principle equations. Disturbance rejection can be significantly improved for long time-constant-to-deadtime processes by assuming they are deadtime-plus-integrator forms. Finally, the rules are being successfully applied to industrial processes. The first example given illustrated how the application of the rules quickly led to high-quality tuning parameters for a troublesome level control loop. The second illustrated how the rules were applied to obtain very responsive tuning for control of composition for a high-purity distillation column.■

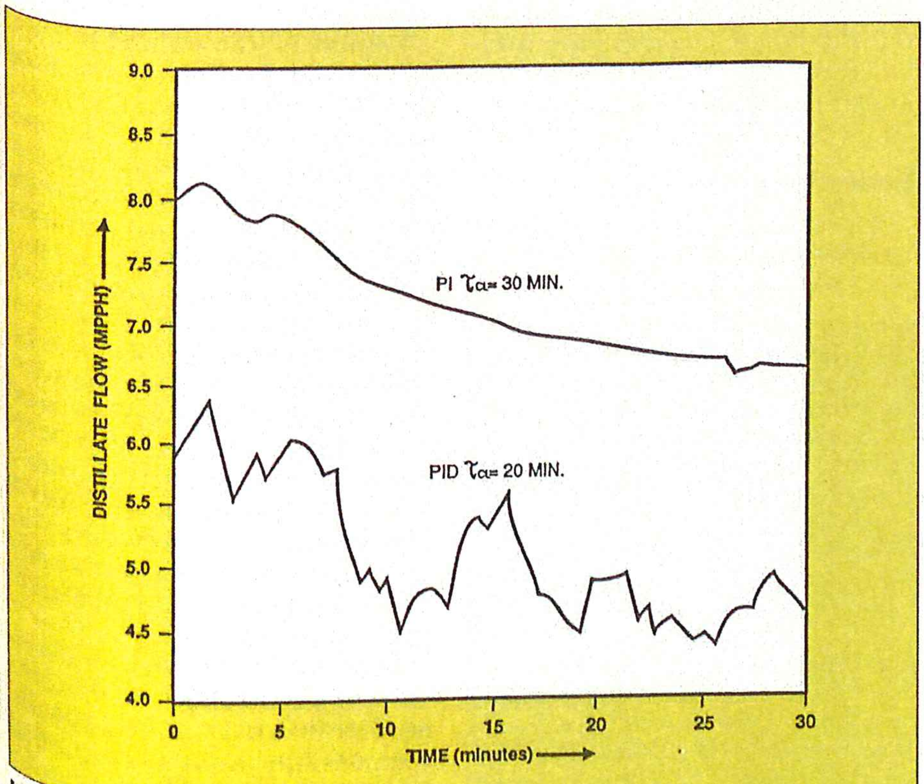

Figure 14. Closed-loop response, distillate flow with and without derivative action (real data).

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