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Appendix I. Phase-margin Tuning Formula

If the ultimate gain K_d and period t_d are determined using a relay with cascade integrator, the corresponding point on the Nyquist curve will be the intersection on the imaginery axis at a frequency of $2/t_d$ and magnitude of $-1/K_d$. It is then straight forward to use this information, following Astrom and Hagglund (1984), to derive the following tuning formula to satisfy a phase margin specification of θ_m:

Proportional Gain $K = K_d \sin \theta_m$

Integral Time $T_i = \dfrac{t_d (1 - \cos \theta_m)}{\pi \sin \theta_m}$

Derivative Time $T_d = T_i / 4$

Appendix II. Weighting On Setpoint

In the dominant pole design of Hagglund and Astrom (1985), a new tuning parameter 'b', which can be interpreted as a weighting factor on setpoint in the proportional term, has been introduced to modify the setpoint response independent of the load recovery response. This technique can be applied to the controller tuned by the Ziegler-Nichols formula which has been found to be near optimum for a step load change but producing an excessive overshoot for a step setpoint change. Using the symbol of Fig. 1 the PID controller becomes:

$$u = K \left[(b\, y_{ref} - y) + \frac{1}{T_i} \int e \, dt - T_d \frac{dy}{dt} \right]$$

It has been found experimentally that a small 'b' of 0.3 - 0.6 will reduce the large overshoot in setpoint response when the loop gain is high. Empirical formulae for computing 'b' are proposed in Hang and Astrom (1988).

"IMC-PID CONTROLLER DESIGN - AN EXTENSION"

I-Lung Chien

Engineering Department
E. I. du Pont de Nemours and Co (Inc)
P. O. Box 6090, Newark, Delaware 19714

Abstract. PID controller design based on Internal Model Control (IMC) design procedure by Rivera and co-workers has been extended to cover a wider range of process models. PID tuning parameters for processes up to second-order with first-order numerator dynamics and dead time will be given with only one tuning parameter (closed-loop time constant). In order to obtain the PID formulation, dead time is approximated by either a first-order Padé or a first-order Taylor series expansion depending on whether a PI or PID controller is desired. Various industrial PID implementations which require different PID settings will also be examined in this paper.

Several examples including a first-order plus dead time, second-order plus dead time and inverse response, an industrial level control problem, and a nonlinear distillation column model will be used to compare the performance of IMC-PID controller design with more traditional tuning rules, eg, Ziegler-Nichols, Cohen-Coon, and frequency response maximum peak criterion. Model mismatches in process gain, dead time, and time constant will also be introduced to compare the robustness of various controller design methods.

1. INTRODUCTION

Although many advance control concepts have been introduced within the last twenty years, the vast majority of the controllers in chemical industry are still of the PID type. From a maintenance and reliability point of view, most plants prefer the commercially available control vendor hardware. Thus, properly tuned PID controllers are still an important factor for successful plant operation.

There are many controller tuning methods proposed in the literature. Perhaps the most well-known ones are the Ziegler-Nichols rules in 1942 and the Cohen-Coon rules in 1953. Ziegler-Nichols uses critical gain and frequency information to determine P, I, and/or D parameters based on quarter decay criterion. Later extensions include Hägglund and Åström (1985) and Hwang and Chang (1987). Cohen-Coon utilizes a first-order plus dead time model with quarter decay criterion to determine the tuning parameters.

Later extensions were made to higher-order models, with a variety of performance criteria (eg, IAE, ITAE, ISE, maximum peak, phase margin, etc) by Gallier and Otto (1968), Smith (1972), Fertik (1975), Weber and Bhalodia (1979), Huang and Chao (1982), Ralston and co-workers (1985), Tan and Weber (1985), Harris and Mellichamp (1985), and, Patwardhan and co-workers (1987). Yuwana and Seborg proposed a closed-loop method to find the first-order plus dead time model in 1982 and was extended by Jutan and Rodriguez (1984).

Rivera and co-workers introduced a PID controller design method based on Internal Model Control in 1986. One tuning parameter and first-order closed-loop response makes this method very attractive to an industrial user. The flexibility of this tuning method gives the user a convenient way to find a compromise between performance and robustness.

In this paper, the IMC-PID design by Rivera and co-workers will be extended to cover a wider range of process models. After a brief review of IMC-PID design in Section 2, various industrial PID implementations will be discussed in Section 3. A summary table of all the IMC-PID tuning rules will be presented in Section 4. The performance and robustness of this tuning method will be examined via simulation of several numerical examples.

2. REVIEW OF IMC-PID CONTROLLER DESIGN

The Internal Model Control (IMC) was introduced by Garcia and Morari in 1982. By using a control structure as Figure 1, it provides a straightforward, two-step controller design procedure as follows:

Step 1. Factor the model

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

Such that G_{M+} contains all the dead times and right half plane zeros; consequently G_{M-} is stable and does not involve predictors.

Step 2. Define the IMC controller by

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

where f is a user-specified low pass filter.

A common selection for the filter which is suitable for the chemical industry is a first-order filter with filter time constant selected to be the desired closed-loop time constant. For systems with an integrator, the closed-loop system needs to satisfy the property of zero offset for ramp inputs. For this case, a proper selection for the filter is

$$f = \frac{(2\tau_{CL} - G'_{M+}(0))\ S + 1}{(\tau_{CL}\ S + 1)^2} \qquad (3)$$

where τ_{CL}, a user-specified parameter, corresponds to closed-loop speed of response. Figures 2 and 3 show how IMC structure can be put into the conventional feedback control structure through cancellation of two dashed line paths involving G_M. The relation between the feedback controller and the IMC controller is

$$G_C = \frac{G_{IMC}}{1 - G_{IMC}\ G_M} \qquad (4)$$

In a subsequent paper (Rivera, Morari, and Skogestad, 1986), the authors use above Eqns 1, 2, and 4 to derive PID controller settings for some simple process models mostly without dead time and also gave "improved" PI controller settings for a first-order process with dead time.

In this paper, the above work will be extended to include up to a second order process with dead time. Different industrial PID implementations which result in different PID tuning parameters will also be given.

3. PID CONTROLLER IMPLEMENTATION

There is no industrial standards for PID controller implementation, thus different controller vendor may provide different PID algorithms.

Gerry (1987) gave a good comparison of different PID algorithms. The most common PID algorithms that control vendors provide are as follows: (in Laplace transformation)

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

$$\text{PID(2)}: \quad G_C = K_C \left(1 + \frac{1}{\tau_I S}\right)\left(\frac{1 + \tau_D S}{\frac{\tau_D}{DG} S + 1}\right) \qquad (6)$$

$$\text{PID(3)}: \quad G_C = K_C \left(1 + \frac{1}{\tau_I S} + \frac{\tau_D S}{\frac{\tau_D}{DG} S + 1}\right) \qquad (7)$$

where K_C = Controller proportional gain (dimensionless)

τ_I = Reset time (units of time)

τ_D = Derivative time (units of time)

DG = Derivative gain

Moore Products uses PID form Eqn 5, Honeywell uses Eqn 6, and Bailey uses PID form Eqn 7. Most of the control vendors have derivative action acting only on the measurement signal to prevent derivative kick when a set point change is made. Some vendors also provide an option to let proportional action act only on the measurement signal. This will provide a smoother set point response.

4. IMC-PID TUNING RULES

With the above industrial PID implementation, the IMC-PID tuning rules can be extended to many process models commonly used in the chemical industry. The derivation of the PID parameters is straightforward:

(1) Approximate dead time by either a first-order Padé or a first-order Taylor series.

(2) Select f as:

$$f = \frac{1}{\tau_{CL}\ S + 1}$$

for first and second order processes, or as Eqn 3 for processes with an integrator.

(3) Use Eqns 1, 2, and 4 to find the feedback controller G_C.

(4) Compare with Eqns 5-7 and equate the coefficients to find the relationship between PID parameters and model parameters.

The PID tuning rules for commonly used process models in chemical industry is given in Table 1. This derivation is not new; a portion of the tuning rules has already been given in Smith and co-workers (1975) by using a direct controller synthesis method. For first-order processes with dead time and processes with an integrator and dead time, both PI and PID settings are given. PID form, Eqns 5 and 6, do not result in a real solution for a second-order underdamped process.

Some process models result in two sets of PID parameters, but the noise filtering characteristic will be different. Most control vendors do not allow the user to select derivative gain (DG) with few exceptions (eg, Moore Products). If derivative gain is set at a constant (eg, 10) by the control vendor, then the solution with larger τ_D will result in a better noise filtering characteristic. Simulation results show that a desired closed-loop response can be achieved even though the derivative gain is fixed by the vendor.

5. SIMULATION RESULTS

In this section, simulation results for four numerical examples are presented.

Example 1 - first-order plus dead time process (see notation in Figure 1).

$$G_P = G_L = \frac{e^{-3S}}{3S + 1}$$

Figure 4 shows the load response for IMC-PI settings with τ_{CL}=3, Ziegler-Nichols, and Cohen-Coon tuning methods. Notice that IMC-PI gives a smooth closed-loop response for both process output and manipulated variable changes. On the contrary, the Ziegler-Nichols method gives a sluggish response for this large dead time process and Cohen-Coon results in an overshoot oscillatory closed-loop response. In order to test the robustness of the tuning rules, model mismatches in process gain and dead time are presented in Figures 5 and 6. Both figures clearly show the superior closed-loop behavior of the PI settings based on IMC under severe model mismatch conditions.

In a separate simulation (not shown), two different first-order with dead time processes with the same cross-over frequency are used to illustrate the different closed-loop behavior of Ziegler-Nichols and IMC-PID tuning methods. These two processes are:

$$G'_P = G'_L = \frac{e^{-6S}}{2S + 1}$$

$$G''_P = G''_L = \frac{e^{-4.135S}}{20S + 1}$$

Comparing set point response, Ziegler-Nichols PI gives a very sluggish response on G'_p but overshoot on G''_p; but the IMC-PI with same selected τ_{CL} gives very similar closed-loop responses for these two vastly different cases. This indicates that IMC-PID method is suitable for a much wider dead time/lag ratio range for set point responses. For load responses since IMC-PID retains the open-loop load response into the closed-loop, the response for G''_L is slow compared to Ziegler-Nichols no matter how fast τ_{CL} is chosen. This might be the less desirable property of IMC-PID method for lag dominant processes.

Example 2 - second-order nonminimum phase system with dead time.

$$G_p = G_L = \frac{(-2S + 1)e^{-4S}}{(5S + 1)(3S + 1)}$$

Table 1 shows that no PI setting can be derived from the IMC method so a PID controller is required. In the simulation, PID algorithm 1 (Eqn 5) is used for controller implementation. IMC-PID method is compared with classical frequency response maximum peak criterion method. The PI setting for maximum peak criterion method is chosen as follows: (Buckley, 1964)

(1) Find P-only controller setting so that closed-loop maximum peak is at 2 db.

(2) Set reset time at 5.0 divided by the resonant frequency.

(3) Find K_C in the PI controller with reset time in step 2 so that the closed-loop maximum peak is at 2 db again.

Figure 7 shows the comparison for load response between IMC-PID and the maximum peak method. The maximum peak method results in very sluggish response but the IMC-PID again gives smooth load rejection performance. The process gain and dead time mismatches are shown in Figures 8 and 9 respectively. Notice again that IMC-PID tuning rules is much more robust than the maximum peak criterion method.

Example 3 - level control system (process with integrator)

The IMC-PID tuning rules has been applied to a industrial level control system with great success. The process fluid from a reactor is pumped through a flasher into a separator. Water in the process fluid is removed through the vapor phase. The process fluid is pumped out to a downstream process. The separator level is controlled by the positive displacement pump speed which is located upstream of the flasher. By using material balance equation, this level system can easily be modeled as:

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

where

$$K_p = \frac{\text{Maximum flow (lb/min)}}{\text{Level span(ft) x Density(lb/ft}^3\text{) x cross sectional area(ft}^2\text{)}}$$

Parameters in the model are all readily available and dead time can easily be obtained by open-loop pulse test. For this system, the model is

$$G_p = -G_L = \frac{1.35e^{-4S}}{S}$$

Figure 10 shows the IMC-PI response for a downstream pump speed step increase of 1%. Three different τ_{CL} are used, they are 4, 8, and 12 minutes. Smaller τ_{CL} gives a faster return to the level set point but with more overshoot on the manipulated variable. On the other hand, larger τ_{CL} results in slower response but with smoother movement in the manipulated variable. Figures 11 and 12 show the disturbance responses with process gain and dead time mismatches respectively. These figures clearly show that although the larger τ_{CL} sacrifices performance, it can tolerate more model mismatch. The user should choose τ_{CL} so that it satisfies the performance requirement but still allows some model mismatch. τ_{CL}=8 minutes was used in the plant.

Example 4 - nonlinear distillation column model.

A nonlinear distillation column model is used to demonstrate IMC-PID tuning rules under severe model mismatch condition. The column is a typical purge fractionation column with nine trays and feed is introduced on tray six. The base case condition is:

Overhead composition = 99.7 % light component
Bottom composition = 8.22 % light component
Feed composition = 90 % light component
Feed flow = 100.0 lb mole/min
Reflux flow = 250.6 lb mole/min
Boilup flow = 340.0 lb mole/min
Distillate flow = 89.4 lb mole/min
Bottom flow = 10.6 lb mole/min
Measurement dead time = 10 minutes

Process noise is simulated by adding white noise with a standard deviation of 1.0 lb mole/min in the column feed and another white noise with a standard deviation of 3.0 lb mole/min in the boilup flow rate. The composition of the overhead which is of most importance is controlled by reflux flow. A typical disturbance to the column is a feed composition change. In order to find the overhead composition tuning parameters, a open-loop test is performed. Reflux flow is perturbed by a pseudo-random binary sequence with a magnitude of 5 lb mole/min. The time response of overhead composition is recorded. Then, an identification routine is used to find an approximate first-order plus dead time model as follows:

$$G_M = \frac{1.06e^{-10S}}{23.1S + 1}$$

(Assume reflux flow span of 0-500 lb mole/min and overhead composition span of 0-1).

IMC-PI setting is calculated by selecting τ_{CL}=20 min, ie, K_C=0.73 and τ_I=20 min. Figure 13 shows the closed-loop response of the above PI settings with feed composition step changes as follows:

$t < 200$ min, feed composition = 90 %
$200 \leq t < 400$ min, feed composition = 92 %
$400 \leq t < 600$ min, feed composition = 90 %
$600 \leq t < 800$ min, feed composition = 88 %
$800 \leq t < 1000$ min, feed composition = 90 %

The disturbance rejection capability is quite good even for this highly nonlinear distillation column.

6. CONCLUSIONS

A simple and robust PID tuning method suitable for industrial application is presented. The IMC-PID design procedure by Rivera and co-workers has been extended to cover a wider range of process models. A complete list of the PID tuning rules for most commonly used process models in the chemical industry is given. Only one tuning parameter needs to be specified by the user which determines the closed-loop speed of response.

The model parameters can be obtained from first principle (ex 3) or system identification (ex 4). Simulation results show that this method with proper choice of τ_{CL} is very robust under even severe model mismatch conditions. A safe first trial for selecting τ_{CL} is to set it equal to the open-loop dominant time constant. Then the on-line adjustment of this parameter for the performance requirement is trivial because the user can anticipate the direction the controller will perform.

This PID tuning method can easily be connected to a on-line system identification routine to control processes with time-varying parameters.

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TABLE 1. IMC-PID TUNING RULES

PROCESS MODEL	PID (1)			PID (2)			PID (3)		
	KcKp	τ_I	τ_D	KcKp	τ_I	τ_D	KcKp	τ_I	τ_D
$\frac{Kpe^{-\theta S}}{\tau S+1}$	$\frac{\tau}{\tau_{CL}+\theta}$	τ	--	$\frac{\tau}{\tau_{CL}+\theta}$	τ	--	$\frac{\tau}{\tau_{CL}+\theta}$	τ	--
$\frac{Kpe^{-\theta S}}{\tau S+1}$	$\frac{\tau}{\tau_{CL}+\frac{\theta}{2}}$	τ	$\frac{\theta}{2}$	$\frac{\tau}{\tau_{CL}+\frac{\theta}{2}}$	τ	$\frac{\theta}{2}$	$\frac{\tau+\frac{\theta}{2}}{\tau_{CL}+\frac{\theta}{2}}$	$\tau+\frac{\theta}{2}$	$\frac{\tau\theta}{2\tau+\theta}$
$\frac{Kpe^{-\theta S}}{\tau S+1}$	$\frac{\frac{\theta}{2}}{\tau_{CL}+\frac{\theta}{2}}$	$\frac{\theta}{2}$	τ	$\frac{\frac{\theta}{2}}{\tau_{CL}+\frac{\theta}{2}}$	$\frac{\theta}{2}$	τ	--	--	--
$\frac{Kp(\tau_3 S+1)e^{-\theta S}}{(\tau_1 S+1)(\tau_2 S+1)}$	$\frac{\tau_1}{\tau_{CL}+\theta}$	τ_1	$\tau_2-\tau_3$	$\frac{\tau_1}{\tau_{CL}+\theta}$	τ_1	τ_2	$\frac{\tau_1+\tau_2-\tau_3}{\tau_{CL}+\theta}$	$\tau_1+\tau_2-\tau_3$	$\frac{\tau_1\tau_2-(\tau_1+\tau_2-\tau_3)\tau_3}{\tau_1+\tau_2-\tau_3}$
$\frac{Kp(\tau_3 S+1)e^{-\theta S}}{(\tau_1 S+1)(\tau_2 S+1)}$	$\frac{\tau_2}{\tau_{CL}+\theta}$	τ_2	$\tau_1-\tau_3$	$\frac{\tau_2}{\tau_{CL}+\theta}$	τ_2	τ_1	--	--	--
$\frac{Kp(\tau_3 S+1)e^{-\theta S}}{\tau^2 S^2+2\zeta\tau S+1}$	--	-	--	--	-	--	$\frac{2\zeta\tau-\tau_3}{\tau_{CL}+\theta}$	$2\zeta\tau-\tau_3$	$\frac{\tau^2-(2\zeta\tau-\tau_3)\tau_3}{2\zeta\tau-\tau_3}$
$\frac{Kp(-\tau_3 S+1)e^{-\theta S}}{(\tau_1 S+1)(\tau_2 S+1)}$	$\frac{\tau_1}{\tau_{CL}+\tau_3+\theta}$	τ_1	$\tau_2+\frac{\tau_3\theta}{\tau_{CL}+\tau_3+\theta}$	$\frac{\tau_1}{\tau_{CL}+\tau_3+\theta}$	τ_1	τ_2	$\frac{\tau_1+\tau_2+\frac{\tau_3\theta}{\tau_{CL}+\tau_3+\theta}}{\tau_{CL}+\tau_3+\theta}$	$\tau_1+\tau_2+\frac{\tau_3\theta}{\tau_{CL}+\tau_3+\theta}$	$\frac{\tau_3\theta}{\tau_{CL}+\tau_3+\theta}+\frac{\tau_1\tau_2}{\tau_1+\tau_2+\frac{\tau_3\theta}{\tau_{CL}+\tau_3+\theta}}$
$\frac{Kp(-\tau_3 S+1)e^{-\theta S}}{(\tau_1 S+1)(\tau_2 S+1)}$	$\frac{\tau_2}{\tau_{CL}+\tau_3+\theta}$	τ_2	$\tau_1+\frac{\tau_3\theta}{\tau_{CL}+\tau_3+\theta}$	$\frac{\tau_2}{\tau_{CL}+\tau_3+\theta}$	τ_2	τ_1	--	--	--

TABLE 1. IMC-PID TUNING RULES (CONT'D)

PROCESS MODEL	PID (1)			PID (2)			PID (3)		
	KcKp	τ_I	τ_D	KcKp	τ_I	τ_D	KcKp	τ_I	τ_D
$\frac{Kp(-\tau_3 S+1)e^{-\theta S}}{\tau^2 S^2+2\zeta\tau S+1}$	--	-	--	--	-	--	$\frac{2\zeta\tau+\frac{\tau_3\theta}{\tau_{CL}+\tau_3+\theta}}{\tau_{CL}+\tau_3+\theta}$	$2\zeta\tau+\frac{\tau_3\theta}{\tau_{CL}+\tau_3+\theta}$	$\frac{\tau_3\theta}{\tau_{CL}+\tau_3+\theta}+\frac{\tau^2}{2\zeta\tau+\frac{\tau_3\theta}{\tau_{CL}+\tau_3+\theta}}$
$\frac{Kpe^{-\theta S}}{S}$	$\frac{2\tau_{CL}+\theta}{(\tau_{CL}+\theta)^2}$	$2\tau_{CL}+\theta$	--	$\frac{2\tau_{CL}+\theta}{(\tau_{CL}+\theta)^2}$	$2\tau_{CL}+\theta$	--	$\frac{2\tau_{CL}+\theta}{(\tau_{CL}+\theta)^2}$	$2\tau_{CL}+\theta$	--
$\frac{Kpe^{-\theta S}}{S}$	$\frac{2\tau_{CL}+\frac{\theta}{2}}{(\tau_{CL}+\frac{\theta}{2})^2}$	$2\tau_{CL}+\frac{\theta}{2}$	$\frac{\theta}{2}$	$\frac{2\tau_{CL}+\frac{\theta}{2}}{(\tau_{CL}+\frac{\theta}{2})^2}$	$2\tau_{CL}+\frac{\theta}{2}$	$\frac{\theta}{2}$	$\frac{2\tau_{CL}+\theta}{(\tau_{CL}+\frac{\theta}{2})^2}$	$2\tau_{CL}+\theta$	$\frac{\tau_{CL}\theta+\frac{\theta^2}{4}}{2\tau_{CL}+\theta}$
$\frac{Kpe^{-\theta S}}{S}$	$\frac{\frac{\theta}{2}}{(\tau_{CL}+\frac{\theta}{2})^2}$	$\frac{\theta}{2}$	$2\tau_{CL}+\frac{\theta}{2}$	$\frac{\frac{\theta}{2}}{(\tau_{CL}+\frac{\theta}{2})^2}$	$\frac{\theta}{2}$	$2\tau_{CL}+\frac{\theta}{2}$	--	--	--
$\frac{Kpe^{-\theta S}}{S(\tau S+1)}$	$\frac{2\tau_{CL}+\theta}{(\tau_{CL}+\theta)^2}$	$2\tau_{CL}+\theta$	τ	$\frac{2\tau_{CL}+\theta}{(\tau_{CL}+\theta)^2}$	$2\tau_{CL}+\theta$	τ	$\frac{2\tau_{CL}+\tau+\theta}{(\tau_{CL}+\theta)^2}$	$2\tau_{CL}+\tau+\theta$	$\frac{(2\tau_{CL}+\theta)\tau}{2\tau_{CL}+\tau+\theta}$
$\frac{Kpe^{-\theta S}}{S(\tau S+1)}$	$\frac{\tau}{(\tau_{CL}+\theta)^2}$	τ	$2\tau_{CL}+\theta$	$\frac{\tau}{(\tau_{CL}+\theta)^2}$	τ	$2\tau_{CL}+\theta$	--	--	--

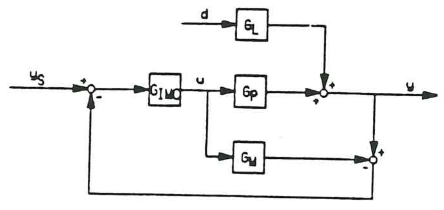

Fig.1. Internal model control structure.

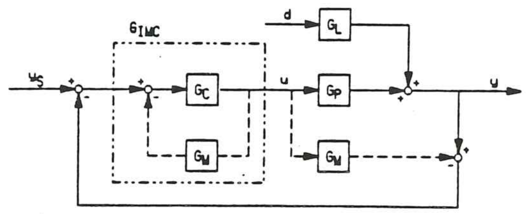

Fig.2. Relationship between IMC structure to feedback control structure

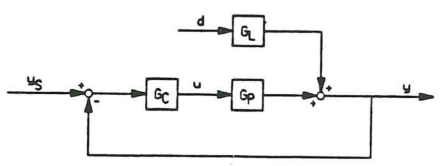

Fig.3. Feedback control structure

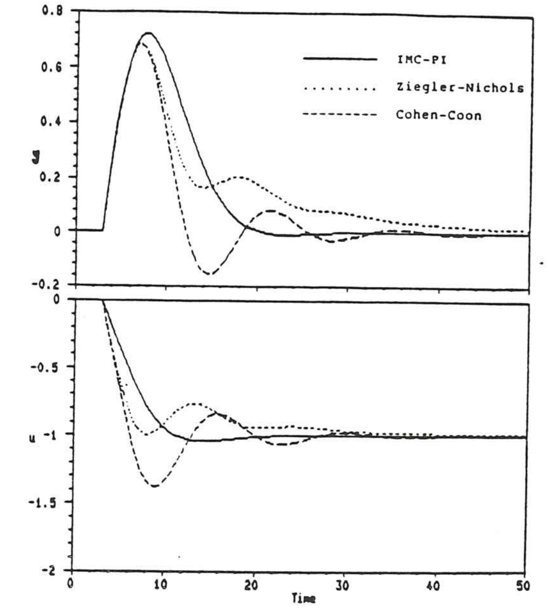

Fig. 4. Closed-loop load response for Example 1.

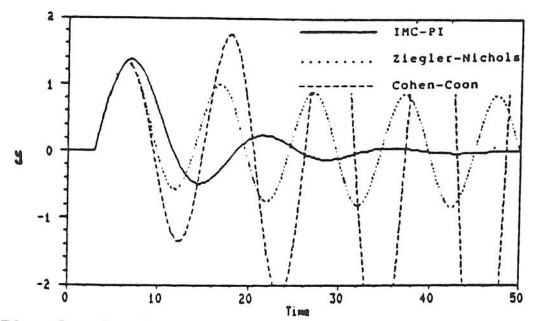

Fig. 5. Load response for Ex.1 with process gain mismatch, process gain of 2.0 instead of 1.0.

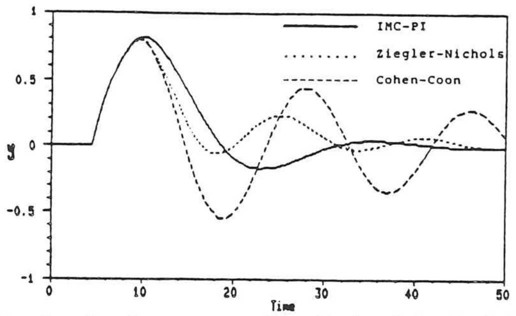

Fig. 6. Load response for Ex.1 with deadtime mismatch, process having deadtime of 4.5 instead of 3.0.

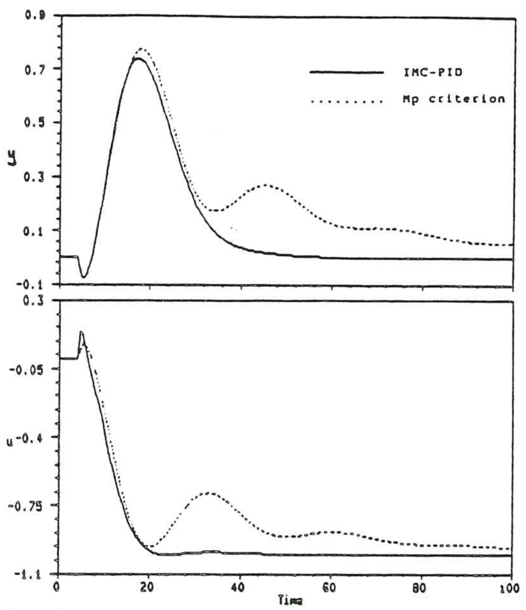

Fig. 7. Closed-loop load response for Example 2.

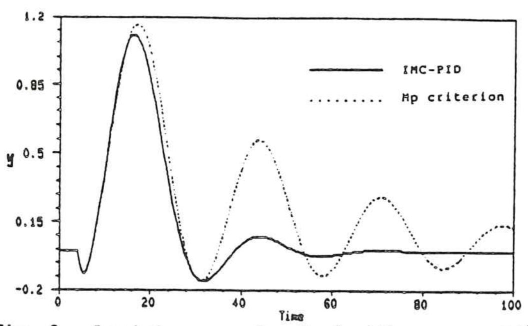

Fig. 8. Load response for Ex.2 with process gain mismatch, process gain of 1.5 instead of 1.0.

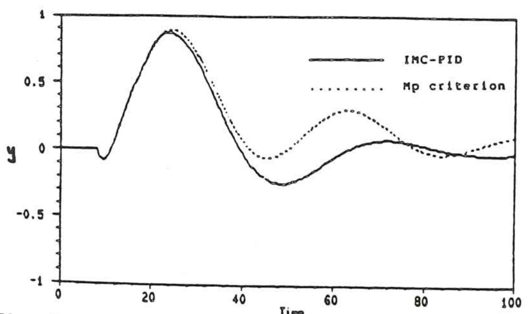

Fig. 9. Load response for Ex.2 with deadtime mismatch, process having deadtime of 8.0 instead of 4.0.

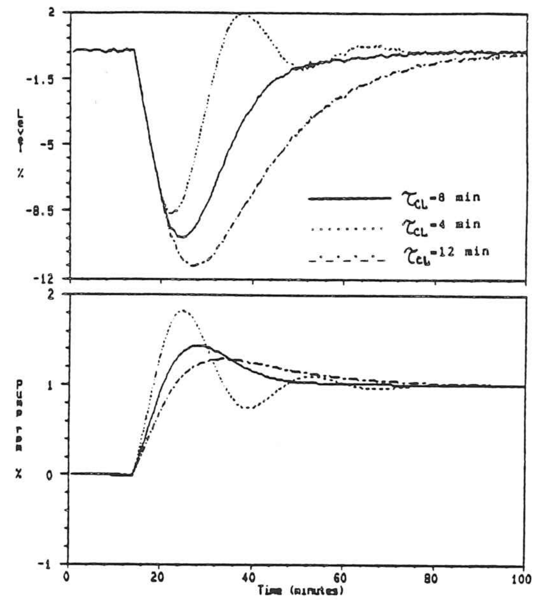

Fig. 10. Closed-loop load response for Example 3.

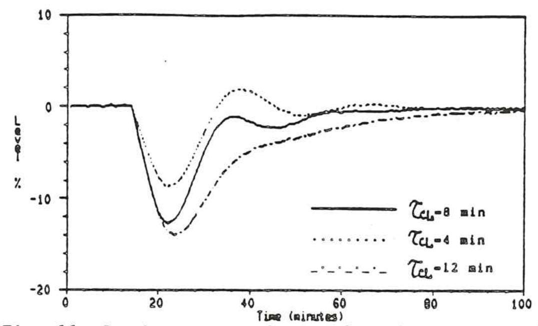

Fig. 11. Load response for Ex.3 with process gain mismatch, process gain of 2.0 instead of 1.35.

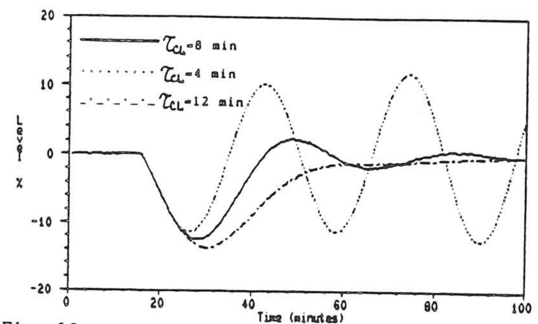

Fig. 12. Load response for Ex.3 with deadtime mismatch, process having deadtime of 6.0 instead of 4.0.

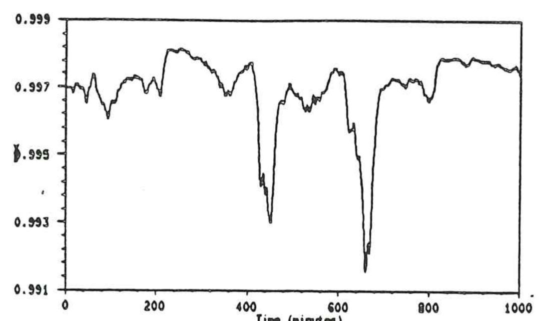

Fig. 13. Response for Ex.4 under feed composition change.

ADAPTIVE PID CONTROL - A POLE PLACEMENT ALGORITHM WITH A SINGLE CONTROLLER TUNING PARAMETER

P.J.Vermeer, A.J.Morris and S.L.Shah

Department of Chemical Engineering
University of Alberta
Edmonton
Alberta. Canada

Abstract. Two approaches to the design of self-tuning PI and PID control strategies are considered. These are based upon a generalized minimum variance control law and a pole-placement control law. The two designs are analysed, using Root Locus techniques, with respect to their stability and transient performance. The control performance of the pole placement design is studied in the presence of model-plant mismatch. Simulation studies using both transfer function representations and a detailed mechanistic model of a binary distillation column demonstrate the capabilities of the pole-placement based design.

Keywords. Self-tuning control, adaptive control, pole placement, Self-tuning PID.

INTRODUCTION

Many process control problems can be adequately and routinely solved by conventional PID control strategies. The overriding reason that the PID controller is so widely accepted is its simple structure which has proved to be very robust with respect to many commonly met process control problems. Eg. unknown process disturbances, process non-linearities, changing process conditions, etc., provided that such occurrences are not too severe. The large number of PID control loops on typical process plant, however, precludes their 'optimal' tuning to account for all possible process disturbances or time varying process changes. Tuning of the PID settings is quite a subjective procedure, relying heavily on the knowledge and skill of the control engineer or even plant operator. Although some tuning guidelines are available (eg. Cohen and Coon and Ziegler-Nichols methods) the process of controller tuning can still be time consuming with the result that many plant control loops are often poorly tuned and full potential of the control system is not achieved.

The fact, however, that PI(D) controllers are so flexible and efficient for many applications would tend to indicate that most process control systems have dynamics which are particularly suited to this type of control. It can be shown that PI(D) controller structures arise naturally from a realistic description of the anticipated process disturbances and an assumption of dominant first (PI) or second (PID) order process dynamics. Allowing controller integral action to arise from a suitable model of the disturbance process results in a well conditioned parameter estimation algorithm and the natural removal of offsets.

Previous work in the area of adaptive and self-tuning PI/PID control has been presented by, for example, Astrom and Hagglund (1984), Hetthessy et al (1983), Cameron and Seborg (1983), Song et al (1984) and Gawthrop (1986). This paper, in contrast, concentrates on the analysis of two different forms of self-tuning (STu) PI/PID algorithm - one based on a generalized minimum variance (GMV) approach and the other based upon a pole-placement (PP) technique. The controller is based on an explicit scheme in that the parameter

P J Vermeer is at present with Polysar, Sarnia, Canada, and A J Morris the Univ. of Newcastle, UK.

estimator identifies model parameters which are then used to calculate the P, I, and D coefficients of the controller. A pole-placement algorithm is recommended which incorporates a single controller 'tuning knob' that is directly related to transient response peak overshoot of a second order system. In the general case of model-plant mismatch this 'tuning knob' becomes just that, and allows operator (or automatic) adjustment of the controller gain (ie. pole position) to achieve a desired transient performance.

SYSTEM MODELS AND CONTROLLER DESIGN

There are a number of ways of deriving a self-tuning control law with the structure of a PID controller. Two approaches that are analyzed and compared in the paper are described below.

1. Process Model and Disturbance Representation

Consider a discrete time, SISO plant, that can be represented by:

$$A(z^{-1})y(t) = z^{-k}B(z^{-1})u(t) + x(t) \qquad (1)$$

The signal $x(t)$ models the disturbances and noise acting on the process and can be represented as:

$$x(t) = C(z^{-1})e(t) \qquad (2)$$

For the most common types of noise and process disturbances encountered in practice, $x(t)$ is nonstationary. A more appropriate model for the noise term is therefore (Box and Jenkins, 1970; MacGregor et al, 1975; Tuffs and Clarke, 1985):

$$\Delta x(t) = C(z^{-1})e(t) \qquad (3)$$

where Δ is the differencing operator $(1-z^{-1})$. Combining eqn. (3) with eqn. (1) gives the CARIMA model of the plant as:

$$A(z^{-1})y(t) = z^{-k}B(z^{-1})u(t) + C(z^{-1})e(t)/\Delta \qquad (4)$$

The design of control strategies (adaptive or otherwise) based on this representation leads to systems with inherent integral action. However, in a deterministic framework a more general and natural way of removing offset is to use the internal model principle of Francis and Wonham (1976), eg. Song et al (1984). The advantage of

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