ANALYTICAL CONTROLLER DESIGN OF INTEGRATING AND FIRST ORDER UNSTABLE TIME DELAY PROCESS

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Abstract: The design of the PID controller cascaded with first order filter has been proposed for the integrating and first order unstable time delay process. The design algorithm is based on the IMC criterion which has single tuning parameter to adjust the performance and robustness of the controller. The setpoint filter is used to diminish the overshoot in servo response. The simulation results of the suggested method are compared with other recently published tuning methods to demonstrate the superiority of the proposed method. For the reasonable comparison the controllers are tuned to have the same degree of robustness by the measure of maximum sensitivity and the robustness of the controllers also investigated for model mismatch, where proposed method has clear advantage. For the ease of the selection of closed-loop time constant (λ), a guideline is provided at two different robustness levels for a broad range of θ/τ ratio. *Copyright © 2007 IFAC*

Keywords: Unstable process, Integrating delay process, PID controller, Disturbance rejection, Time delay, Distillation columns control

1. INTRODUCTION

Proportional-integral-derivative (PID) control has three-term functionality offering treatment of both transient and steady-state responses; it provides a generic and efficient solution to real world control problems. The wide application of PID control has stimulated and sustained research and development to "get the best out of PID'', and "the search is on to find the next key technology or methodology for PID tuning".

The numerous important chemical processing units are open-loop unstable process in industrial and chemical practice are well known to be difficult to control especially when there exists time delay, such as continuous stirred tank reactors, polymerization reactors and bioreactors are inherently open loop unstable by design. Furthermore, many of the processes are usually run batch-wise, and as a result of possible formulation changes, may operate with significant batch-to-batch variability. Clearly, the tuning of controllers to stabilize these processes and impart adequate disturbance rejection is critical. On the other hand the integrating processes are also frequently encountered in the process industries and many of the researchers suggested that considerable numbers of chemical processes can be modeled for the purpose of designing controller by an integrating process with time delay. Consequently, there has been much recent interest in the literature on tuning the industrially standard PID controllers for open-loop unstable systems and integrating processes. The effectiveness of internal model control (IMC) design principle has attracted in process industry, which causes many efforts made to exploit the IMC principle to design the equivalent feedback controllers for stable and unstable processes (Morari and Zafiriou, 1989). The IMC based PID tuning rules have the advantage of only one tuning parameter to achieve a clear trade-off between closed-loop performance and robustness. It is well know that the IMC structure is very powerful for controlling stable processes with time delay and cannot be directly used for unstable processes by reason of the internal instability (Morari and Zafiriou, 1989), some modified IMC methods of two-degree-of-freedom (2DOF) control such as Lee *et al*. (2000), Yang *et al*. (2002), Wang and Cai (2002), Tan *et al*. (2003), Liu *et al*. (2005) had been developed for controlling unstable processes with time delay. In addition, 2DOF control methods based on the Smith-Predictor (SP) had been proposed by Majhi & Atherton (2000), Zhange *et al.* (2004) and achieved smooth nominal

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setpoint response without overshoot for first order unstable processes with time delay. It is a notable merit that the nominal setpoint response tends to be faster without overshoot for unstable process according to either the modified IMC methods or the modified SP methods. In fact, the common characteristic of the abovementioned modified IMC and SP methods is utilizing the nominal process model in their control structures, which effectively contributes to acquire the above merit. It should be noted that most existing 2DOF control methods restricted attention on unstable processes modeled in the form of a first order rational part plus time delay, which in fact, cannot represent a variety of industrial and chemical unstable processes well enough. Besides, there usually exist the process unmodeled dynamics that inevitably tend to deteriorate the control system performance. The delay integrating process has clear advantage in the identification test, because the model contains only two parameters and simple for identification. Some of the well acceptable PID tuning methods for the delay integrating process are Chien and Fruehauf (1990); Lubyen (1996) and Chen & Seborg (2002).

It is well known that in recent time the controller hardware support the microprocessor implementation for the PID cascaded filter. Therefore, PID controller cascaded with first order lead lag filter can be easily implemented using the modern control hardware. The important reason for using the controller in this form is availability of these facilities in the present microprocessor implementation to achieve the better performance. Many authors have suggested the PID controller cascaded with first order filter either to get PID structure or for the better performance. The PID controller cascaded first or second order filter type structure in Eq. (1) are suggested by Rivera *et al.* (1986), Lee *et al.* (1998), Horn *et al.* (1996) and Dwyer (2003).

$$G_c = K_c\left(1+\frac{1}{\tau_I s}+\tau_D s\right)\frac{1+as}{1+bs} \qquad (1)$$

where K_c , τ_I and τ_D are the proportional gain, integral time constant, and derivative time constant of the PID controller, respectively, and a & b are the filter parameters. It has essential to emphasize that design principle of the aforementioned tuning methods for the unstable and delay integrating process is either complicated or providing modified IMC structure which is difficult to implement in the real process plant.

Therefore, in the present study a simple method has been proposed for the design of the PID controller cascaded with first order filter to accomplish the improved performance for the first order unstable and delay integrating processes. A closed-loop time constant (λ) guidelines has been recommend cover a wide range of θ/τ ratio. Simulation study has been performed to show the superiority of the proposed method for both the nominal and perturbed processes.

2. DESIGN PROCEDURE

The IMC controller (Fig. 1-a) has been shown to be a powerful method for control system synthesis (Morari and Zafiriou, 1989). However, for unstable processes the IMC structure cannot be implemented, since any input *d* will make *y* grow without bound if G_p is unstable. Nevertheless, as discussed in (Morari and Zafiriou, 1989), we could still use IMC approach to design a controller for an unstable process, if only the following conditions are satisfied for the internal stability of the closed-loop system:

(i) q stable.

(ii) $G_p q$ stable.

(iii) $(1-G_p q)G_p$ stable.

These conditions result in the well known standard interpolation conditions (Morari and Zafiriou, 1989):

- If the process model G_p has unstable poles, $up_1, up_2, \cdots, up_m$, then q should have zeros at $up_1, up_2, \cdots, up_m$.
- If the process model G_D has unstable poles, $dup_1, dup_2, \cdots, dup_m$, then $1-\tilde{G}_p q$ should have zeros at $dup_1, dup_2, \cdots, dup_m$.

Since the IMC controller q is designed as $q = p_m^{-1} f$ in which p_m^{-1} includes the inverse of the model portion, the controller satisfies the first condition. The second condition could be satisfied through the design of the IMC filter f . For this, the filter is designed as

$$f = \frac{\sum_{i=1}^{m} \alpha_i s^i + 1}{(\lambda s+1)^r} \qquad (2)$$

where *r* is the number of poles to be canceled; α_i are determined by Eq. (3) to cancel the unstable poles in G_D ; r is selected large enough to make the IMC controller proper.

$$1-\tilde{G}_p q\Big|_{s=dup_1, dup_2, \cdots, dup_m} = \left|1-\frac{p_m^{-1}(\sum_{i=1}^{m} \alpha_i s^i+1)}{(\lambda s+1)^r}\right|_{s=dup_1, dup_2, \cdots, dup_m} = 0 \qquad (3)$$

Then, the IMC controller comes to be

$$q = p_m^{-1}\frac{(\sum_{i=1}^{m} \alpha_i s^i+1)}{(\lambda s+1)^r} \qquad (4)$$

Thus, the resulting setpoint and disturbance rejection is obtained as (nominal case *i.e.*, $G_P = \tilde{G}_P$):

$$\frac{y}{r} = G_p q = \frac{p_A\left(\sum_{i=1}^{m} \alpha_i s^i+1\right)}{(\lambda s+1)^r} \qquad (5)$$

$$\frac{y}{d} = (1-G_p q)G_D = \left(1-\frac{p_A\left(\sum_{i=1}^{m} \alpha_i s^i+1\right)}{(\lambda s+1)^r}\right)G_D \qquad (6)$$

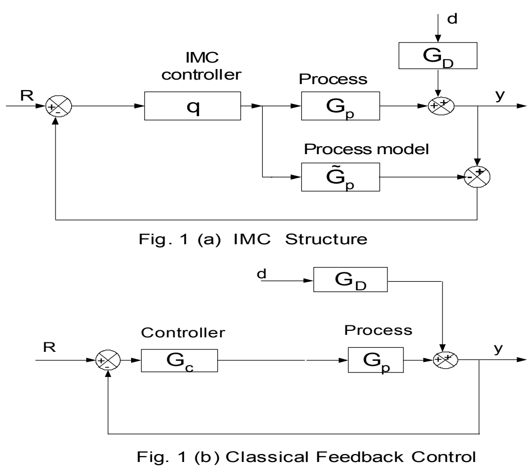

Fig. 1. Block diagram of classical feedback control

The numerator expression $\left(\sum_{i=1}^{m}\alpha_i s^i+1\right)$ in Eq. (5) causes an unreasonable overshoot in the servo response, which can be eradicated by adding the setpoint filter f_R as:

$$f_R=\frac{1}{\left(\sum_{i=1}^{m}\alpha_i s^i+1\right)} \tag{7}$$

From Figs. (1-a) and (1-b), a feedback controller G_c which is equivalent to the IMC controller q is represented by

$$G_c=\frac{q}{1-\tilde{G}_P q} \tag{8}$$

The resulting ideal feedback controller obtained as

$$G_c=\frac{p_m^{-1}\dfrac{\left(\sum_{i=1}^{m}\alpha_i s^i+1\right)}{(\lambda s+1)^r}}{1-\dfrac{p_A\left(\sum_{i=1}^{m}\alpha_i s^i+1\right)}{(\lambda s+1)^r}} \tag{9}$$

The aforementioned resulting controller in Eq. (9) does not have a standard PID controller configuration. The remaining objective is to design the PID controller cascaded with first order filter that resemble the equivalent feedback controller most closely and is discussed in the next section.

3. PROPOSED TUNING RULE

The first order delay unstable process (FODUP) is the typical representative model which is commonly utilized in the chemical process industries. Consequently this section comprises the design of the tuning rule for FODUP and it also extended for the delay integrating process (DIP).

3.1. First-Order Delay Unstable Process (FODUP)

$$G_P=G_D=\frac{Ke^{-\theta s}}{\tau s-1} \tag{10}$$

where K is the gain, τ the time constant and θ is time delay. The IMC filter structure exploited is given as:

$$f=\frac{\alpha s+1}{(\lambda s+1)^3} \tag{11}$$

The resulting IMC controller can be obtained as follows

$$q=\frac{(\tau s-1)(\alpha s+1)}{K(\lambda s+1)^3} \tag{12}$$

The IMC controller in the Eq. (12) is proper and the ideal feedback controller which is equivalent to the IMC controller is:

$$G_c=\frac{(\tau s-1)(\alpha s+1)}{K\left[(\lambda s+1)^3-e^{-\theta s}(\alpha s+1)\right]} \tag{13}$$

Approximating the dead time $e^{-\theta s}$ with a 1/2 Pade expansion

$$e^{-\theta s}=\frac{(6-2\theta s)}{\left(6+4\theta s+\theta^2 s^2\right)} \tag{14}$$

It is important to note that the 1/2 Pade approximation is precise enough to convert the ideal feedback controller into a PID cascaded first order filter with barely any loss of accuracy as well as retain the desired controller form. Therefore, dead time in Eq. (13) is approximated with 1/2 Pade expansion and the resulting G_c is

$$G_c=\frac{(\tau s-1)(\alpha s+1)\left(6+4\theta s+\theta^2 s^2\right)}{K\left[(\lambda s+1)^3\left(6+4\theta s+\theta^2 s^2\right)-(\alpha s+1)(6-2\theta s)\right]} \tag{15}$$

Expanding and rearranging the above Eq. (15)

$$G_c=\frac{(-\tau s+1)\left(6+4\theta s+\theta^2 s^2\right)(\alpha s+1)}{-K(6\theta+18\lambda-6\alpha)s\left[1+\frac{\left(2\alpha\theta+\theta^2+12\lambda\theta+18\lambda^2\right)}{(6\theta+18\lambda-6\alpha)}s+\frac{\left(3\lambda\theta^2+12\lambda^2\theta+6\lambda^3\right)}{(6\theta+18\lambda-6\alpha)}s^2+\frac{\left(3\lambda^2\theta^2+4\lambda^3\theta\right)}{(6\theta+18\lambda-6\alpha)}s^3\frac{\lambda^3\theta^2}{(6\theta+18\lambda-6\alpha)}s^4\right]} \tag{16}$$

The analytical PID formula can be obtained by rearranging the above Eq. (16) and presented as:

$$K_C=-\frac{4\theta}{K(6\theta+18\lambda-6\alpha)};\quad \tau_I=2\theta/3;\quad \tau_D=\theta/4;\quad a=\alpha \tag{17}$$

The parameters b in filter can be obtained by equating the remaining part of the denominator of Eq. (16) with the process pole and filter $(bs+1)$. Since the remaining part of the denominator of Eq. (16) contains the factor of the process pole, filter $(bs+1)$ and a high order polynomial terms in s. The high order polynomial term in s has barely any impact because it is not in control relevant frequency range. Therefore filter parameters can be obtained as

$$(bs+1)=\frac{1+\frac{\left(2\alpha\theta+\theta^2+12\lambda\theta+18\lambda^2\right)}{(6\theta+18\lambda-6\alpha)}s+\frac{\left(3\lambda\theta^2+12\lambda^2\theta+6\lambda^3\right)}{(6\theta+18\lambda-6\alpha)}s^2+\frac{\left(3\lambda^2\theta^2+4\lambda^3\theta\right)}{(6\theta+18\lambda-6\alpha)}s^3\frac{\lambda^3\theta^2}{(6\theta+18\lambda-6\alpha)}s^4}{(-\tau s+1)} \tag{18}$$

Taking the first derivative of the above Eq. (18) and substituting $s=0$, the parameter b can be easily obtained below

$$b=\frac{\left(2\alpha\theta+\theta^2+12\lambda\theta+18\lambda^2\right)}{(6\theta+18\lambda-6\alpha)}+\tau \tag{19}$$

The G_c in Eq. (15) contains the RHP (zero) which will eliminate from the controller after factorization and cancel out with remaining part of Eq. (16). The value of the extra degree of freedom α is selected so that it cancels out the open-loop unstable pole at $s=1/\tau$. This means certainly adopt α so that the term $[1-Gq]$ has a zero at the pole of G_D. That required $[1-Gq]\big|_{s=1/\tau}=0$ and $\left[1-(\alpha s+1)e^{-\theta s}/(\lambda s+1)^3\right]_{s=1/\tau}=0$. The value of α is obtained after some simplification

$$\alpha=\tau\left[\left(1+\frac{\lambda}{\tau}\right)^3 e^{\theta/\tau}-1\right] \tag{20}$$

3.2. Delayed Integrating Process (DIP)

$$G_p=G_D=\frac{Ke^{-\theta s}}{s} \tag{21}$$

The DIP can be modeled by considering the integrator as an unstable pole near zero. This is necessary because it is not possible to apply the aforementioned IMC procedure for DIP, since the term of α disappears at $s=0$. Therefore, DIP can be approximated to FODUP as below:

$$G_p=G_D=\frac{Ke^{-\theta s}}{s}=\frac{Ke^{-\theta s}}{s-1/\psi}=\frac{\psi Ke^{-\theta s}}{\psi s-1} \tag{22}$$

where ψ is an arbitrary constant with a sufficiently large value. Accordingly, the optimum filter structure for DIP is same as that for the FODUP model, i.e., $f=(\alpha s+1)/(\lambda s+1)^3$. Therefore, the resulting IMC controller becomes $q=(\psi s-1)(\alpha s+1)/K\psi(\lambda s+1)^3$ and the consequently PID tuning rules are obtained follows:

$$K_C=-\frac{4\theta}{K\psi(6\theta+18\lambda-6\alpha)};\quad \tau_I=2\theta/3;\quad \tau_D=\theta/4;\quad a=\alpha \tag{23a}$$

$$b = \frac{\left(2\alpha\theta + \theta^2 + 12\lambda\theta + 18\lambda^2\right)}{\left(6\theta + 18\lambda - 6\alpha\right)} + \psi \qquad (23b)$$

$$\alpha = \psi\left[\left(1 + \frac{\lambda}{\psi}\right)^3 e^{\theta/\psi} - 1\right] \qquad (23c)$$

4. SIMULATION RESULTS

This section deals with the simulation study for the two different examples and the results are compared with some of the recently reported methods. To evaluate the robustness of a control system, the maximum sensitivity, Ms, which is defined by $Ms = \max\left|1/[1 + G_p G_c(iw)]\right|$, is used. Since the Ms is the inverse of the shortest distance from the Nyquist curve of the loop transfer function to the critical point $(-1,0)$, a small Ms value indicates that the stability margin of the control system is large. The Ms is a well known robustness measure and is used by Chin and Seborg, (2002). Therefore, throughout all our simulation examples, all of the controllers compared were designed to have the same robustness level in terms of the maximum sensitivity.

4.1. Example 1. FODUP

A widely published example of a FODUP has been considered for the comparisons (Lee *et al.,* 2000; Tan *et al.*, 2003, Liu *et al.,* 2005) is:

$$G_p = G_D = \frac{1e^{-0.4s}}{1s - 1} \qquad (24)$$

In the recently published paper of Liu *et al.* (2005) which had already demonstrated its superiority over many widely accepted previous approaches (Tan *et al.* 2003 and Majhi & Atherton, 2000). The proposed method is compared with the Lee *et al.* (2000) and Liu *et al.* (2005).

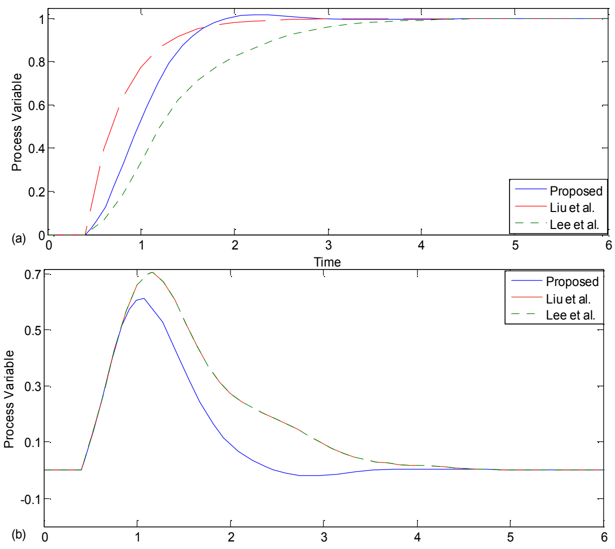

Fig. 2. Simulation results for Example 1 (FODUP)

The three controller parameters for Liu *et al.* (2005) method were taken as $K_c = 2$, $C(s) = (s+1)/(0.4s+1)$ and $\lambda = 0.5$, they suggested the disturbance estimator $F(s) = 2.634 + \frac{1}{0.9566s} + 0.4058s$. For the fair comparison λ has been adjusted for each tuning methods which gives the same Ms, because Ms is well know robustness measure and used by many researchers. In order to achieve the same $Ms = 3.03$ with Liu *et al.* (2005), for the proposed method $\lambda = 0.20$ has been adjusted and corresponding tuning parameters are $K_c = 0.4615$, $\tau_I = 0.2667$, $\tau_D = 0.1$, $a = 1.5779$, $b = 0.1053$ and $f_R = 1/(1.5779s + 1)$. The tuning parameters for the Lee *et al.* (2000) method is identical with Liu *et al.* (2005) disturbance estimator at the same Ms value. It is very obvious that the disturbance estimator design of the Liu *et al.* (2005) method is exactly identical with the Lee *et al.* (2000), but the setpoint response is different in both cases, because both of them have different approach. Lee *et al.* (2000) PID controller setting are $K_c = 2.634$, $\tau_I = 2.5197$, $\tau_D = 0.1541$ and setpoint filter $f_R = 1/(2.3566s + 1)$.

Figure 2(a) and 2(b) show the comparison of the proposed method with Liu *et al.* (2005) and Lee *et al.* (2000), by introducing a unit step change in the setpoint and a unit step input in the load disturbance respectively. For the servo response the setpoint filter is used for both the proposed and Lee *et al.* (2000) methods whereas three control element structure is used for the Liu *et al* (2005).

It is clear from the Fig. (2), the proposed method results in the improved load disturbance response. For the servo response the Liu *et al.* (2005) methods appears better but the settling time of the Liu *et al.* (2005) method and the proposed method is almost similar. In Fig. (2-a) Lee *et al.* (2000) response is very slow and it requires long settling time. It is important to note that in the well known modified IMC structure has theoretical advantage of eliminating the time delay from the characteristic equation. Unfortunately, this advantage is lost if the process model is inaccurate. Besides, there usually exists the process unmodeled dynamics in real process plant that inevitably tends to deteriorate the control system performance severely. For the disturbance rejection the proposed methods has big advantage over other methods as shown in Fig. (2-b). As discussed the disturbance estimator design of the Liu *et al.* (2005) and Lee *et al.* (2000) method is exactly similar which cause the same PID tuning setting and consequently same response for disturbance rejection and have overlapping in Fig. (2-b).

Despite the fact that the comparison has been performed on the same robustness by equalizing the Ms of the compared tuning methods by adjusting their λ value. It is worthwhile to check the robustness of the controller by inserting a perturbation uncertainty of 10% in all three parameters simultaneously to obtain the worst case model mismatch, i.e., $G_p = G_D = 1.1e^{-0.44s}/(0.9s - 1)$. The simulation results for the proposed and other tuning methods are presented in Fig. (3) for both the set-point and the disturbance rejection. It is clear from Fig. (3) that the proposed controller tuning method has the best setpoint as well as load response while the modified IMC controller structure which contains the three-element controller of the Liu *et al.*(2005) method has worst response for the model mismatch for the setpoint. As we have seen in the setpoint response for the nominal case the Liu *et al.*(2005)

has smooth and fast response, which is achieved by the sacrificing the robustness of the closed-loop system. The model mismatch case for the disturbance rejection of Liu *et al.* (2005) and Lee *et al.* (2000) is identical and overlapping. The setpoint response of Lee *et al.* (2000) is achieved by simple feed back controller with setpoint filter as in the proposed study also. The performance and robustness from the above study obviously exhibit that the proposed method has better nominal as well as robust performance among others.

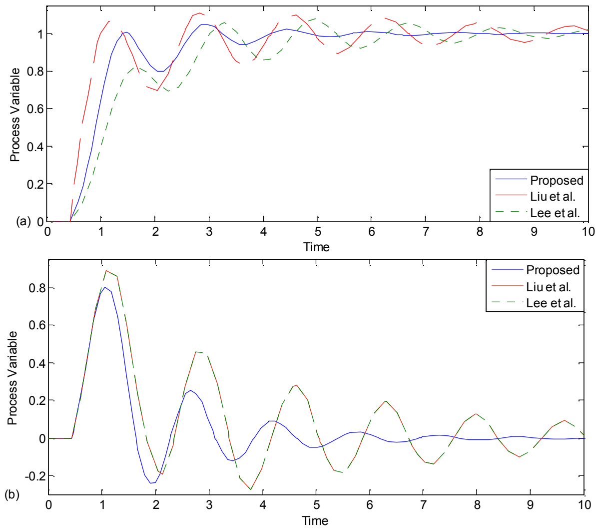

Fig. 3. Perturbed response for Example 1 (FODUP)

4.2. Example 2. DIP (Distillation Column Model)

The distillation column model studied by Chien & Fruehauf (1990) and Chen & Seborg (2002) was considered for the present study. The distillation column separates a small amount of a low-boiling material from the final product. The bottom level of the distillation column is controlled by adjusting the steam flow rate. The process model for the level control system is represented as the following DIP model which can be approximated by the FODUP model as follows:

$$G_p = G_D = \frac{0.2e^{-7.4s}}{s} = \frac{20Ke^{-7.4s}}{100s-1} \quad (25)$$

The proposed method, Chen and Seborg (2002), and Lee *et al.* (2000) was used to design the PID controllers, as shown in Fig. (4) and λ was selected $5.56, 9.15,$ and 11.0 respectively, which have resulting $Ms = 1.90$ for each tuning rule. Figure (4) shows the output response, where the proposed tuning rule result in the least settling time for both the servo and disturbance rejection and followed by Chen and Seborg (2002). Lee *et al.*'s (2000) method has the slowest response and requires the maximum settling time for both the setpoint and disturbance rejection. On the basis of Fig. (4), it is clear that the proposed method performs better than the other conventional methods for both the servo as well as regulatory performance.

The robustness of the controller is evaluated by inserting a perturbation uncertainty of 50% in the gain and 20% in the dead time simultaneously towards the worst case model mismatch. The resulting worst case plant-model mismatch after perturbation is obtained as $G_p = G_D = 0.3e^{-8.88s}/s$. The simulation results for plant-model mismatch are given in Fig. (5) for both servo and regulatory problem. It needs to clarify that the controller settings are those calculated for the process with nominal process parameters. The responses indicate that the proposed method has less oscillatory response for both disturbance rejection and setpoint and required less settling time. Chen and Seborg (2002), method has more oscillation and followed by Lee *et al.* (2000). It seems that the proposed method clearly gives good performance, even for high process uncertainties. Also, the proposed method is more robust than other tuning rules for large uncertainness in process parameters.

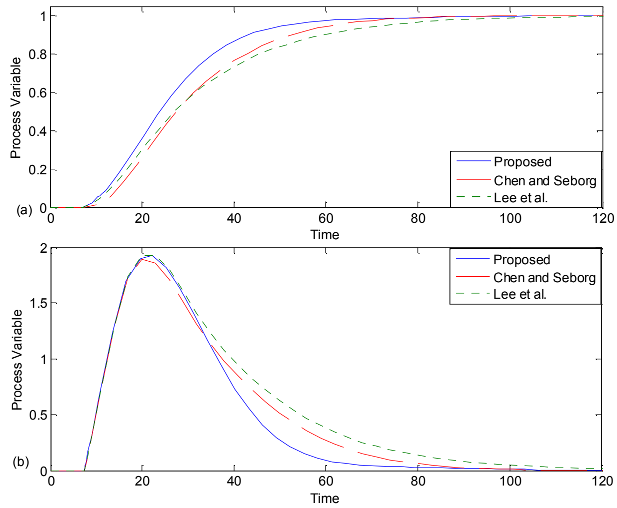

Fig. 4. Simulation results for Example 2 (DIP)

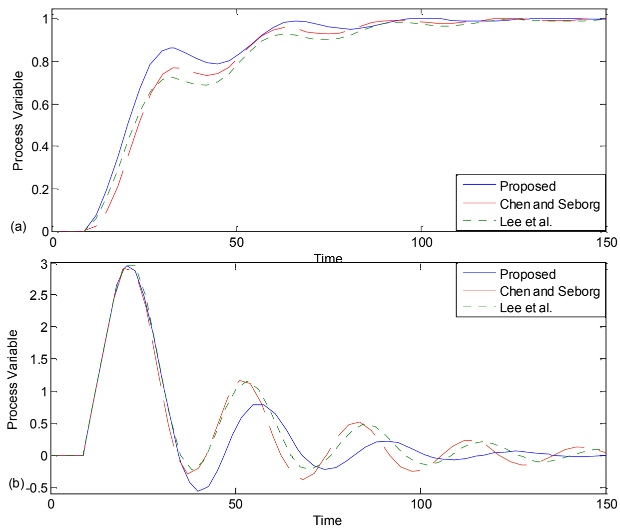

Fig. 5. Perturbed response for Example 2 (DIP)

4.3. Closed-loop time constant λ guidelines

The closed-loop time constant λ is an only one user-defined tuning parameter in the proposed tuning rule. It is directly related to the performance and robustness of the proposed tuning method, which is why it is important to have some λ guidelines in order to provide both a fast and robust performance for a desirable range of θ/τ ratio.

Figure (6) shows the plot of λ/τ versus θ/τ ratios for FODUP. The presented λ guideline is for the nominal model for $Ms = 3.0$ and 3.6 values. It is important to mention that the proposed tuning method is applicable for dead time dominant process also. The λ guideline is not extended for the lager θ/τ value in the Fig. (6) because it is difficult to obtain the above suggested Ms value for the large θ/τ ratio. So, for the $\theta/\tau > 0.6$, based on many simulation studies, it is observed that the starting value of λ can be considered to be equal as process time delay, which can gives robust control performance. If not, the value should be increased carefully until both the nominal and robust control performances are achieved.

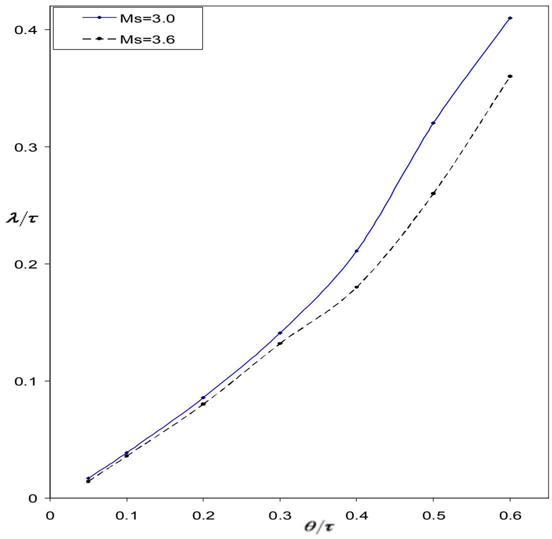

Fig. 6. λ guidelines for FODUP

5. CONCLUSIONS

A simple design method of the analytical PID cascaded filter tuning method has been proposed based on the IMC principle. Two important representative processes have been considered in the present study which is frequently used in the chemical process industries. The proposed method has excellent improvement in both setpoint and disturbance rejection for the FODUP and DIP process. The simulation has been conducted for the fair comparison when the various controllers were tuned to have the same degree of robustness by the measure of Ms value. The robustness study has been conducted by inserting a perturbation uncertainty in all parameters simultaneously to obtain the worst case model mismatch, where proposed study has clear advantage. The closed-loop time constant λ guideline was also proposed for over a wide range of θ/τ ratio.

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