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Set-Point Tracking Analysis of Controllers Designed for a Trickle-Bed Catalytic Reactor

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ABSTRACT: Set-point tracking is one of the key performance benchmark for a controller. In an industrial process, it is important to choose the optimal controller that offers the best performances. In this work, set-point tracking of 3 different controllers designed by different methods has been evaluated by comparing the closed loop responses for a selected third order delayed model.

Keywords: Catalytic Reactor, TOPDT, Delay, process control.

1 INTRODUCTION

The evaluation of a controller's performance is the most important subject of controller's analysis. A controller capable of rejecting disturbance efficiency and exhibits rapid and smooth responses to any variation is considered a good controller. PID controllers are well known for their disturbance rejection capabilities and provide excellent set-point.

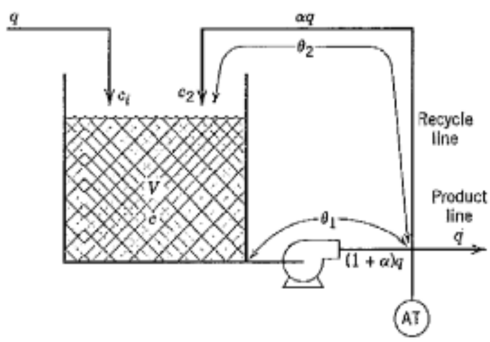

Fig. 1. A trickle-bed reactor with recycle line

In this present work, a "trickle-bed catalytic reactor" process is selected. It is a chemical reactor that uses the downhill progress of liquid and the down or upward progress of gas over a crowded bed of particles. It is the simplest reactor that performs catalytic reactions where a gas and liquid are present in the reactor. It is widely used in processing plants [1-3]. Trickle-bed catalytic reactor is represented by Fig.1.

where, AT: analyzer transmitter

Θ_1: delay time for material flow from reactor outlet to the composition analyzer

Θ_2: delay time for material flow from analyzer to reactor inlet

The parameter values are selected as –

$V = 5\ m^3$, $\alpha = 12$, $q = 0.05\ m^3/min$, $\Theta_1 = 0.9$ min, $k = 0.04\ min^{-1}$, $\Theta_2 = 1.1$min

The transfer function used in this process is a Third Order plus Delay Time (TOPDT), and it has the following form [4]:

$$G(s) = \frac{0.2(s+1)e^{-0.9s}}{s(25s+1)(0.8s+1)} \tag{1}$$

Easy implementation and operation of PID controllers makes it extensively applicable in process manufacturing, to control various industrial processes [4]. In the literature, scarce PID tuning techniques are available for TOPDT model compared to First Order plus Delay Time (FOPDT) model [5, 6].

It is required of process control that the tuning methods should be simple and easy to implement which require little information and give moderate performance. A tuning method should be widely applicable [7].

II. METHODOLOGY

To design the appropriate controllers for this process, flowchart to be implemented is given by Fig. 2. The transfer function of the process after approximating delay with Skogestad's half rule approximation is given by:

$$G(s) = \frac{0.2e^{-0.3s}}{s(25.4s+1)} \tag{2}$$

It is a SOPDT with an integrator. This integrator doesn't allow us to use the Skogestad's half rule approximation for a second time in order to get a FOPDT model. Therefore, IMC (internal model control) tuning technique is used to design a PID controller for the approximated SOPDT model. Approximated SOPDT model type one is given in general as:

$$G(s) = \frac{Ke^{-\theta s}}{s(1+\tau s)} \tag{3}$$

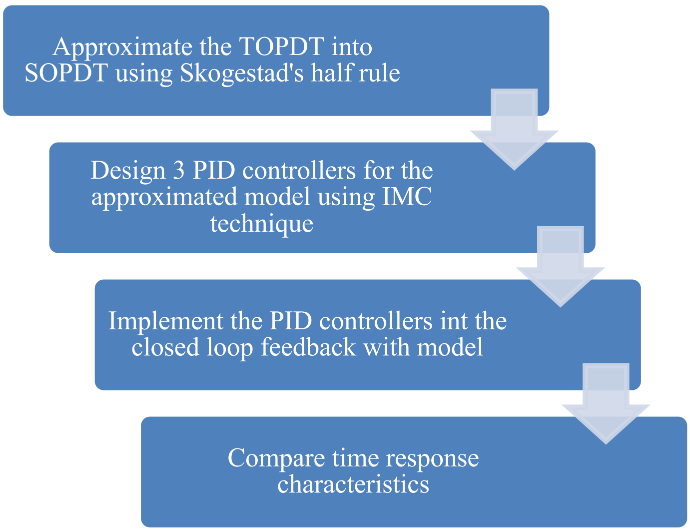

Fig. 2. Methodology

Comparing with approximated model, we have k = 0.2; ϴ = 0.3; τ = 25.4;
In that case, the value of the controller component can be calculated with the following equations:

$$K_c = \frac{2\tau_c + \tau + \Theta}{k(\tau_c + \Theta)^2} \quad (4)$$

$$\tau_i = 2\tau_c + \tau + \Theta \quad (5)$$

and

$$\tau_d = \frac{(2\tau_c + \Theta)\tau}{2\tau_c + \tau + \Theta} \quad (6)$$

The choice of design parameter τc is a key decision in both the DS and IMC design methods. For general process models with a time constant which is dominant, guideline can be generalized to [7]

$$\tau_{dom} > \tau_c > \Theta$$

Table 1. PID controller settings for different values of τ_c

Parameter	K_c	τ_i	τ_d
C1 : Controller with τ_c = 3	14.5546	31.7	5.0479
C2 : Controller with τ_c = 6	4.7493	37.7	8.2870
C3 : Controller with τ_c = 12	1.6425	49.7	12.4189
C4 : Controller with τ_c = 17	0.9973	59.7	14.5932
C5 : Controller with τ_c = 21	0.7461	67.7	15.8703
C6 : Controller with τ_c = 25	0.5913	75.7	16.8774

III. RESULTS AND DISCUSSION

From fig. 3, it can be depicted that the response C1 (yellow) has the best performances compared to C2 and C3. Its rise and settling time are far better and it doesn't exceed the step value a lot.

The three controllers have approximately the similar performances, but C4 controller is slightly better performing as it rise and settling time are less than the two other controllers, as shown in fig. 4.

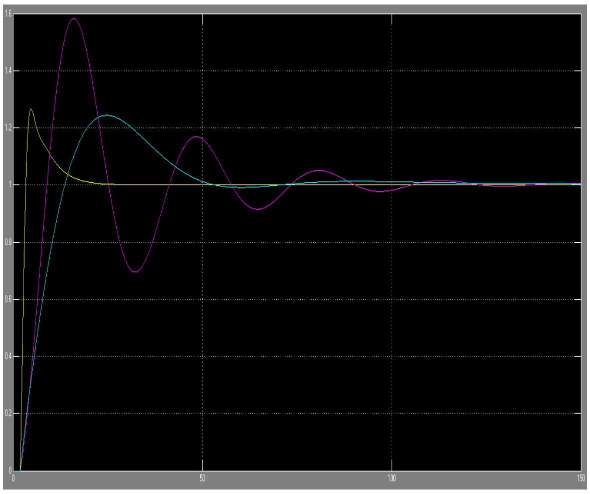

Fig. 3. Closed loop step response of C1 (yellow), C2 (pink) and C3 (blue)

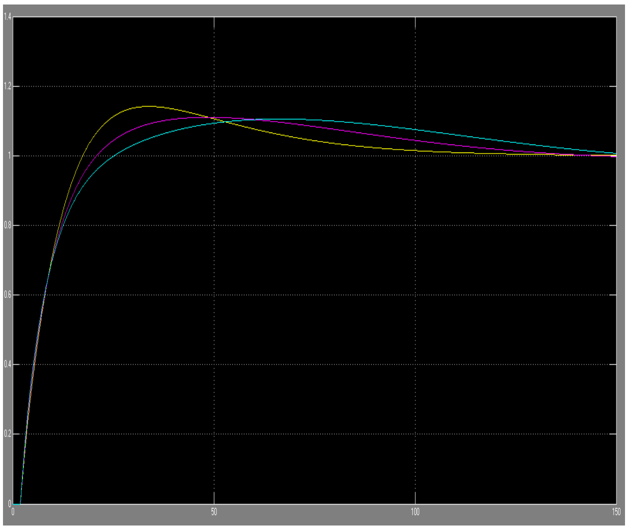

Fig. 4. Closed loop step response of C4 (yellow), C5 (pink) and C6 (blue)

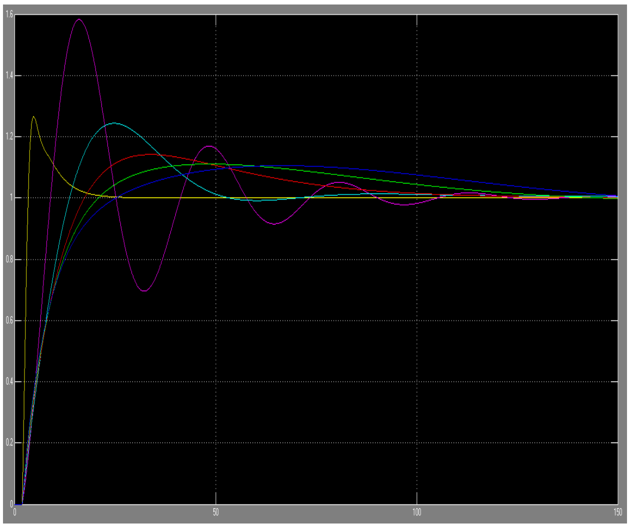

Fig. 5. Closed loop step response of all the controllers C1 (yellow), C2 (pink) and C3 (blue) C4 (yellow), C5 (pink) and C6 (blue)

TABLE 2: COMPARISON OF CONTROLLER PERFORMANCES

Controller	Rise Time	Settling Time	% Overshoot
C1 (τ_c = 3)	3.41	13.2	26.6
C2 (τ_c = 6)	8.96	81.6	58
C3 (τ_c = 12)	13.68	44.4	24.5
C4 (τ_c = 17)	17.42	71.5	14
C5 (τ_c = 21)	20.91	96	11
C6 (τ_c = 25)	25.25	116.7	10.5

It is evident from Table 2, except for C2 controller, the smaller is τ_c, the better are the controllers. Indeed, the more τ_c decreases, the rise and settling time decreases as well. On the other hand, the overshoot of the controller is more in smaller τ_c.

IV. CONCLUSION

To choose the most performing controller, some compromises are required. It depends on the system and its ability to accept the overshoot of the corrector. In this case, the maximum overshooting value is 26.6 %. For a trickle-bed catalytic reactor, this much overshoot can be handled without causing any damage. Indeed, the appropriate controller for this TOPDT system is the C1 controller with $\tau_c = 3$.

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