Analysis of different control strategies for different industrial processes using SIMULINK

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Analysis of different control strategies for different industrial processes using SIMULINK

A Thesis submitted in partial fulfillment

of the requirements for the award of the degree of

Bachelor of Technology

In

Electronics & Instrumentation Engineering

In

May 2013

to the department of

Electronics & Communication Engineering
National Institute of Technology, Rourkela
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CERTIFICATE

This is to certify that the Thesis Report entitled "**ANALYSIS OF DIFFERENT CONTROL STRATEGIES FOR DIFFERENT INDUSTRIAL PROCESSES**" submitted by **KAMAL PRAKASH** and **RAVI RAJ** bearing roll no. **110EI0248** and **110EI0250** respectively, in partial fulfilment of the requirements for the award of Bachelor of Technology in Electronics and Instrumentation Engineering carried out during the academic session 2013-2014 at National Institute of Technology, Rourkela is an authentic work carried out by them under my supervision and guidance.

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Acknowledgment

This project work would have not been possible without the guidance and the help of several individuals who in one way or another, contributed and extended their valuable assistance in our project work.

Our utmost gratitude to **Prof. Tarun Kumar Dan**, our dissertation adviser whose sincerity and encouragement we will never forget. He has been our inspiration as we hurdle all the obstacles in the completion of this project work and has supported us throughout the project work with patience and knowledge.

Sincere thanks to people, faculty and non-technical staffs of Dept. of Electronics & Communication Engineering, NIT Rourkela who have encouraged us throughout the project work of our Bachelor's Degree.

We would like to thank all friends for their help during the course of this project work. And finally thanks to our parents, our brothers (who has been a constant source of inspiration) and our lovely sisters whose faith, patience and teaching had always inspired us to work upright in life. Without all these beautiful people our world would have been an empty place.

Once again, we especially thank Prof. T.K. Dan. It was a great pleasure for us to conduct the project under his supervision.

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Abstract

This project work analyses the different control strategies for different types of industrial process systems. The design and implementation of these processes are done in the SIMULINK software. Firstly, we need mathematical models of the systems before the analysis of processes. For this we generally use Laplace transfer and linearization which is carried out in each and every system considered in this project work. We have distinguished higher order systems on the basis of Non-interacting & Interacting system.

In level process, a single-tank, a two-tank, and three-tank systems are analyzed. The objective of these systems is to maintain the process liquid at a desired level or change it at a pre - determined rate by controlling and using the manipulated variables.

This project work also focusses on the various PID controller design methods such as Zeigler-Nichol method, Cohen-Coon method, Tyreus-Luyben method, Internal Model Control (IMC) and Minimum error criteria (IAE, ISE) method. The performances of the PID tunings methods for the different control strategies are also used to compare different processes. The implementation of IMC based PID controller for handling unstable processes with dead time (SOPTD) is also done.

SIMULINK in MATLAB is used to simulate all these processes. The response curves of these systems are generated to different forcing functions.

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CHAPTER 1:

INTRODUCTION:

This project work breaks down the diverse control techniques for distinctive sorts of mechanical process systems. The design and implementation of these processes are carried out in the SIMULINK programming. Firstly, we require mathematical models of the systems before the analysis of processes. For this we by and large utilize Laplace Transfer and linearization which is done in every single system recognized in this project work. We have recognized higher order systems on the premise of Non-Interacting & Interacting systems.

In level process, a single-tank, a two-tank, and three-tank systems are dissected. The target of these systems is to keep up the process fluid at a desired level or transform it at a decided ahead of time rate by controlling and utilizing the manipulated variables.

This project work likewise focuses on the different PID controller strategies, for example, Ziegler-Nichols technique, Cohen-Coon system, Tyreus-Luyben strategy, Internal Model Control (IMC) and Minimum error criteria (IAE, ISE) system. The exhibitions of the PID tunings techniques for the distinctive control methods are additionally used to compare different processes. The usage of IMC based PID controller for taking care of unstable processes with dead time (SOPTD) is likewise done.

SIMULINK in MATLAB is utilized to simulate all these processes. The response curve of these systems are created to distinctive forcing functions.

1.1 CONTROL SYSTEM

Control is paramount for some reasons. Those that take after are not by any means the only ones, yet we feel they are the most paramount. They are focused around our modern experience, and we might want to pass them on. Control is vital to

1. Avert harm to plant facility, secure nature's space by foreseeing releases and minimizing waste, and turn away mischief to the philosophy gear.
2. Maintain cost and quality.
3. Maintain plant generation rate at cheap rate.

There are two systems for controlling any methodology.

1. manual control
2. automatic control

Not with remaining, there are a couple of issues with such manual control. In any case, the occupation obliges that the pro look at the temperature frequently to make therapeutic move at whatever point it strays from the needed quality. Second, assorted directors settle on different decisions about how to move the steam valve, and these impacts in a short of what perfectly foreseeable operation. Third, in light of the fact that in most process plants there are a few variables that must be kept up at some ached for quality, manual correction obliges a broad number of overseers. As an eventual outcome of these issues, we may need to accomplish this control automatically. That is, we may need to have strategies that control the variables without any manual help or regular watch. This is what we mean automatic process control.

A PID controller is a control loop feedback mechanism used in most of the industrial control systems. A PID controller evaluates the error as the difference between a measured process variable and a desired set point. The controller reduces the error, overshoots and increases the response.

The PID controller algorithm includes the 3 constant parameters the proportional (P), the integral (I) and derivative (D) values. *P* depends on the *present* error, *I* on the accumulation of *past* errors, and *D* is a prediction of *future* errors, based on current rate of change. These 3 actions are used together to reduce the error via a control element such as the position of a control valve, a damper, or the power supplied. There are several methods for tuning a PID loop. The most effective methods generally involve the development of some form of process model by choosing P, I and D values. If the system can be taken offline, the best tuning method often involves subjecting the system to a step change in input, measuring the output as a function of time, and using this response to determine the control parameters.

The objective of automatic process control structure is to change the controlled variable to keep up the controlled variable at its set point slighting all aggravations.

1.2 PID Controller tuning methods

There are different tuning methods which are used to calculate the tuning parameters. Different methods are used for different processes depending upon the process and the desired output.

I) Ziegler-Nichols method

II) Cohen-Coon method

III) IMC METHOD

IV) Tyreus-Luyben method

V) Minimum error criteria (IAE, ISE, ITAE) method

Ziegler-Nichols method:

In this method Ki and Kd are first set to zero and then Kp is increased slowly from zero to the value where we get the ultimate gain and at this value the output loop starts oscillating, then we calculate the value of the Pu from the response.

Table I : Ziegler-Nichols Tuning Parameters

Controller type	Kc	Ki	Kd
P	0.5Ku		
PI	0.45Ku	1.2Kp/Pu	
PID	0.6Ku	2Kp/Pu	KpPu/8

Cohen-Coon method:

Control action in this method is removed and an open loop transient is introduced. The step response is observed at the output of the measuring element and then the controller tuning parameters are then evaluated by using the table described.

Table II : Cohen-Coon Tuning Parameters

Controller	Kc	Ti	Td
P	T/ KpTd(1 + Td/3T)		
PI	T/ KpTd(9/10 + Td/12T)	Td(30+3Td/T)/ (9+20Td/T)	
PID	T/ KpTd(4/3 + Td/4T)	Td(32+6Td/T)/ (13+8Td/T)	4Td/ (11+2Td/T)

Tyreus-Luyben method:

In this method Ki and Kd are first set to zero and then Kp is increased slowly from zero to the value where we get the ultimate gain Ku and at this value the output loop starts oscillating, then we calculate the value of the Pu from the response and finally the parameters are calculated using table (III).

Table III : Tyreus-Luyben Tuning Parameters

Controller	Kc	Ki	Kd
PI	0.3125 Ku	Kp/2.2Pu	
PID	0.4545 Ku	Kp/2.2Pu	KpPu/6.3

Integral model control (IMC) method:

In this method we use the process model to calculate the parameters of IMC based PID structure. It uses the invertible part of the process transfer function and also has a filter tuning factor which depends upon the robustness.

Minimum error criteria (IAE, ISE, ITAE) method:

In this method we design controller by using performance indices which considers entire closed loop because tuning for one-fourth decay ratio leads to oscillatory response

$$IAE = \int |e(t)| dt$$

$$ISE = \int (e(t))^2 dt$$

$$IATE = \int t |e(t)| dt$$

Where, t is the time and e(t) is the error which is calculated as the difference between the set point and the output.

CHAPTER 2:

CONTROL STRATEGIES:

2.1 <u>DIFFERENT TYPES OF CONFIGURATIONS</u>

2.1.1 <u>FEEDBACK CONTROL</u>:

Feedback control system is most commonly used process in modern industries and is cheap also. In feed back control system the disturbance is allowed into the feedback loop and is manupulated several times to get the desired output. The process is automatic however it does need any information about the process and disturbance.

A feedback control move makes the accompanying steps:

1. Measures the worth of the yield utilizing proper measuring device
2. Compares the measured worth with the set point of the yield and finds the deviation.
3. The quality of the deviation is supplied to the fundamental controller. The controller in turns changes the worth of the manipulated esteem in a manner so is to lessen the greatness of the deviation.

Advantages of Feedback control system:

1. Corrective action are taken when the variables are deviated from the set point.
2. Feedback control requires insignificant information about the process to be controlled; it specific, a scientific model of the process is not needed, despite the fact that it could be exceptionally valuable for control framework plan.

3. The universal PID controller is both adaptable and strong. In the event that process conditions change, returning the controller generally generates agreeable control

Drawbacks of Feedback control system:

1. No remedial move is made until after a deviation in the controlled variable happens. Along these lines, flawless control, where the controlled variable does not veer off from the set point throughout aggravation or set-point progressions, is hypothetically outlandish.
2. Feedback control does not give prescient control activity to make up for the impacts of known or measurable aggravations.
3. It may not be acceptable for processes with substantial time constants and/or long time delays. In the event that vast and regular aggravations happen, the process may work ceaselessly in a transient state and never achieve the sought unfaltering state.
4. In a few circumstances, the controlled variable can't be measured on-line, and, hence, feedback control is not attainable.

2.1.2 FEED FORWARD CONTROL:

Feed forward control measures the disturbances as they enter the process. The arrangement utilizes a controller to alter manipulated variable with the goal that the influence of the disturbances on the controlled variable is diminished or killed. Feed forward control requires a mindfulness and understanding of the impact that the disturbance will have on the controlled variable. It can compute the precise sum by which the manipulated variables ought to change to compensate for the disturbances. It requires a precise measurement of disturbances.

Advantages of Feed forward system:

1. takes curative movement before the process is upset.
2. theoretically fit for "immaculate control".
3. does not influence system soundness.

Drawbacks of Feed forward system:

1. The disturbance variables must be measured on-line. In numerous requisitions, this is not achievable.
2. To make successful utilization of feed forward control, in any event an estimated process model ought to be accessible. Specifically, we have to know how the controlled variable reacts to changes in both the disturbance and manipulated variables. The nature of feed forward control relies on upon the correctness of the process model.

3. Ideal feed forward controllers that are hypothetically equipped for accomplishing immaculate control may not be physically feasible. Luckily, handy estimates of these perfect controllers regularly give extremely viable control.

2.1.3 **CASCADE CONTROL:**

In cascade control configuration, we have one manipulated variable and all the more then one measurement. It is an elective to think about if immediate feed back control utilizing the primary variable is not palatable and a secondary variable measurement is accessible. It utilizes the yield of primary controller to control the set point of secondary controller. The fundamental standard of cascade control is that if the secondary variable reacts to the disturbance sooner then the primary variables then there is the likelihood to catch and invalidate the impact of the disturbance before it proliferates into the primary variables .

FEATURES OF CASCADE CONTROL:

1. More then one measurment, but one manupulated variable
2. The output of master controller serves as set point for the slave controller.
3. Two feed back loop are nested together.
4. Specially useful in eliminating effect of disturbances that move through the system very slowly.
5. Inner loop has a effect of reducing time lag in outer loop, so that the cascade control responds very quickly and improves dynamic performance.

6. Decreases variation in primary variables.
7. Inhances stability characteristics.
8. Insensetive to modelling errors.

2.2 DIFFERENT TYPES OF CONTROLLERS:

There are three different types of controllers.

1. Proportional Controller(P)
2. Proportional-Integral Controller(PI)
3. Proportional-Integral-Derivative Controller(PID)

2.2.1 Proportional Controller (P):

The proportional controller is the most basic controller which acts as a gain for the process.

The equation that describes its operation is

$$m(t) = m + Kc\ e(t)$$

where, m(t) = controller output,

Kc = controller gain

m = bias value.

When Kc is increased it increases the error or the offset value.

Transfer Function is

G(s) = Kc

Proportional controller is nothing but a gain which increases the output of the response in sluggish way.it reduces the maximum overshoot.

2.2.2 Proportional-Integral Controller(PI):

Most processes have offset value which is difficult to control; i.e, they are to be controlled at the set point. Due to this reason we need an proportional controller which removes offset.

The describing equation is

$$m(t) = m + K_c e(t) + \frac{K_c}{Ʈi} \int e(t)dt$$

Where Ʈi*, = integral (or reset) time*

Transfer function is

$$G(s) = K_C \left(1 + \frac{1}{Ʈis}\right)$$

To summarize the PI controller it reduces the steady state error.

2.2.3 <u>Proportional-Integral-Derivative Controller (PID):</u>

A new controller is merged with PI controller known as derivative controller, Which is also known for pre act. Its purpose is to anticipate where the process is heading by looking at the time rate of change of the error, its derivative.

The describing equation is

$$m(t) = m + K_c e(t) + \frac{K_c}{\tau i} \int e(t)dt + K_C \tau d \frac{de(t)}{dt}$$

where

K_c = controller gain.

e $= SP - PV$

PID controller has advantage of all the control actions, it reduces the over shoot and steady state error and also increase the response of the system.

CHAPTER 3:

HIGHER ORDER DYNAMICS SYSTEM

Higher-order processes and systems are classified as either non-interacting or interacting. This section presents two examples of non- interacting systems.

3.1 <u>Non-interacting Level Process considering 2 tanks:</u>

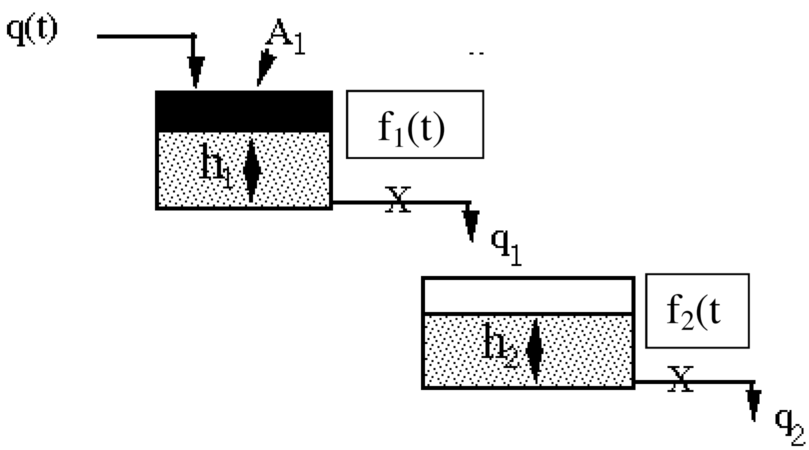


Fig 3.1: 2 level non interacting tanks

In this process both the tanks are open to the atmosphere, and the temperature is steady. The openings of the valves stay consistent, and the flow of liquid through the valves is given by

f(t) = $C_V\sqrt{\Delta P(T)/G_f}$

where

f(t) = flow through valve, m^3/s

C, = valve coefficient, $m^3/s\text{-}Pa^{1/2}$

$\Delta P(t)$ = pressure drop across valve, Pa

G_f = specific gravity of liquid, dimensionless

Because the tanks are open to the atmosphere and the valves discharge to atmospheric pressure, the pressure drop across each valve is given by

$$\Delta P(t) = P_r(t) - P_d = P_a + pgh(t) - P_a = pgh(t)$$

Where

$P_r(t)$ = upstream pressure from valve, Pa

P_d = downstream pressure from valve, Pa

P_a = atmospheric pressure, Pa

p = density of liquid, kg/m3

g = acceleration due to gravity, 9.8 m/s^2

h(t) = liquid level in tank, m

Thus the valve equation for this process becomes

$$f(t) = C_V\sqrt{\Delta P(T)/G_f} \quad = C_V\sqrt{\rho g h(t)/G_f}$$

it is coveted to know how the level in the second tank, h,(t), is influenced by the delta flow into the first tank, h(t), and by the pump flow, f,(t). The destination is to create the numerical model, focus the exchange capacities relating h2(t) to fi(t) and fo(t), and draw the square graph.

Writing an unsteady-state mass balance around the first tank gives

(Rate of mass into the tank) – (Rate of mass out of the tank) = (Rate of accumulation of mass in tank)

$$m_1(t) = \rho A_1 h_1(t)$$

where

A_1 = cross-sectional area of first tank, uniform throughout, m

$h_1(t)$ = liquid level in first tank,

We do not consider the input variables, $f_i(t)$ and $f_o,(t)$, unknowns; it is up to us to specify how they will change. The valve expression provides another equation:

$$f_1(t) = C'v1\sqrt{h1(t)}$$

We now proceed to the second tank. An unsteady-state mass balance around the second tank gives

$$\rho f_1(t) - \rho f_2(t) = \rho A_2 \frac{dh2(t)}{dt}$$

Again, the valve expression provides another equation:

$$f_2(t) = c'_{v2}\sqrt{h2(t)}$$

$$f_1(t) = f_1 + C1[h_1(t) - h_1]$$

We now proceed to obtain the transfer functions. Because Eq2 and 4 are nonlinear, they must first be linearized. This yield

Where

$C_1 = \frac{\delta f1(t)}{\delta h1(t)}$

$C_2 = \frac{\delta f2(t)}{\delta h2(t)}$

Solving these equations and rearranging yield

$$\tau_1 dH_1/dt + H_1(t) = K_1 F_i(t) - K1 F_v(t)$$

$$\tau_2 dh_2/dt + H_2(t) = K_2\ H_1(t) - K1 F_v(t)$$

$\tau 1 = A1/C1$ seconds

$K1 = 1/C1$ ($m/m^3/s$)

$\tau 2 = A2/C2$ seconds

$K2 = C1/C2$

Taking the Laplace transform of these equations and and rearranging, we get

$$H_1(s) = \frac{K1}{\tau 1S+1} Fi(s) - \frac{K1}{\tau 1S+1} Fv(s)$$

$$H_2(s) = \frac{K2}{\tau 2S+1} H1(s)$$

To determine the desired transfer functions, we substitute Eq. 5 into which yields

$$H_2(s) = K_1K_2[F_i(s) - F_v(s)] \ / (\tau_1 s + 1)(\tau_2 s + 1)$$

from which the individual desired transfer functions can be obtained:

$H_2(s) / Fi(s) = K_1K_2 / (\tau_1 s + 1)(\tau_2 s + 1)$

$H_2(s)/F_v(s) = -K_1K_2 / (\tau_1 s + 1)(\tau_2 s + 1)$

At the point when the denominator of these two exchange capacities is ventured into a polynomial structure, the force on the s admin is two. Therefore these exchange capacities are called second request exchange capacities or second-request slacks. Their improvement indicates that they are "shaped" by two first-request exchange capacities, or differential comparisons, in arrangement.

3.2 Non interacting Level Process considering 3 tanks

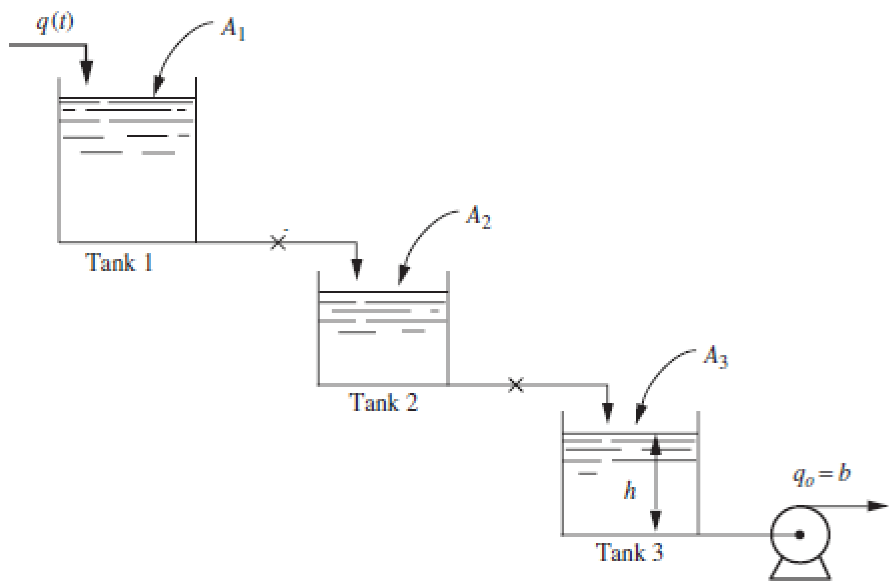


Fig 3.2: 3 level non interacting tanks

$\rho f_2(t) - \rho f_3(t) = \rho A_3 \frac{dh3(t)}{dt}$

The valve expression provides the next required equation:

$f_3(t) = c'_{v3}\sqrt{h3(t)}$

$\tau_3 dH_3/dt + H_3(t) = K_3 h_2(t)$

$C_3 = \frac{\delta f3(t)}{\delta h3(t)}$

$\tau_3 = A_3/C_3$ seconds

$K_3 = C_2/C_3$

Taking the Laplace transform

$H_3(s) = K_3 H_2(s)/\tau 3S+1$

$H_3(s) = K_3 K_1 K_2 [F_i(s) – F_v(s)] / (\tau_1 s +1)(\tau_2 s +1) (\tau_3 S+1)$

$H_3(s) /Fi(s) = K_3 K_1 K_2 / (\tau_1 s +1)(\tau_2 s +1)(\tau_3 S+1)$

$H_2(s)/F_v(s) = -K_3 K_1 K_2 / (\tau_1 s +1)(\tau_2 s +1) (\tau_3 S+1)$

When the denominator of these two transfer functions is expanded into a polynomial form, the power on the s operator is three. Thus they are referred to as third-order transfer functions or third order-lags. shows a block diagram for this process.

SIMULATION FOR 2 NON INTERACTING SYSTEMS:

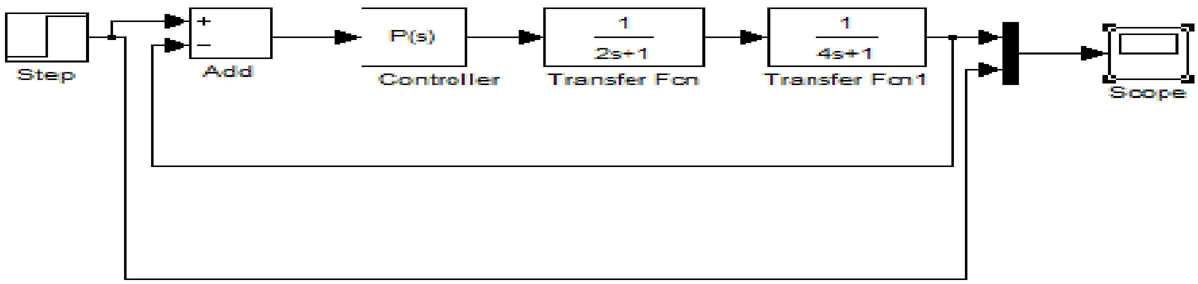


Fig 3.3: block diagram for 2 level non interacting tanks using P controller

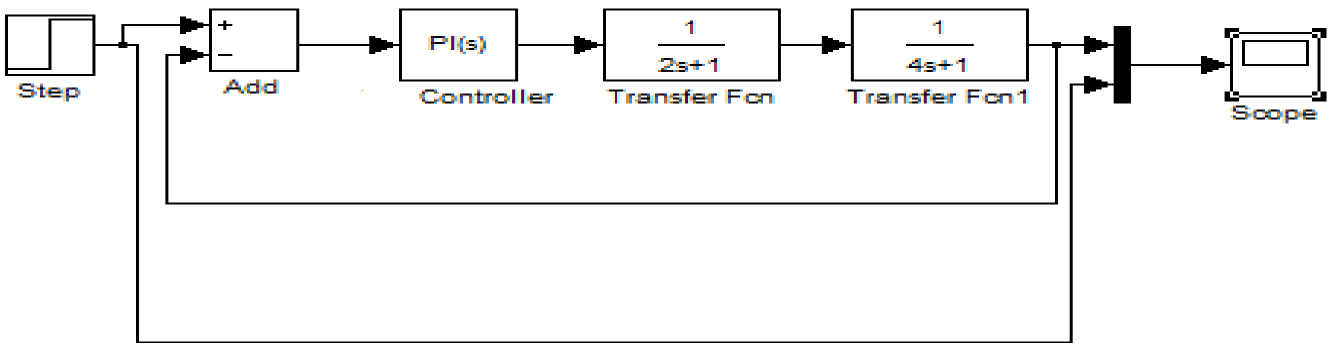


Fig 3.4:block diagram for 2 level non interacting tanks using PI controller

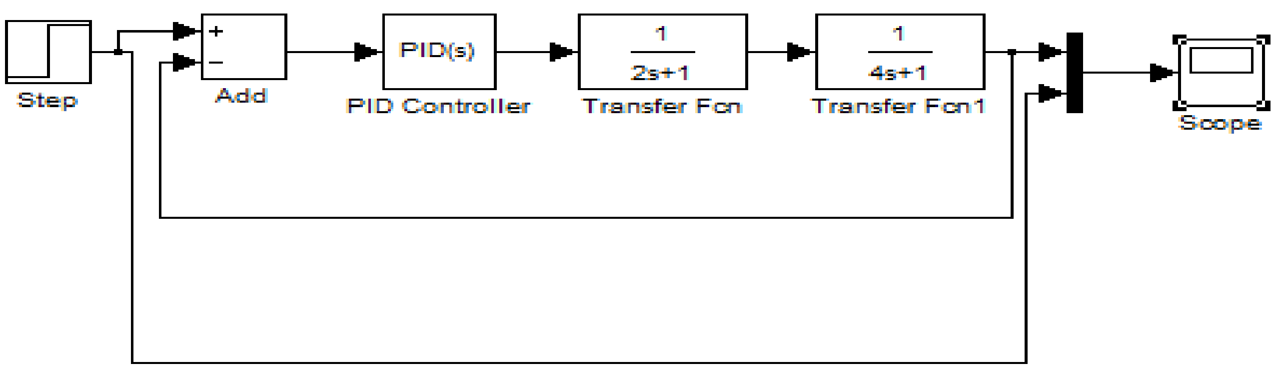


Fig 3.5: block diagram for 2 level non interacting tanks using PID controller

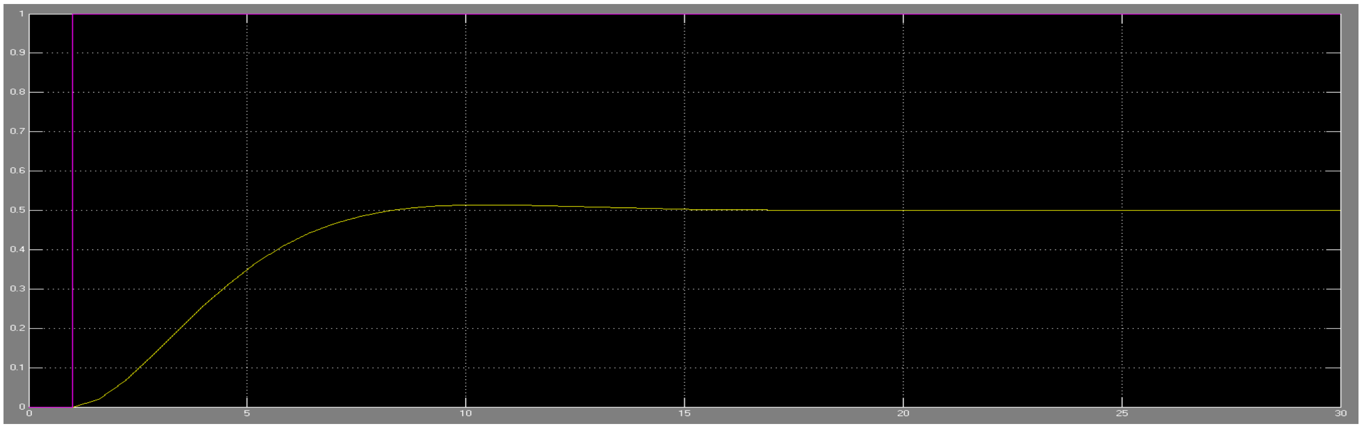


Fig 3.6: simulation for 2 level non interacting tanks using P controller

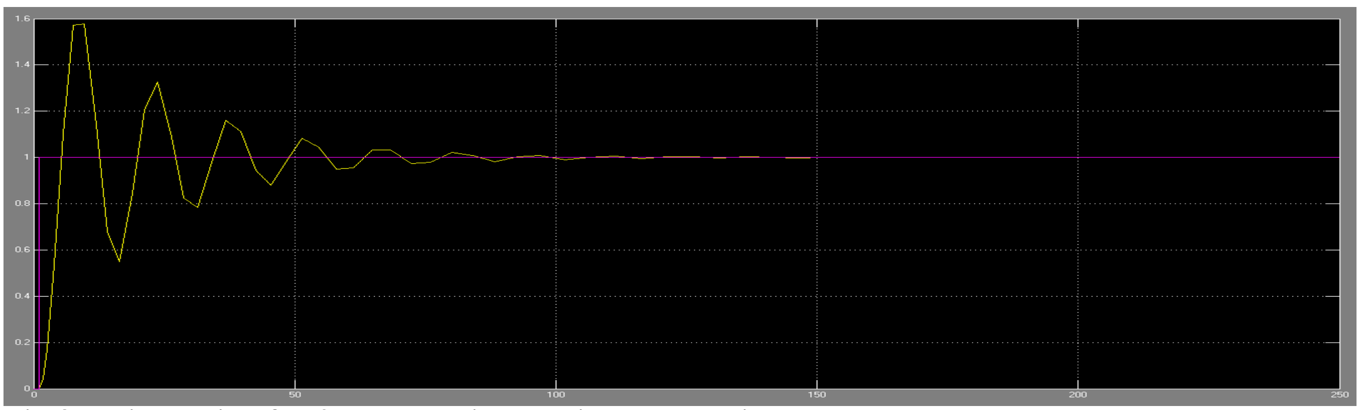


Fig 3.7: simulation for 2 level non interacting tanks using PI controller

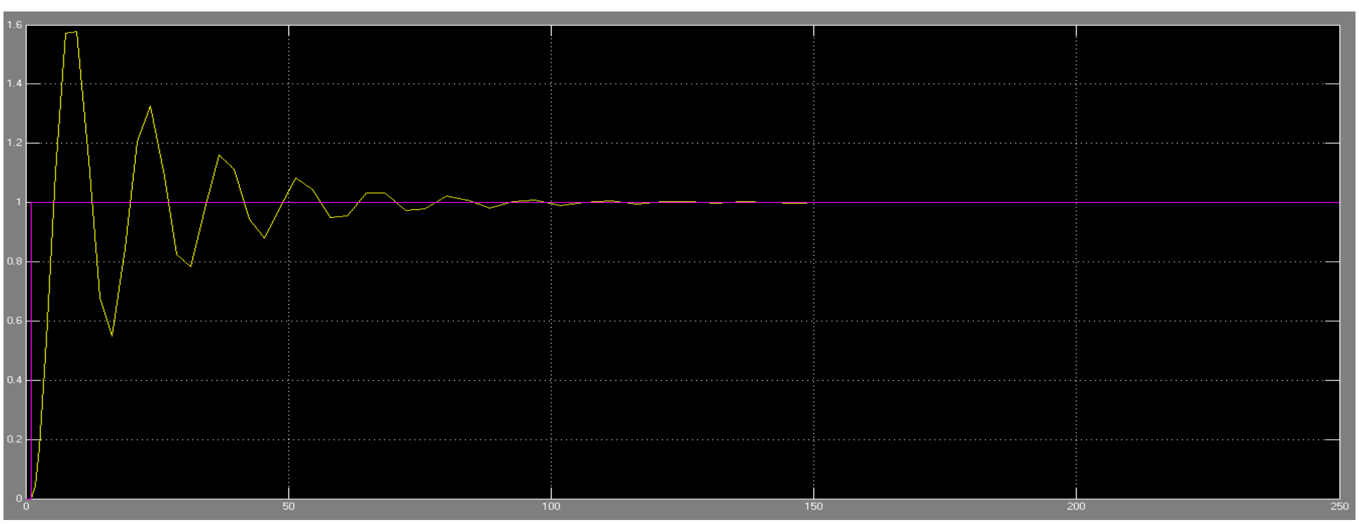


Fig 3.8: simulation for 2 level non interacting tanks using PID controller

SIMULATION FOR 3 NON INTERACTING SYSTEMS:

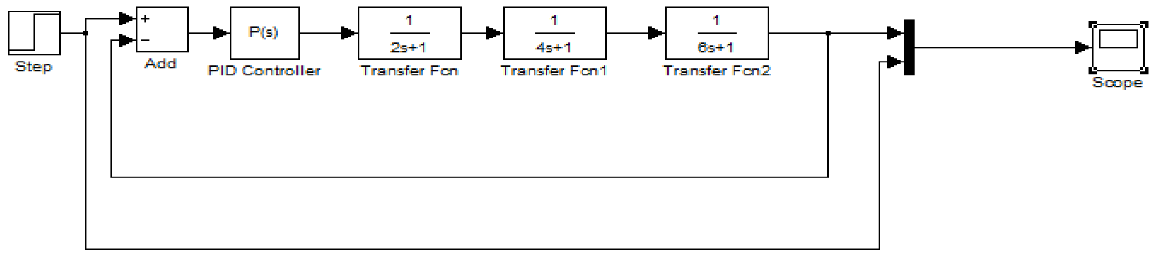


Fig 3.9: block diagram for 3 level non interacting tanks using P controller

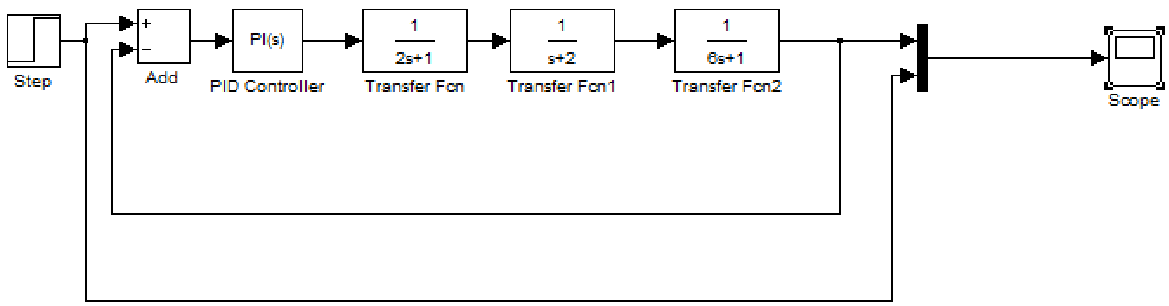


Fig 3.10: block diagram for 3 level non interacting tanks using PI controller

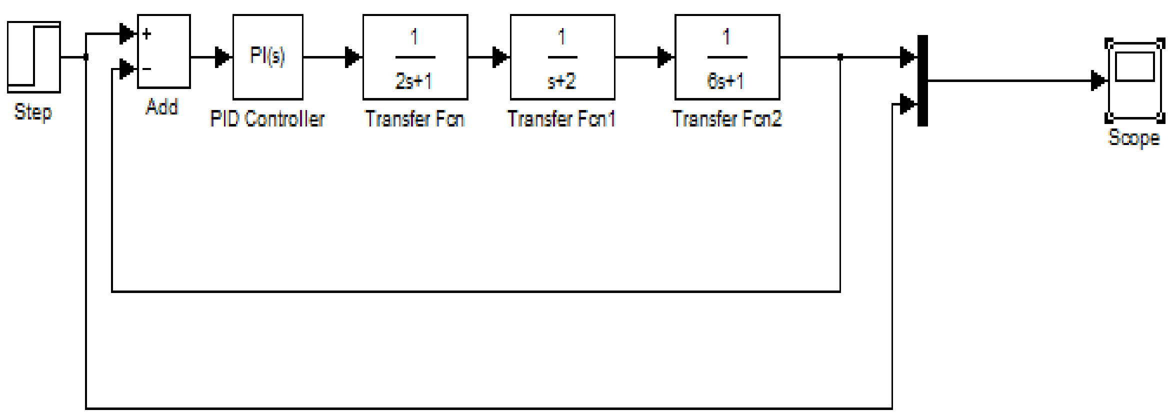


Fig 3.11: block diagram for 3 level non interacting tanks using PID controller

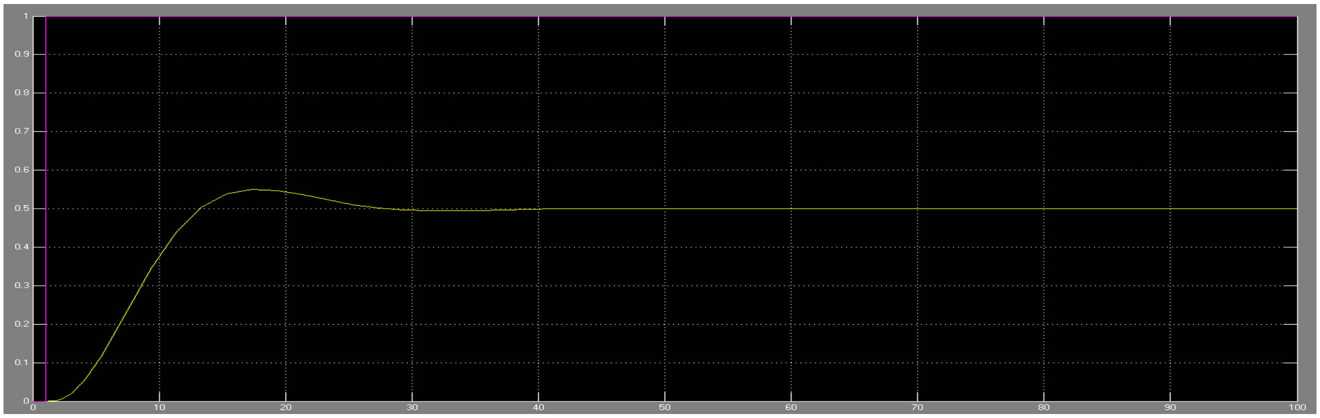


Fig 3.12: simulation for 3 level non interacting tanks using PID controller

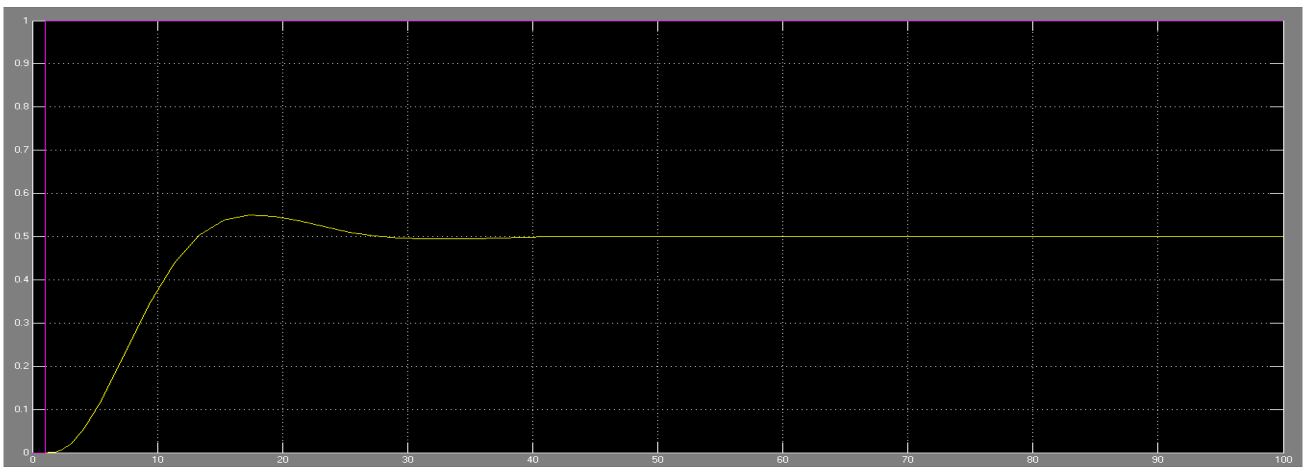


Fig 3.13: simulation for 3 level non interacting tanks using PID controller

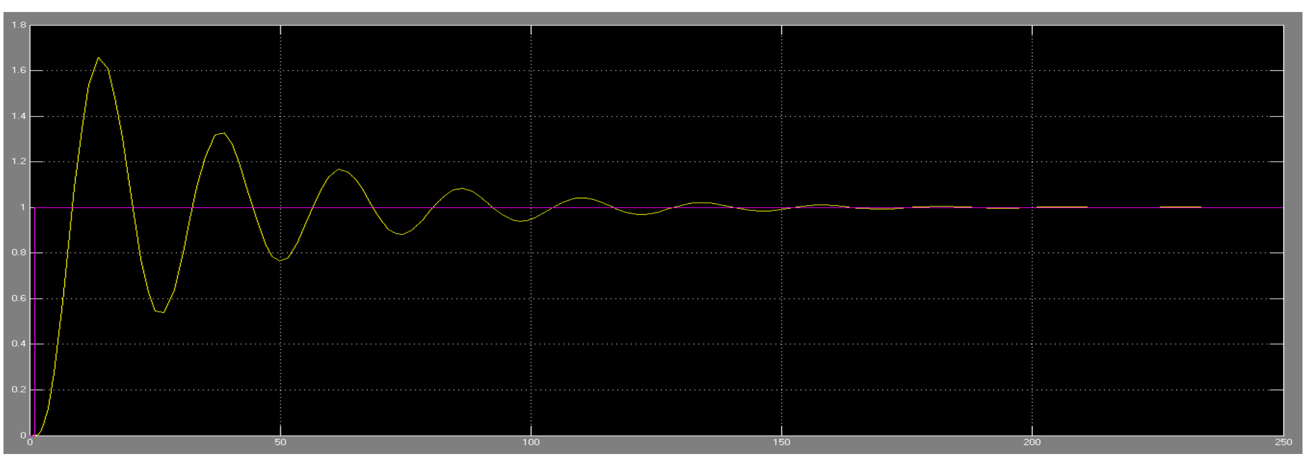


Fig 3.14: simulation for 3 level non interacting tanks using PID controller

3.3 Interacting Level Process

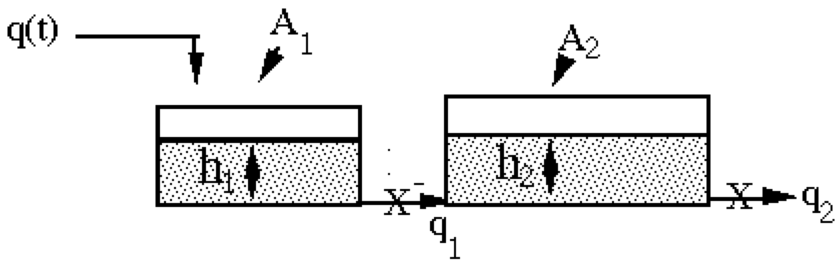


Fig 3.15: 2 level interacting tanks

In this process both the tanks are open to the atmosphere, and the temperature is steady. The openings of the valves stay consistent, and the flow of liquid through the valves is given by

$f(t) = c'_v\sqrt{h1(t) - h2(t)}$

$\rho f_i(t) - \rho f_1(t) - \rho f_0(t) = \rho A_1\frac{dh1(t)}{dt}$

$\rho f_1(t) - \rho f_2(t) = \rho A_2\frac{dh2(t)}{dt}$

Again, the valve expression provides another equation:

$f_2(t) = c'_{v2}\sqrt{h2(t)}$

$f_1(t)= f_1 + C_4[h_1(t) - h_1] - C_4[h_2(t) - h_2]$

$C_4 = \frac{\delta f1(t)}{\delta h1(t)}$

$C_2 = \frac{\delta f2(t)}{\delta h2(t)}$

Solving these equations and rearranging yield

Taking the Laplace transform of these equations and and rearranging, we get

$H_1(s) = \frac{K4}{Ʈ4S+1}[Fi(s) - Fo(s)] - \frac{K1}{Ʈ4S+1}H2(s)$

$H_2(s) = \frac{K5}{Ʈ5S+1}H1(s)$

Where,

$Ʈ_4 = A_1/C_4$ seconds

$K_4 = 1/C_4$

$K_5 = C_4/C_4 + C_2$

$Ʈ_5 = A_2/ C_4 + C_2$ seconds

$H_2/F_i(s) = \frac{K4K5}{1-K5} / \frac{Ʈ4Ʈ5}{1-K5}S^2 + \frac{Ʈ4+Ʈ5}{1-K5}S + 1$

$H_2/F_O(s) = -\{\frac{K4K5}{1-K5} / \frac{Ʈ4Ʈ5}{1-K5}S^2 + \frac{Ʈ4+Ʈ5}{1-K5}S + 1\}$

SIMULATION FOR 2 INTERACTING SYSTEMS:

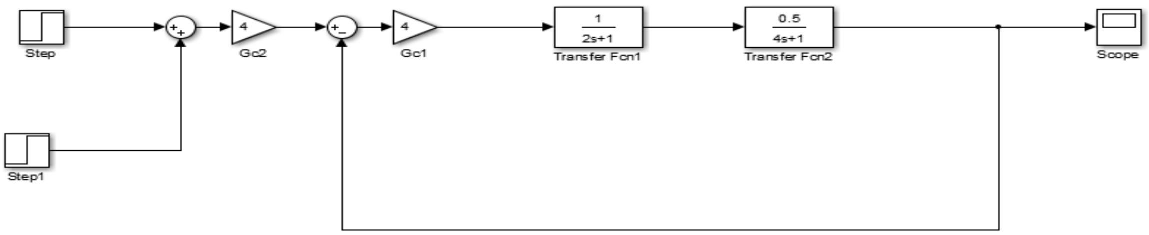


Fig 3.16: block diagram for 3 level non interacting tanks using P controller

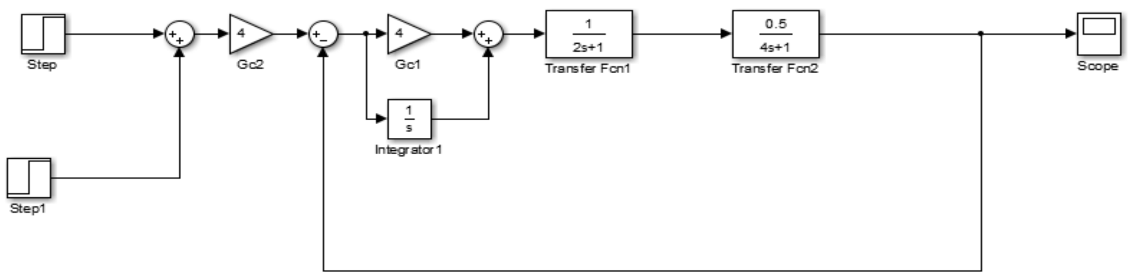


Fig 3.17: block diagram for 3 level non interacting tanks using PI controller

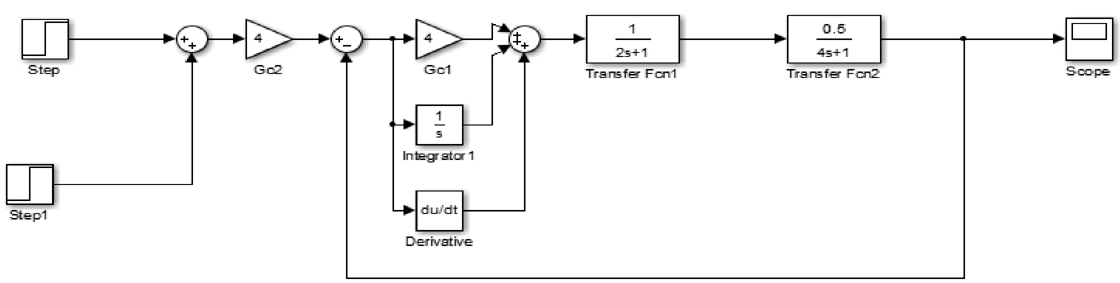


Fig 3.18: block diagram for 3 level non interacting tanks using PID controller

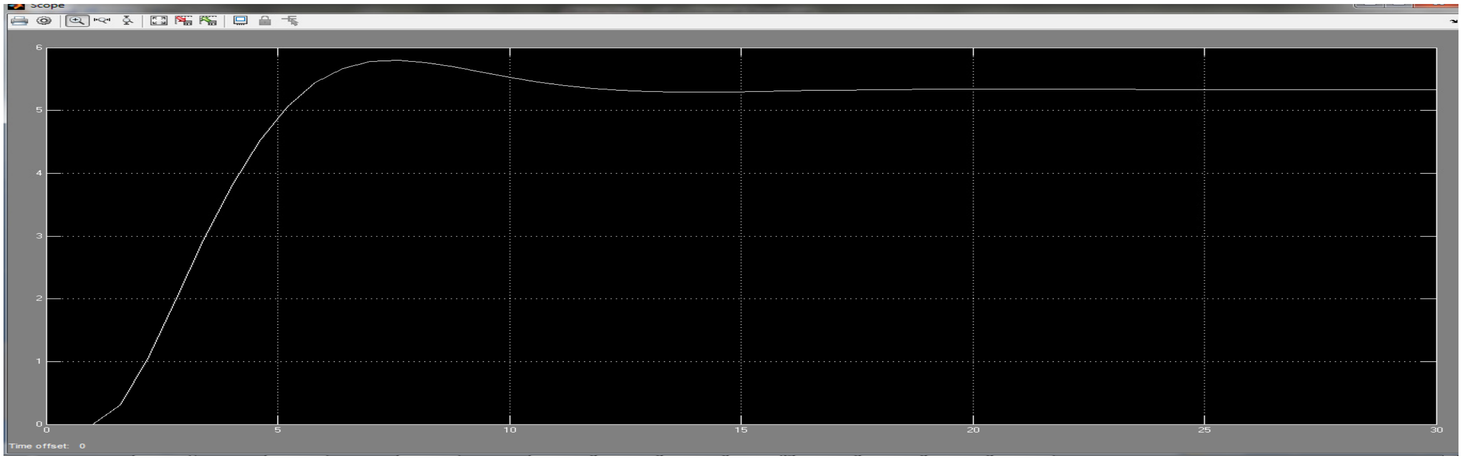


Fig 3.19: simulation for 3 level non interacting tanks using P controller

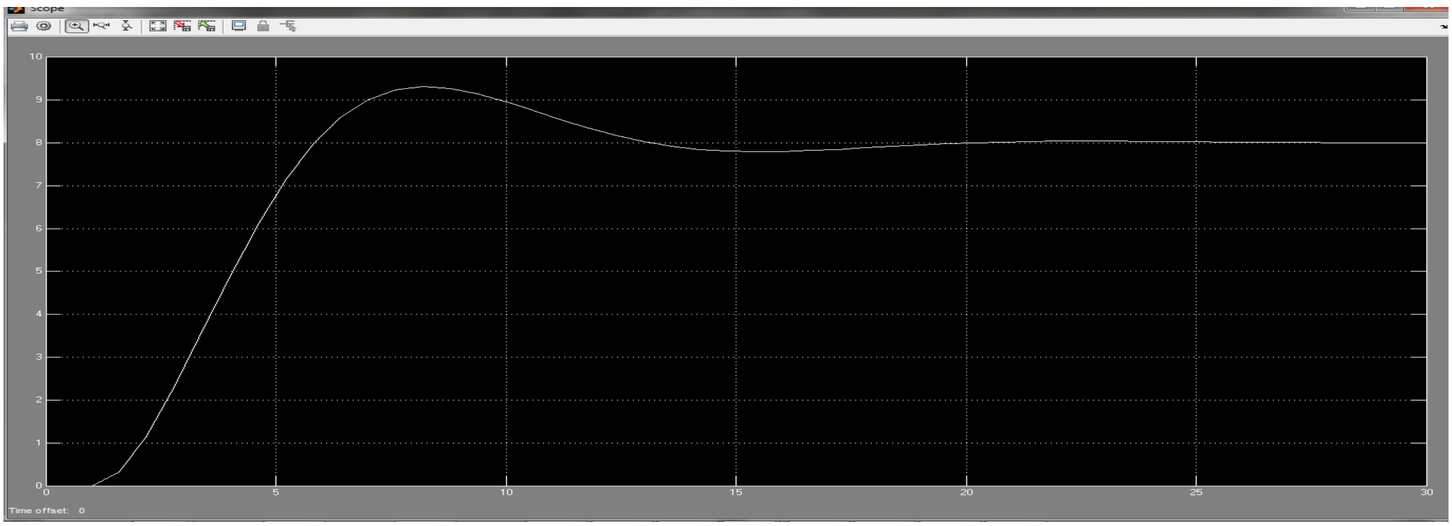


Fig 3.20: simulation for 3 level non interacting tanks using PI controller

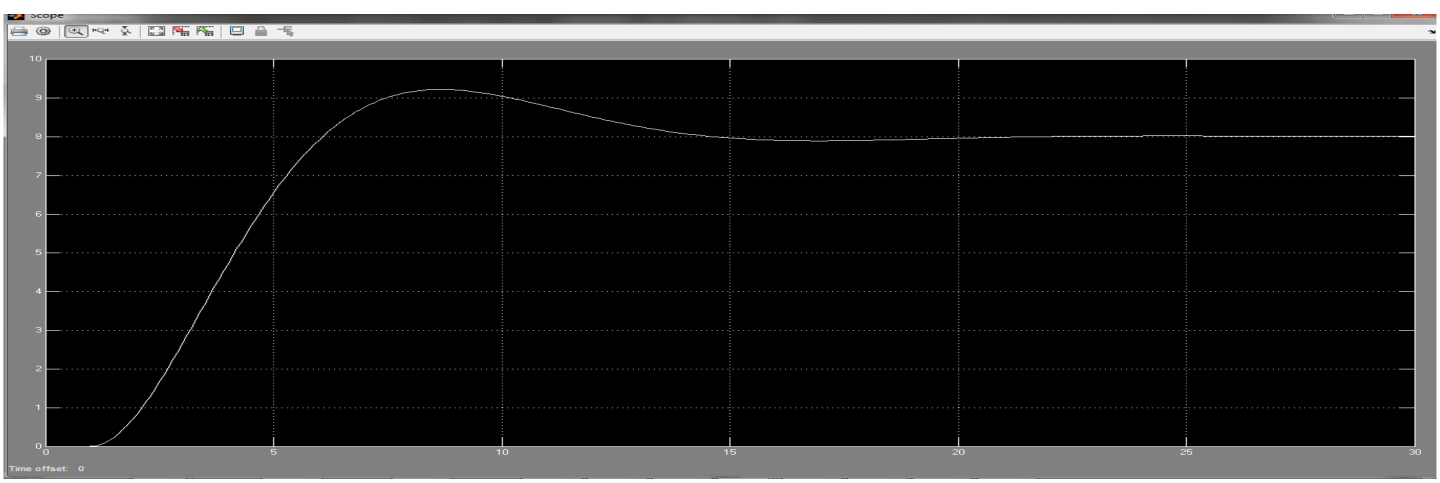


Fig 3.21: simulation for 3 level non interacting tanks using PID controller

CHAPTEER 4:

TUNING BY DIFFERENT METHODS:

4.1 ZIEGLER-NICHOLS METHOD

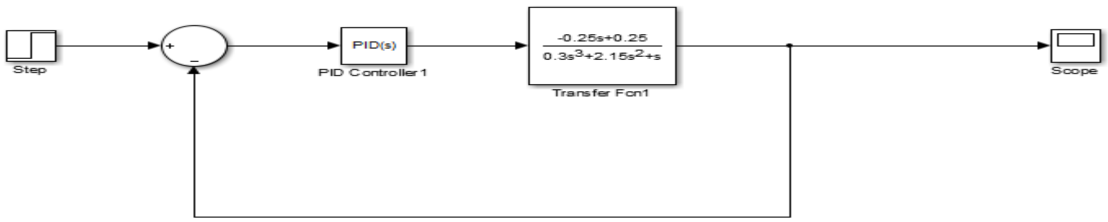


Fig 4.1: block diagram for Ziegler-Nichols tuning method

4.2 TYREUS-LUYBEN METHOD

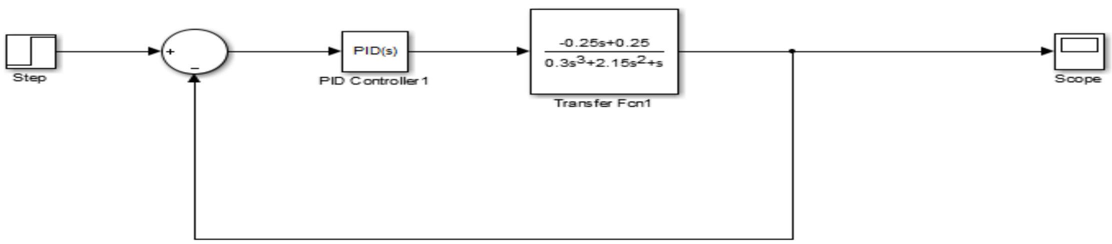


Fig 4.2: block diagram for Tyreus-Luyben tuning method

4.3 COHEN-COON METHOD

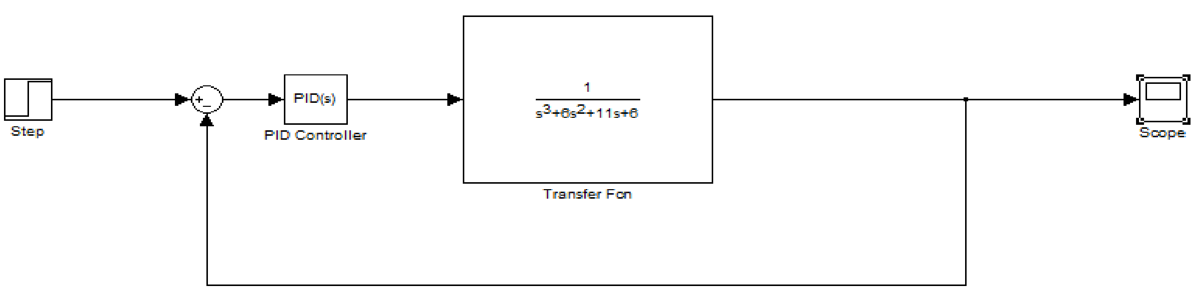


Fig 4.3: block diagram for Cohen-Coon tuning method

4.4 IAE METHOD

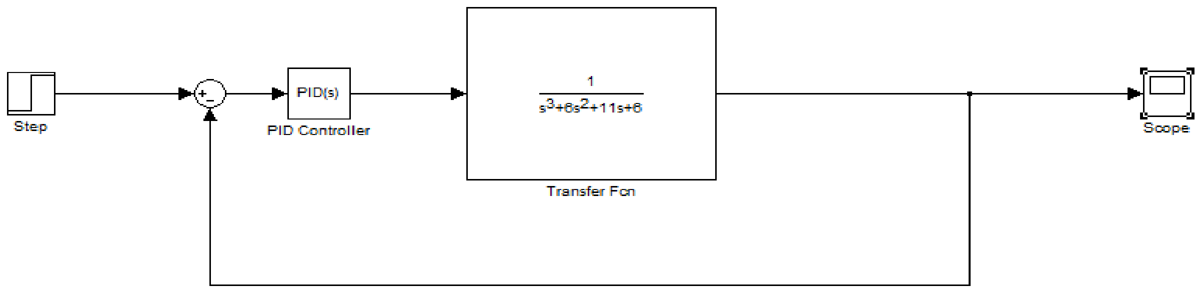


Fig 4.4: block diagram for IAE tuning method

4.5 ISE METHOD

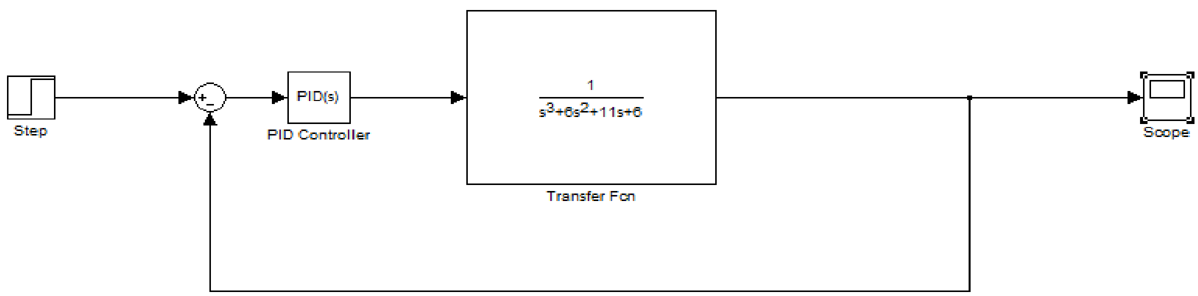


Fig 4.5: block diagram for ISE tuning method

Table IV : OPTIMUM TUNING PARAMETERS VALUES FOR VARIOUS TUNING METHODS

Tuning methods	Proportional gain	Integral gain	Derivative gain
Ziegler-Nichols	6.00	6.33	1.42
Tyreus-Luyben	4.54	1.29	3.00
Cohen-Coon	7.20	7.33	1.09
IAE	12.13	6.41	7.96
ISE	6.84	19.90	19.78

ZIEGLER-NICHOLS

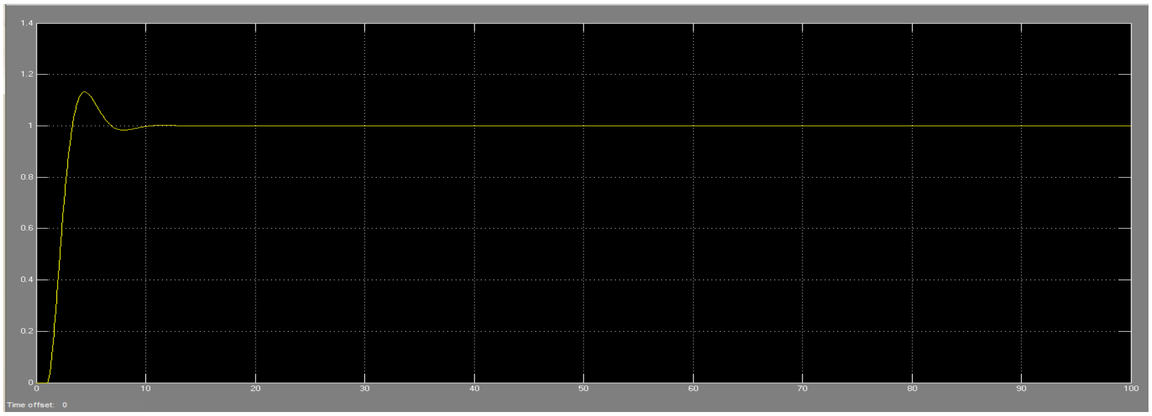


Fig 4.6: Simulation for Ziegler-Nichols tuning method

TYREUS-LUYBEN

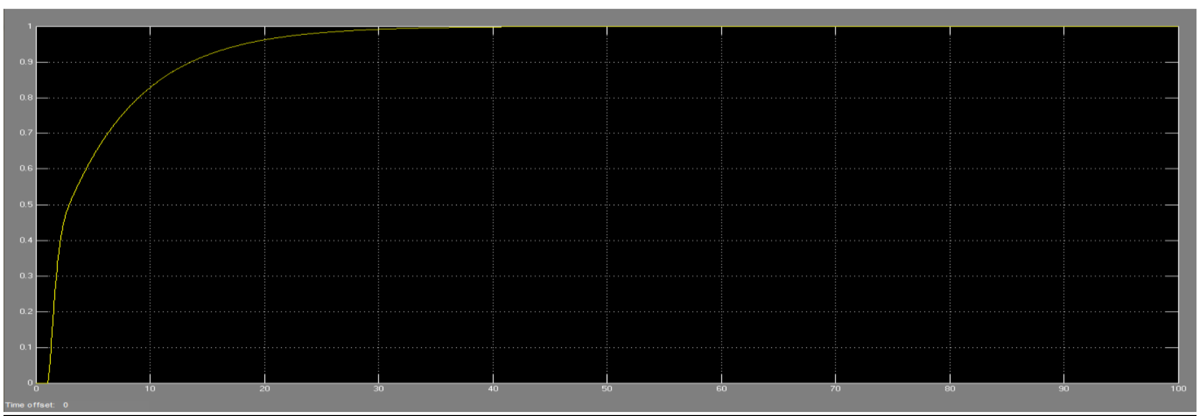


Fig 4.7: Simulation for Tyreus-Luyben tuning method

COHEN-COON METHOD

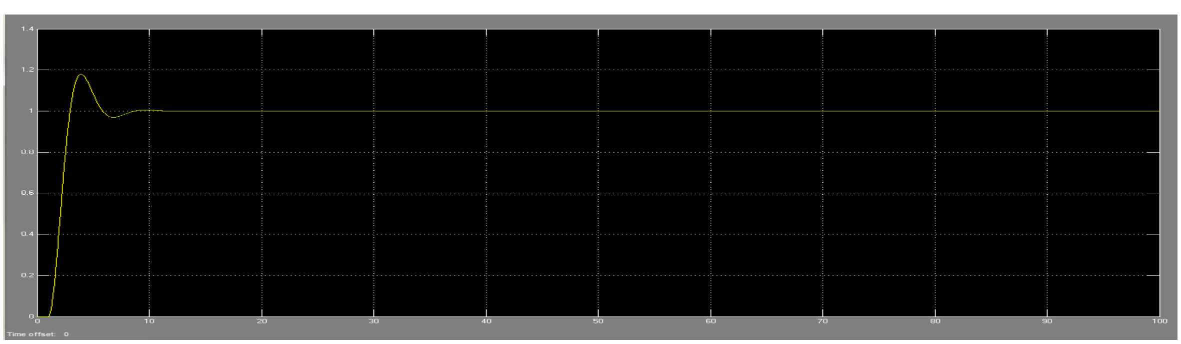


Fig 4.8: Simulation for cohen-coon tuning method

IAE METHOD

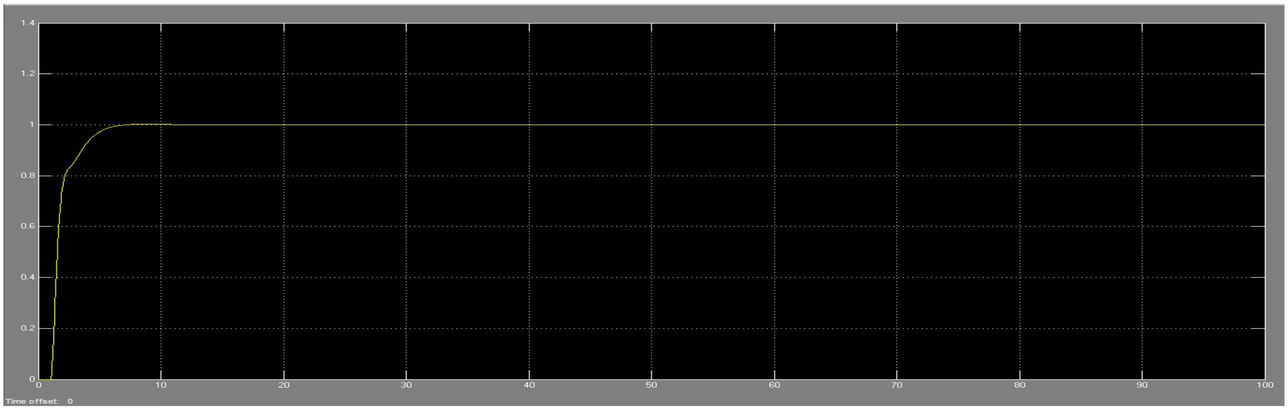


Fig 4.9: Simulation for IAE tuning method

ISE METHOD

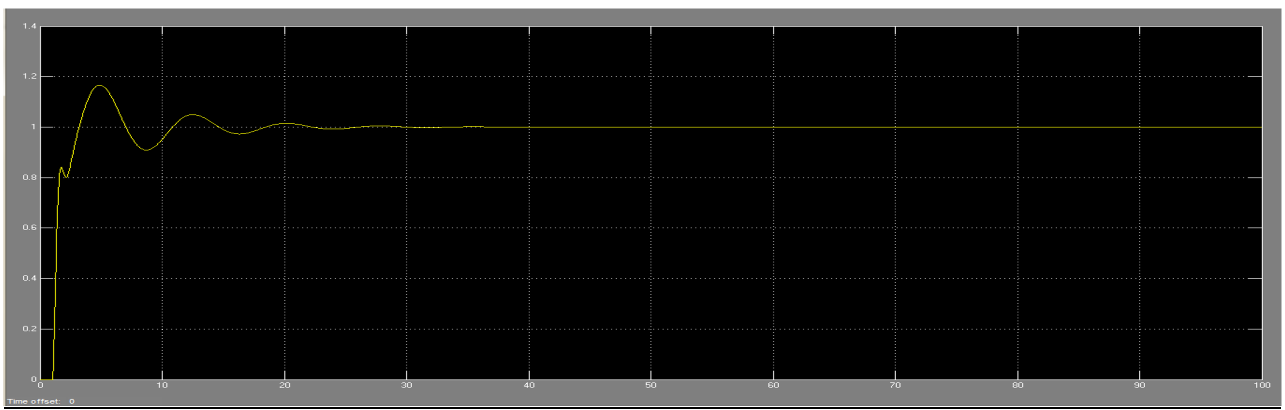


Fig 4.10: Simulation for ISE tuning method

TABLE V : COMPARISON OF PERFORMANCE INDICES OF 3 LEVEL TANK PROCESS USING PID

Time domain specification	Z-N	T-L	C-C	IAE	ISE
Rise Time	0.655	12.50	0.625	2.65	0.18
Settling Time	11.10	33.45	17.50	19.59	20.50
Peak Time	1.13	31.25	1.10	4.28	0.282
Overshoot	1.45	0.00	1.635	1.35	1.43

CHAPTEER 5:

BOILER DRUM LEVEL CONTROL:

Level Control in boiler drum Systems are used in most of the modern industries and in boiler drum we control different parameters such as temperature, level, pressure etc. constant. If the level is high then it discharges the water automatically through a control strategy. If the level is low then it reduces the efficiency of the plants. A control system strategy holds the variables and do not allow them to deviate much from the set point.

Maintaining the level in the drum is tough due to many reasons. If the water level is high then it is discharged. A water level that is too low reduces the efficiency of the plant and it's other function. It results in tube failure when the drum gets overheated due to the lack of cooling water. Normally drum level is 2 to 5cm of the set-point.

During operation bubbles are formed at the top of the surface and it make confusion between true reading and observed value. an increase in water level due to decrease in drum pressure is called swell and decrease in water level due to increase in pressure is called shrink.

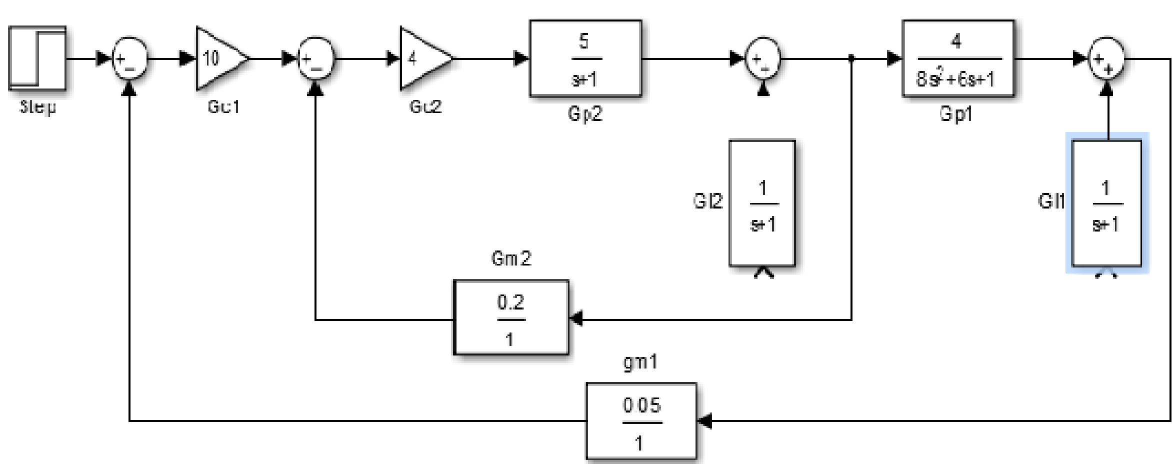


Fig 5.1: Block diagram for boiler drum level control using cascade control strategy

C2/R2 = GC2GP2/ 1+ GC2GP2GM2

$$(C1/R1)CASCADE = \frac{GC1GC2GP1GP2}{1+GP2GM2GC2 + GP1GP2GC1GC2GM1}$$

$$(C1/R1)SIMPLE\ FEEDBACK = \frac{GC1GC2GP1GP2}{1+GP1GP2GC1GC2GM1}$$

$$(e/L2)\ CASCADE = - \frac{GL2GP1GM1}{1+GP2GM2GC2 + GP1GP2GC1GC2GM1}$$

5.1 LEVEL CONTROL SYSTEM USING CASCADE STRATEGY

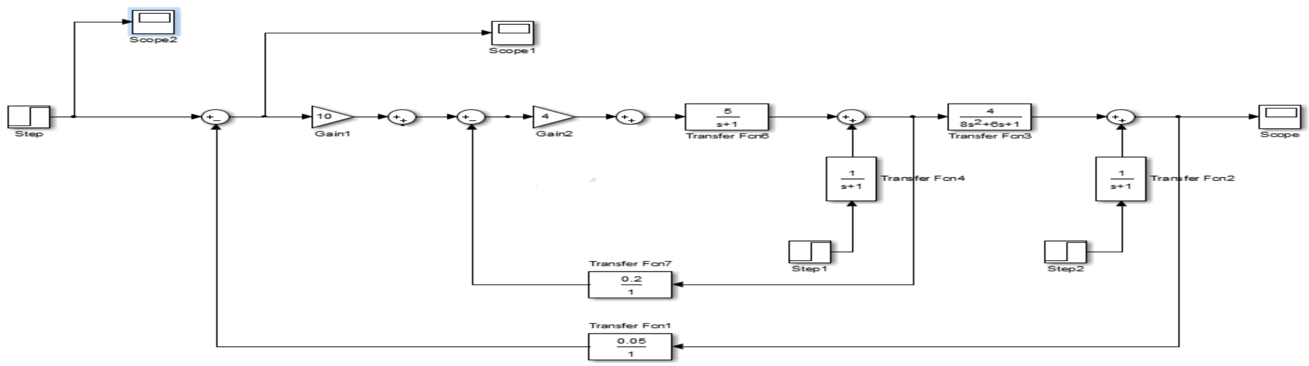


Fig 5.2: Block diagram for boiler drum level control using cascade P-control strategy

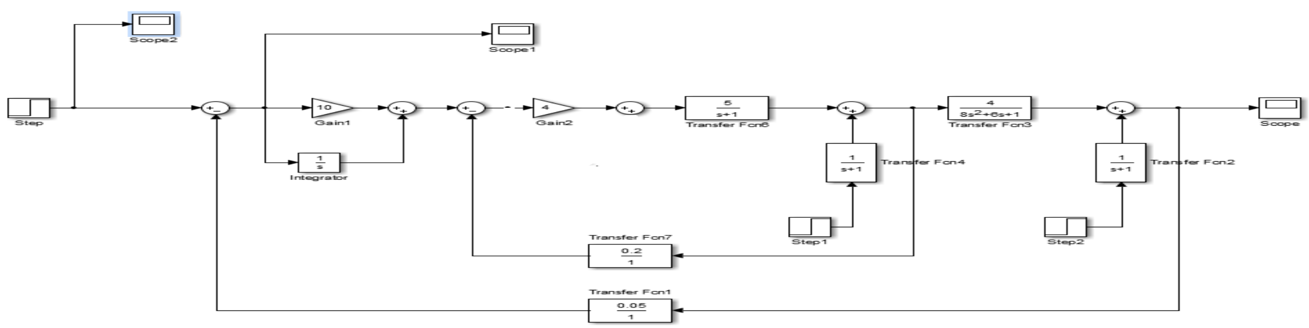


Fig 5.3: Block diagram for boiler drum level control using cascade PI-control strategy

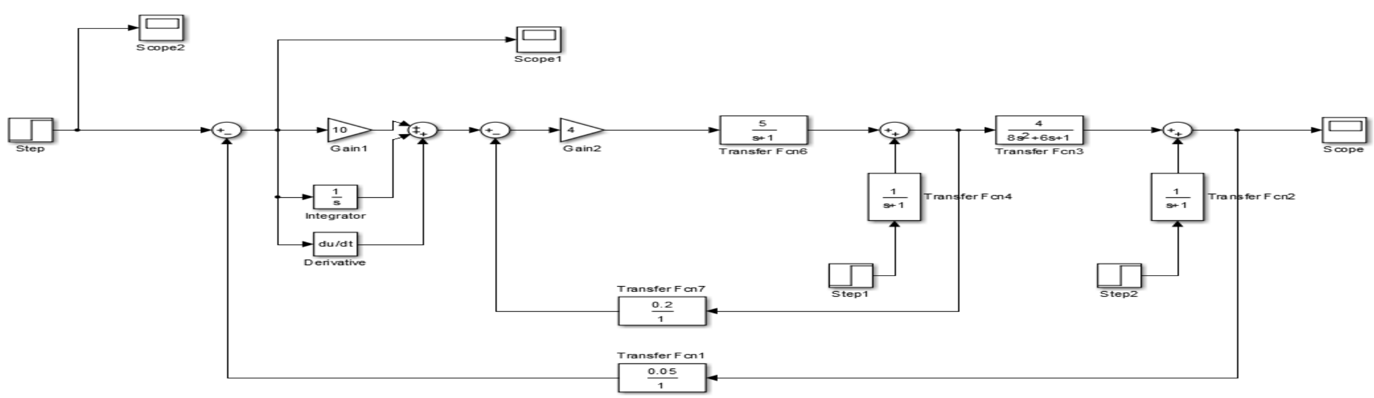


Fig 5.4: Block diagram for boiler drum level control using cascade PID-control strategy

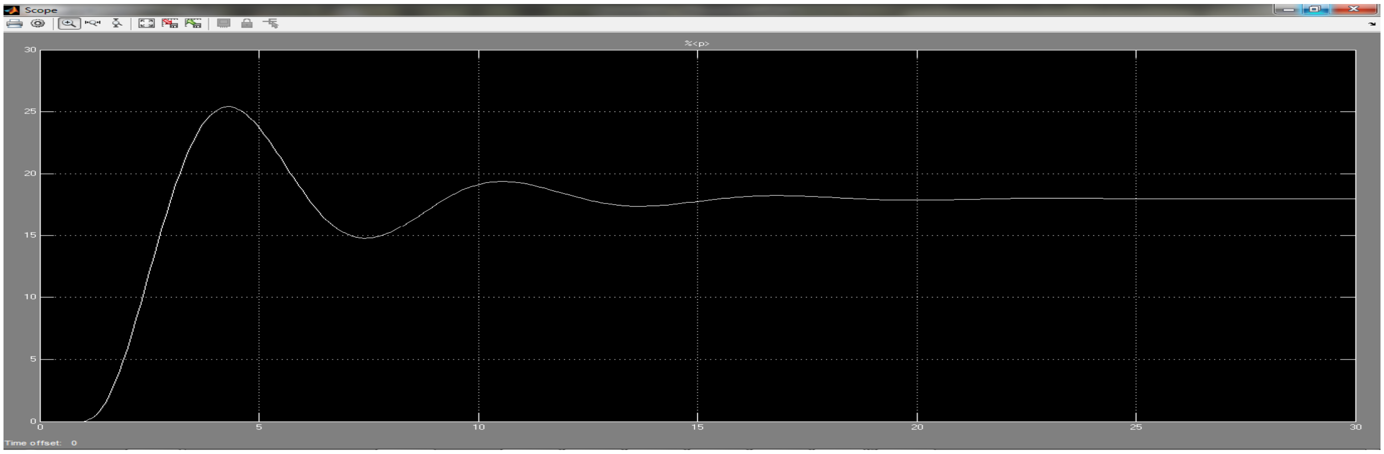


Fig 5.5: Simulation for boiler drum level control using cascade P-control strategy

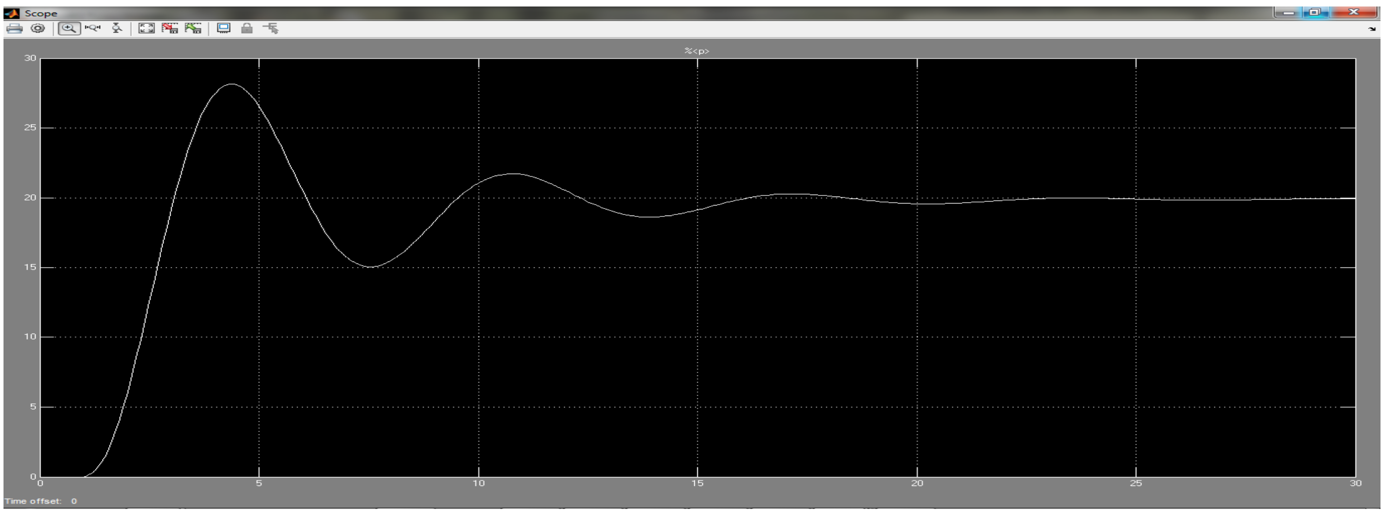


Fig 5.6: Simulation for boiler drum level control using cascade PI-control strategy

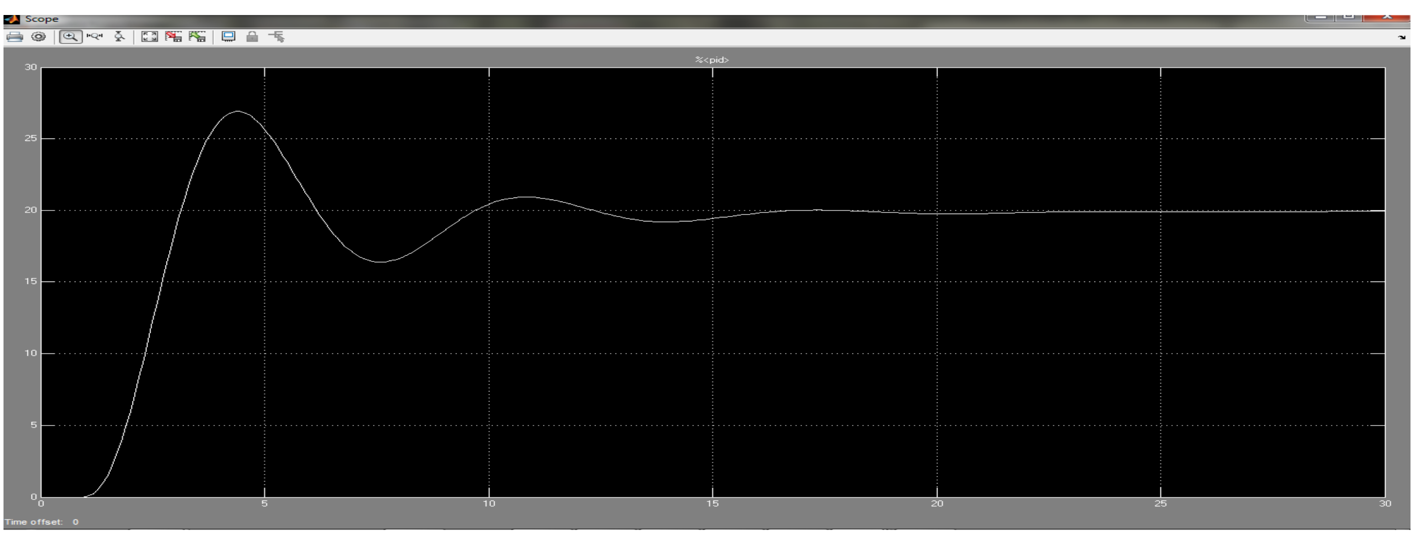
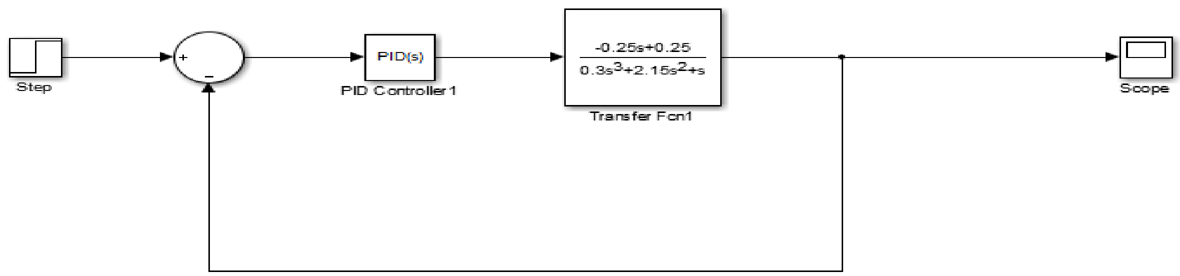


Fig 5.7: Simulation for boiler drum level control using cascade PID-control strategy

5.2 LEVEL CONTROL SYSTEM USING ZIEGLER-NICHOLS METHOD



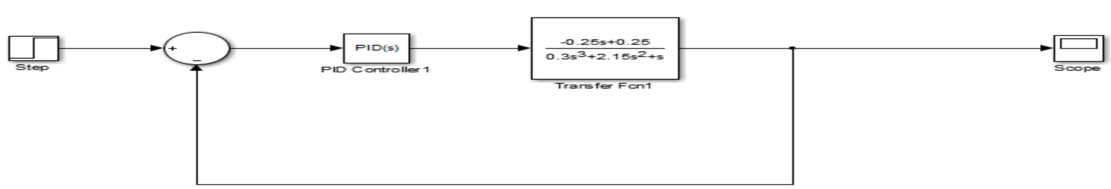
Ziegler-Nichols method

Where Kp = 2.1

Ki = 0.43 & Kd = 2.57

Fig 5.8: Block diagram for boiler drum level control using Ziegler-Nichols tuning method

5.3 LEVEL CONTROL SYSTEM USING ZIEGLER-NICHOLS METHOD



TYREUS- LUBYEN METHOD

Where Kp = 1.59, Ki = 0.073 & Kd = 2.47

Fig 5.9: Block diagram for boiler drum level control using Tyreus-Luyben tuning method

5.4 LEVEL CONTROL SYSTEM USING IMC METHOD

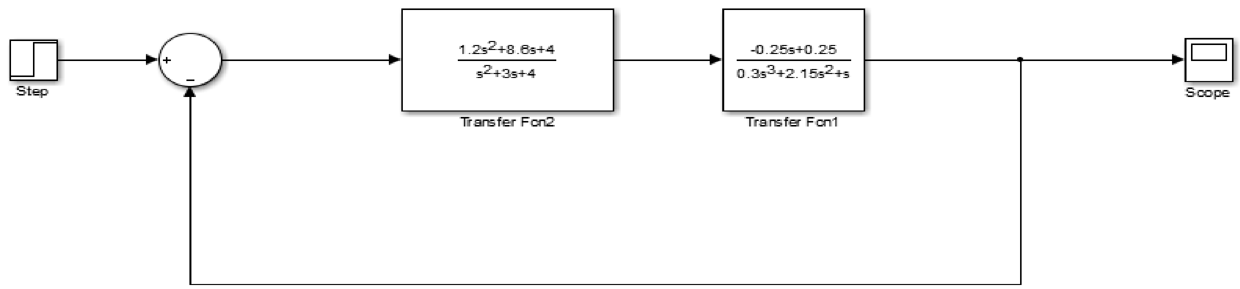


Fig 5.10: Simulation for boiler drum level control using IMC tuning method

5.5 LEVEL CONTROL SYSTEM USING IMCWITH FEEDFORWARD METHOD

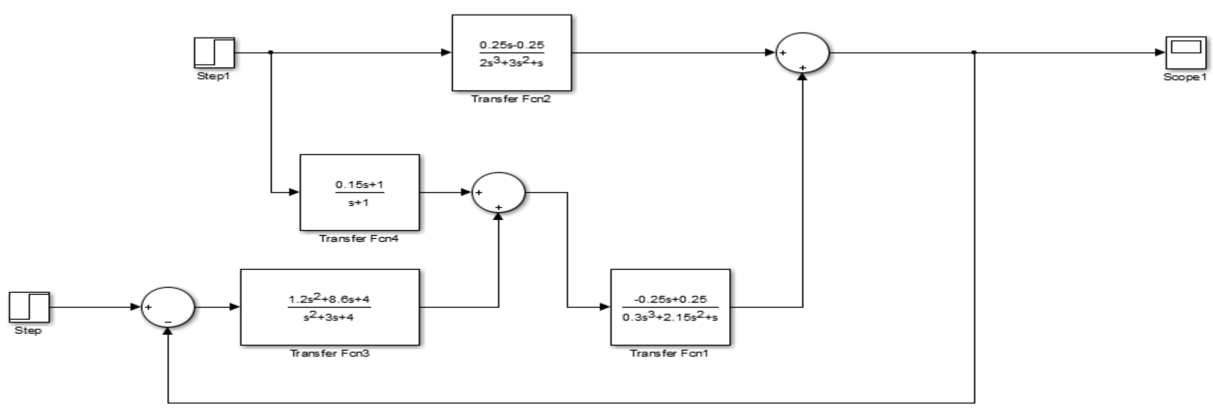


Fig 5.11: Block diagram for boiler drum level control using IMC tuning method with feed forward

TABLE VI : COMPARING OF VARIOUS TIME DOMAIN SPECIFICATION

CONTROLLER	RISE TIME	SETTLING TIME	% Mp
ZLPID	4	30	75
TLPID	4.5	40	20
IMC	4.2	10	0
IMCFF	4.1	9	0

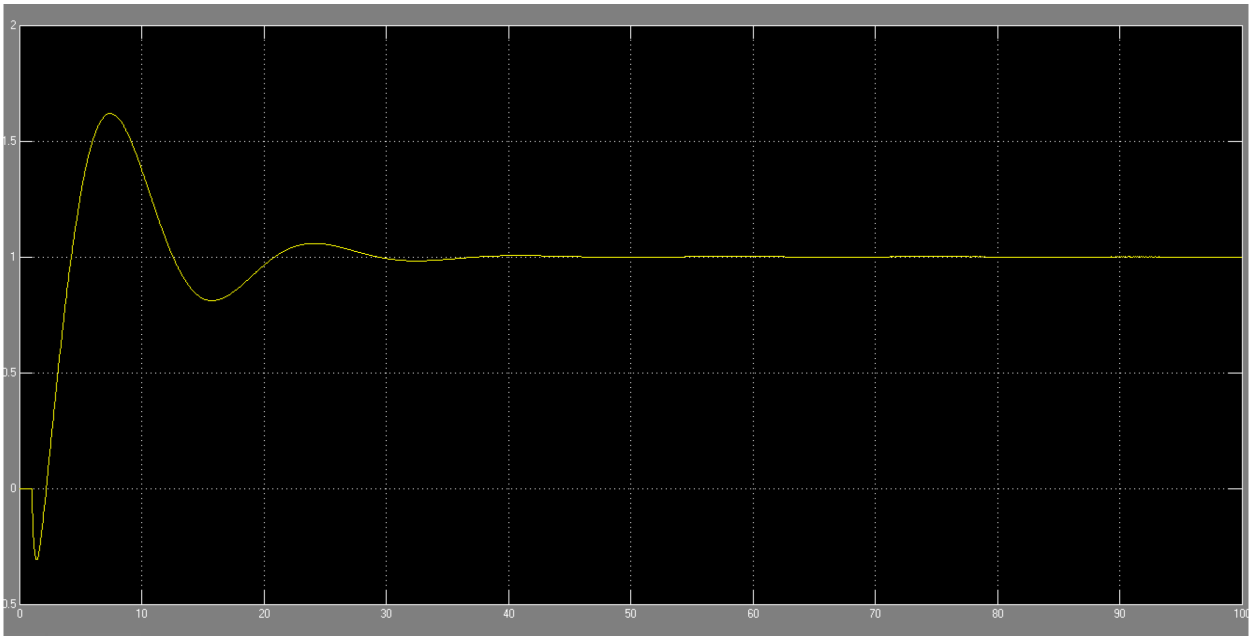


Fig 5.12: Simulation for boiler drum level control using Ziegler-nichols tuning method

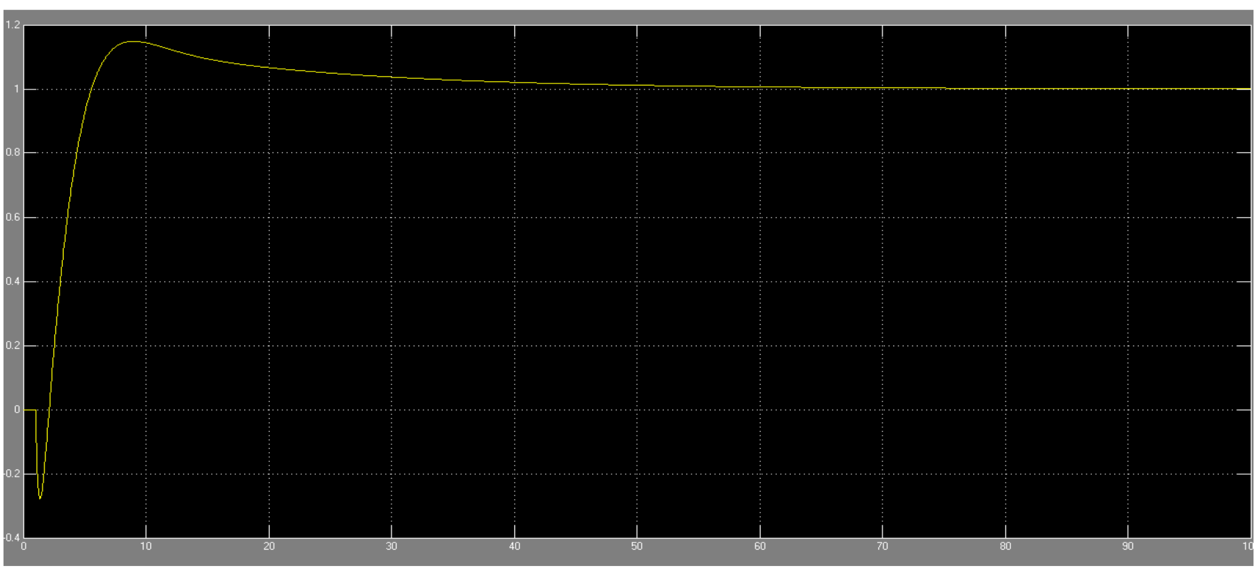


Fig 5.13: Simulation for boiler drum level control using tyreus-luyben tuning method

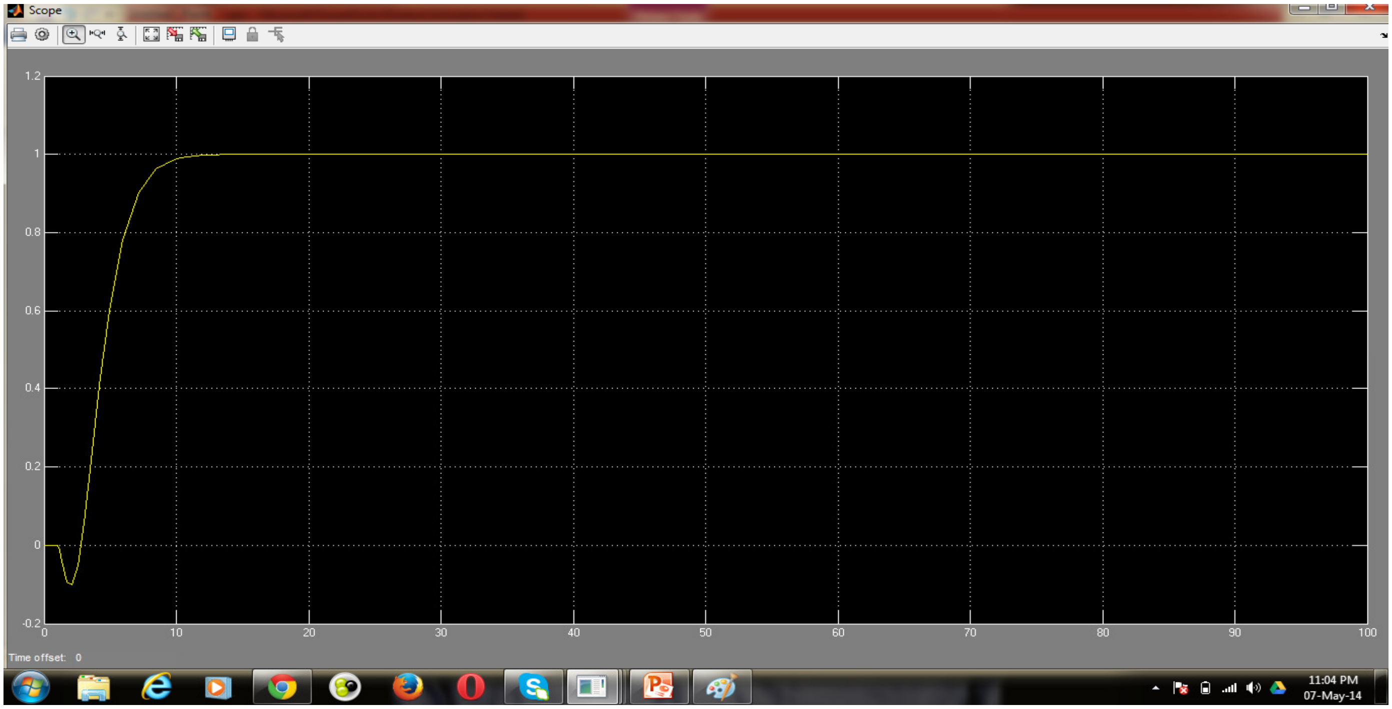


Fig 5.14: Simulation for boiler drum level control using IMC tuning method

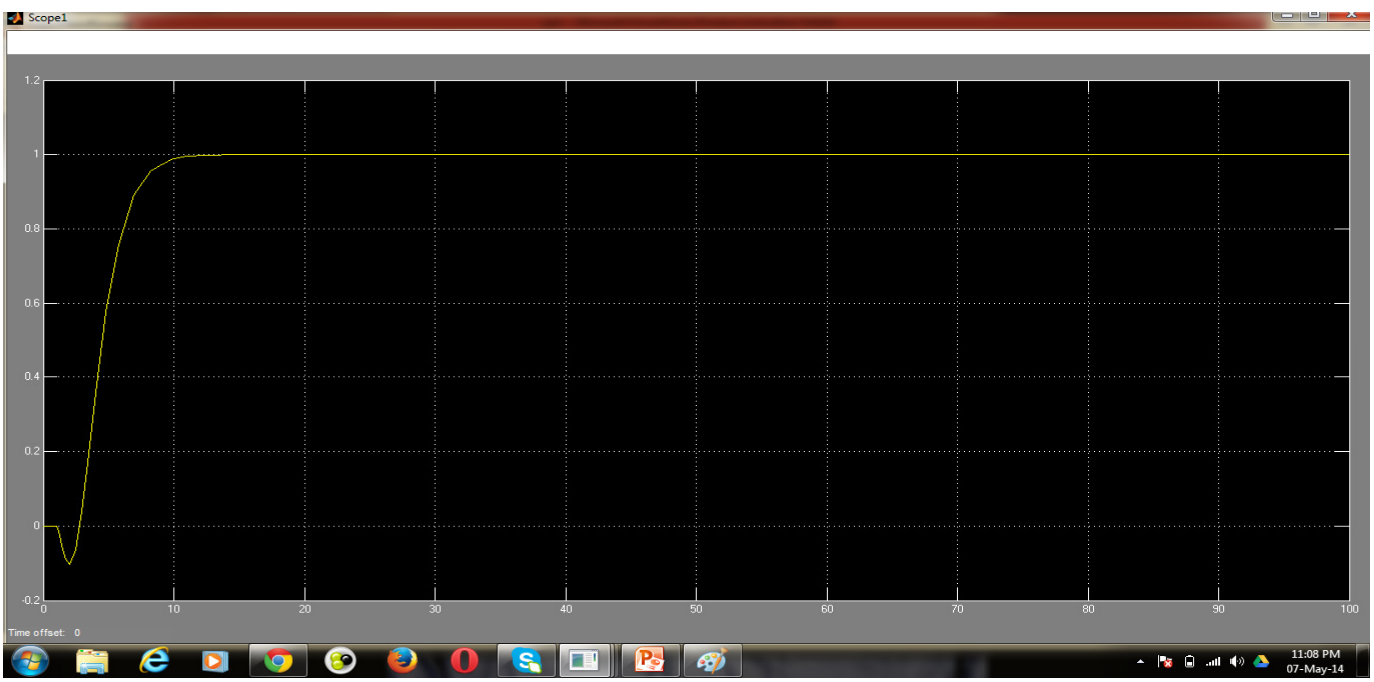


Fig 5.15: Block diagram for boiler drum level control using IMC tuning method with feed forward

CHAPTEER 6:

UNSTABLE CONTINUOUS STIRRED TANK REACTORS

PID controllers give the best performance for many control processes. PID controller has become a powerful and best solution to the control of a large number of industrial processes due to their simplicity and usefulness. Due to the presence of a zero in numerator dynamics, the control systems performance becomes complicated. Many of the processes exhibit second order plus time delay system with a zero transfer function model. Examples for such processes are jacketed CSTR, distillation column, autocatalytic CSTR and crystallizer. Many processes which involves energy and mass recycle are given by SOPTDZ transfer function model.

6.1 <u>IMC METHOD</u> :

Let , $G_p(\mathrm{S}) = \frac{(1+\mathrm{pS})e^{-ls}}{(\tau_1 s+1)((\tau_2 s-1)}$

Now assuming Pade's approximation,

$$e^{-ls} = \frac{1-0.5ls}{1+0.5ls}$$

so, the transfer function becomes,

$$G_p(s) = \frac{K_p(1+ps)(1-0.5ls)}{(\tau_1 s+1)((\tau_2 s-1)(1+0.5ls)}$$

And,

$$G_p(s) = G_{p^+}(s)G_{p^-}(s)$$

$$\mathrm{Q(s)} = [G_{p^-}(s)]^{-1}F(s)$$

Where, F(s) is transfer function of a filter and is given by

$$F(s) = \frac{ns+1}{(\lambda s+1)^3}$$

So, the transfer function for the controller is given as,

$$G_c(s) = \frac{(1+0.5ls)(\tau_1 s+1)((\tau_2 s-1)(ns+1)}{K_p(1+ps)[(\lambda s+1)^3-(1-0.5ls)(ns+1)]}$$

The above equation can be rearranged into the form of,

$$G_c(s) = K_c[1 + \frac{1}{\tau_i s} + \tau_d s]\,\frac{(1+\alpha_0)}{(\alpha_1 s^2+\alpha_2 s+1)}$$

Where,

$$K_c = \frac{\tau_1+n}{K_p(n-3\lambda-0.5l)}$$

$$\tau_i = \tau_1 + n \qquad\qquad \tau_d = \tau_1 \frac{n}{(\tau_1+n)}$$

$$n = \frac{\lambda^3+(3\lambda+0.5L)\tau_2{}^2+3\lambda\tau_2}{(\tau_2{}^2-0.5l\tau_2)}$$

$$\alpha_0 = 0.5l \qquad\qquad \alpha_1 = \frac{\lambda^3 p}{(n-3\lambda-0.5l)\tau_2}$$

$$\alpha_2 = \frac{\lambda^3 p}{(n-3\lambda-0.5l)\tau_2} + p$$

And if we approximate time delay as,

$$e^{-ls} = 1 - ls$$

Then, IMC filter is given by,

$$F(s) = \frac{ns+1}{(\lambda s+1)^2}$$

So, the transfer function of IMC based PID controller is given as,

$$G_c(s) = K_c[1 + \frac{1}{\tau_i s} + \tau_d s] \frac{1}{(\tau_f s+1)}$$

Where,

$$K_c = \frac{\tau_1+n}{K_p(n-2\lambda-l)}$$

$$\tau_i = \tau_1 + n \qquad \tau_d = \tau_1 \frac{n}{(\tau_1+n)} \qquad \tau_f = p$$

$$n = \frac{\lambda^2+(2\lambda+l)\tau_2}{(\tau_2-l)}$$

6.2 STABILITY ANALYSIS METHOD :

In this method, the transfer function of the controller is given by,

$$G_c(s) = K_c'(1+\frac{1}{\tau_i' s})(1+\tau_d' s)(\frac{1}{1+\alpha s})$$

Case study 1 :

Let us assume a jacketed CSTR system consists of first order irreversible exothermic reaction. We use a coolant to remove the heat of reaction in the jacketed CSTR to maintain the temperature of reaction.

The process transfer function which gives the relationship between the reactor temperature and the jacket temperature along with a measurement delay of 0.0317 hr. is given by,

$$\frac{T(s)}{T_j(s)} = \frac{0.82055s + 6.5565}{0.9416s^2 + 2.6977s - 1}\, e^{-0.0317s}$$

Case study 2 :

Let us assume a jacketed CSTR with a constant volume carrying first order irreversible exothermic reaction.

The transfer function relating the reactor temperature to the jacket temperature along with a measurement time delay of 0.06 min is given by,

$$\frac{T(s)}{T_j(s)} = \frac{2.092s + 2.198}{s^2 - 0.4939s - 1.6}\, e^{-0.06s}$$

Case study 3 :

Consider a jacketed CSTR carrying first order irreversible exothermic reaction.

The following transfer function model is considered,

$$\frac{T(s)}{T_j(s)} = \frac{0.3119s + 2.07}{-2.85s^2 - 2.31s - 1}\, e^{-0.3s}$$

The IMC-PID controller and SA-PID controller parameters for different case studies are given in Table VI.

<u>Table VII : PID settings for different methods</u>

Case Study	**Controller**	K_c	τ_i	τ_d	α_1	α_2	α_3
1.	IMC-PID	21.6464	0.482	0.1098	0.0159	0.0015	0.1374
	SA-PID	49.4887	0.4288	0.084			0.12515
2.	IMC-PID	31.1037	1.1979	0.1908	0.03	0.0113	0.9636
	SA-PID	28.036	1.1891	0.1857			0.9518
3.	IMC-PID	-10.7689	1.7175	0.4288	0.15	0.0049	0.1831
	SA-PID	-6.274	2.043	0.5024			0.1507

SIMULATIONS FOR DIFFERENT CASES:

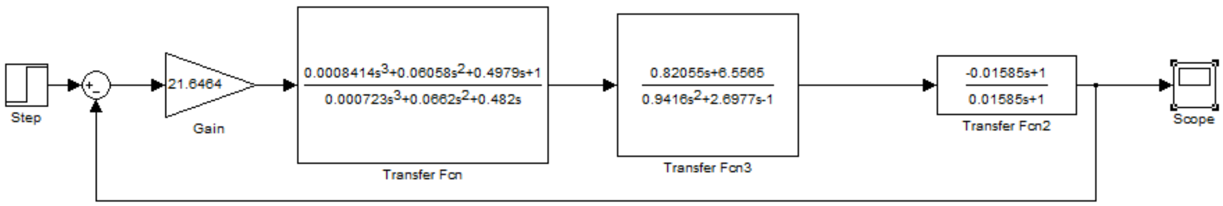


Fig 6.1: IMC based PID controller for case 1

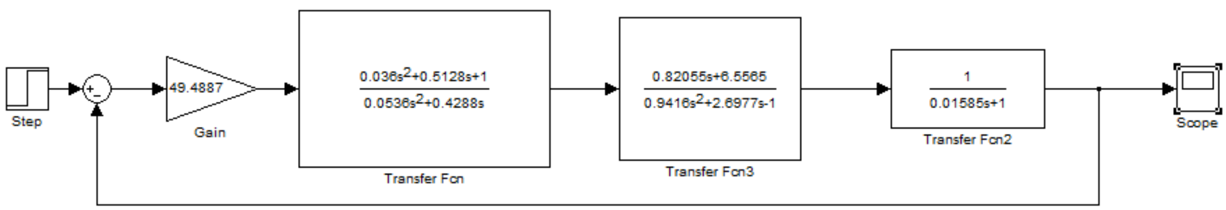


Fig 6.2: SA based PID controller for case 1

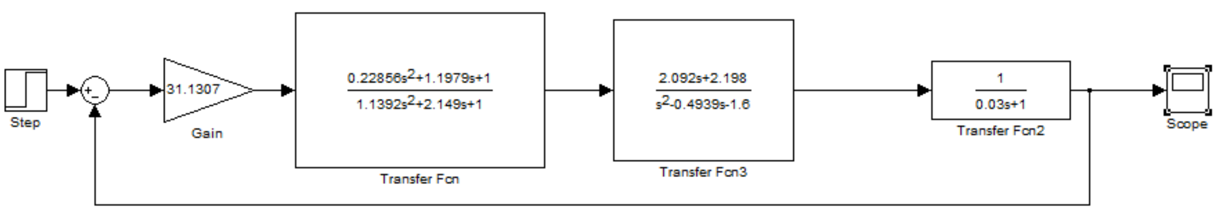


Fig 6.3: IMC based PID controller for case 2

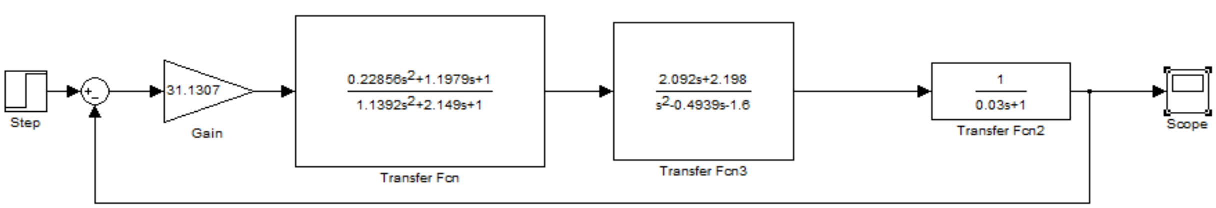


Fig 6.4: SA based PID controller for case 2

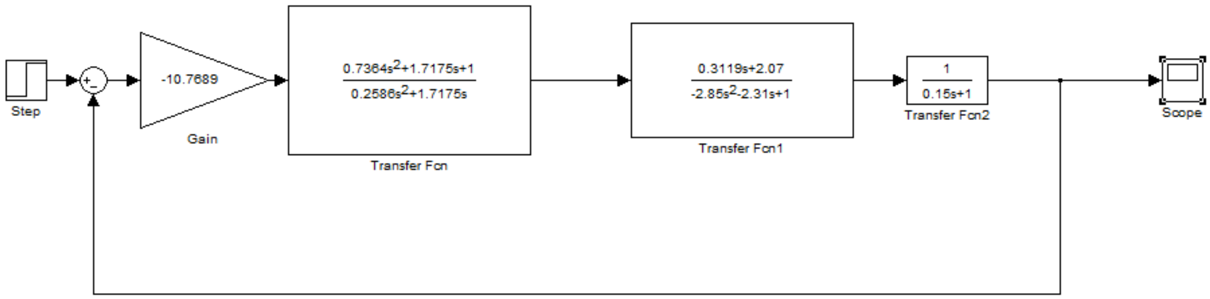


Fig 6.5: IMC based PID controller for case 3

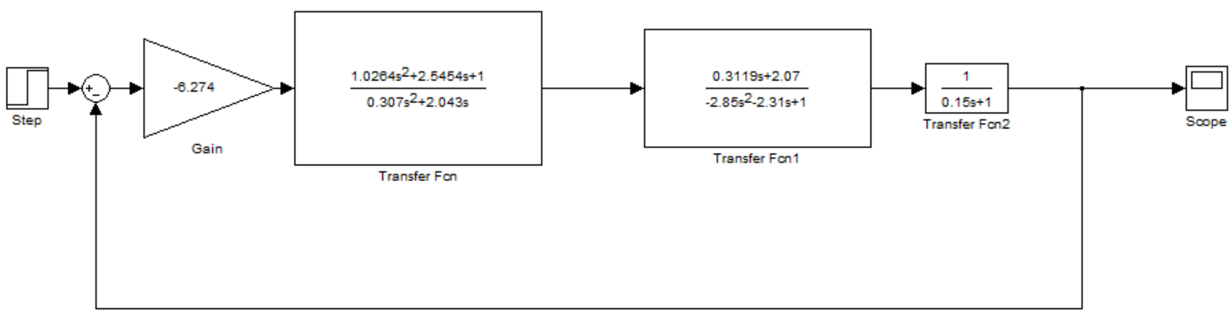


Fig 6.6: SA based PID controller for case 3

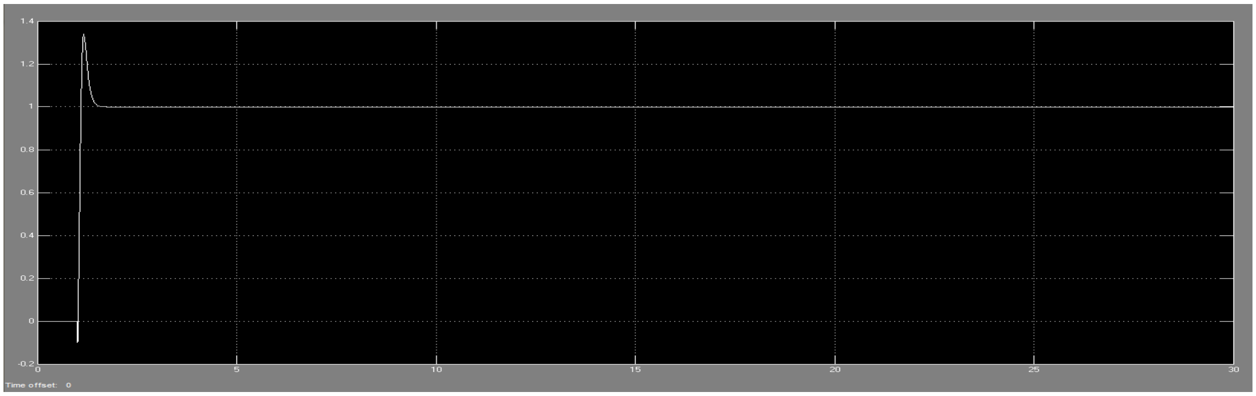


Fig 6.7: IMC based PID controller for case 1

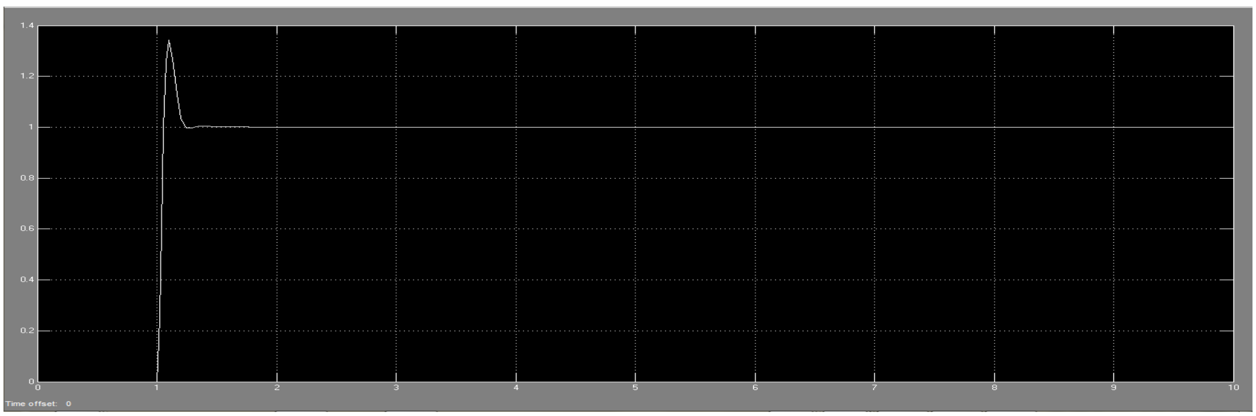


Fig 6.8: SA based PID controller for case 1

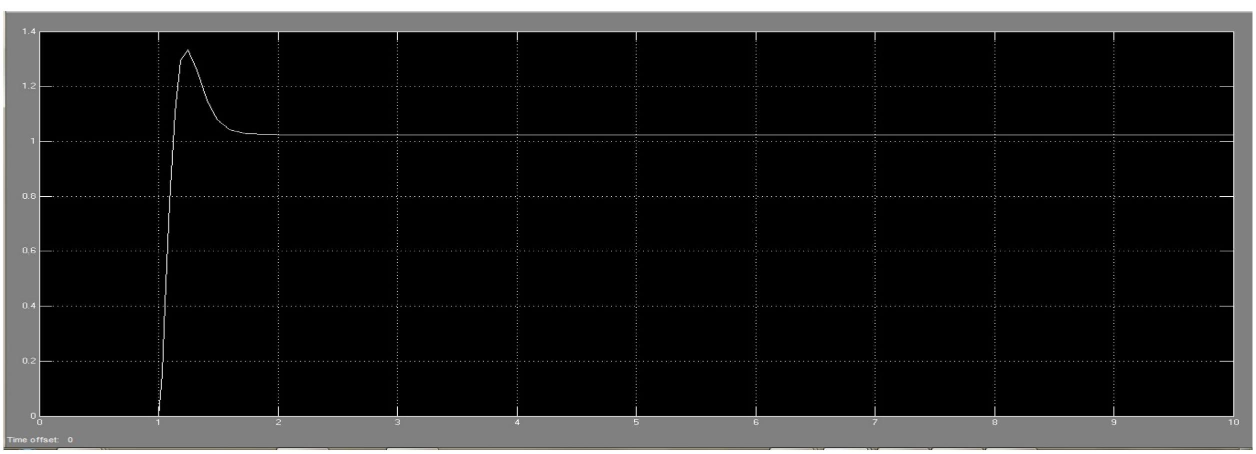


Fig 6.9: IMC based PID controller for case 2

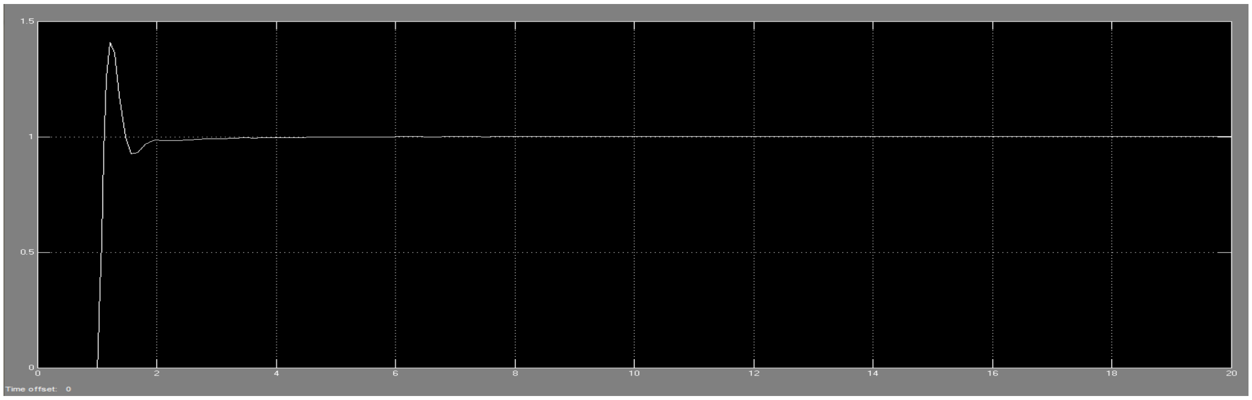


Fig 6.10: SA based PID controller for case 2

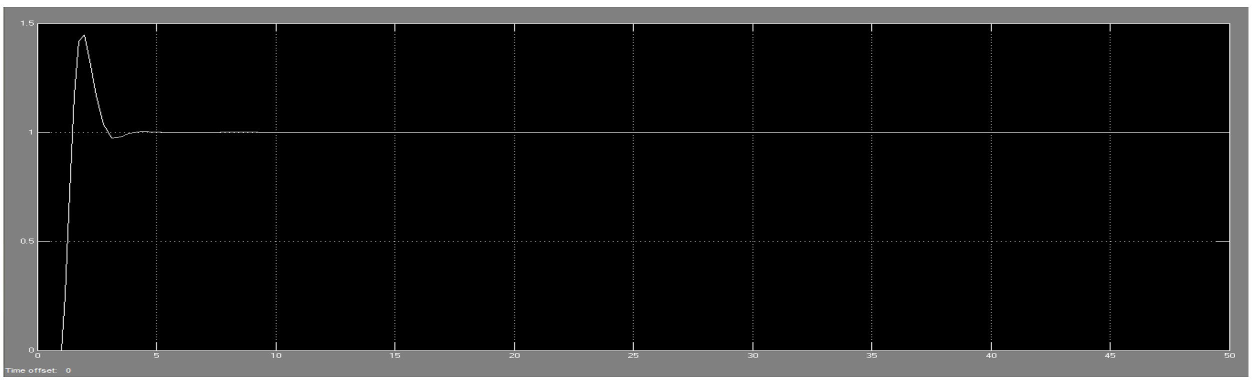


Fig 6.11: IMC based PID controller for case 3

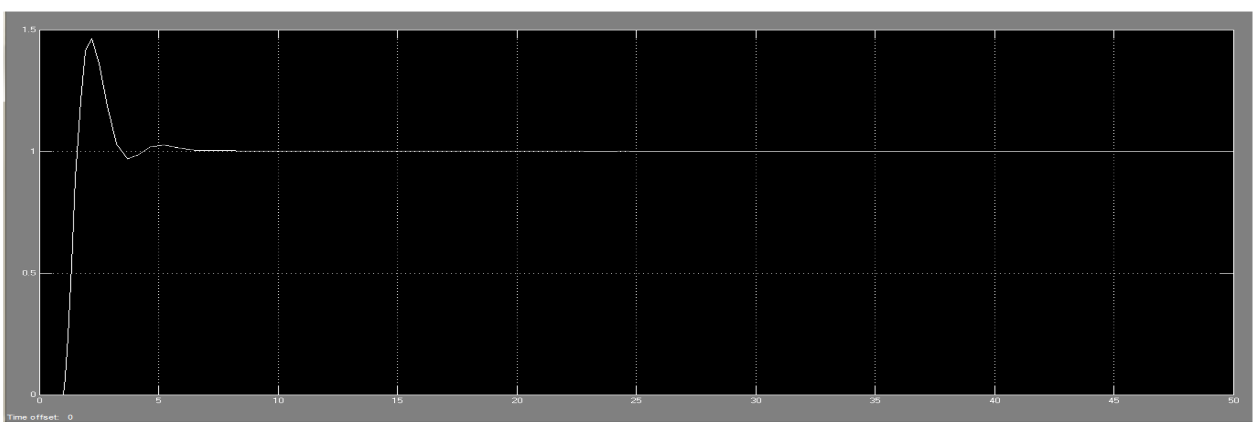


Fig 6.12: SA based PID controller for case 3

Chapter 7:

Conclusion

From the above project, we came across various type of controllers like Proportional controller(P), Proportional-Integral Controller(PI), Proportional-Integral-Derivative(PID) and we have also done the simulation of different type of control strategy of a single, double and triple tank level using Feedback, Feed-forward, Feedback plus Feed-forward, cascade controls; INTERACTING & NON-INTERACTING LEVEL PROCESS ; and BOILER DRUM.

We learnt about the different tuning methods for PID Controller such as Ziegler-Nichols, Tyreus-Luyben, Cohen-Coon, Minimum error criteria (IAE, ISE). We did a complete study of PID controller of boiler drum by different methods and studied the response of it. We proposed Two controller design methods based on internal model control (IMC) principles and stability analysis (SA) method for unstable SOPTD system. As we observed from Simulation results in SIMULINK, the controllers designed by proposed methods perform better on non-linear model equation of jacketed CSTR carrying out irreversible first order chemical reaction.

References

[1] C. A. Smith and A. B. Corripio, “Principles and Practice of Automatic Process Control”, John Wiley & Sons, Inc., ISBN 0-47 1-88346-8.

[2] Bequette, B. W. 2003. “*Process control, Modelling, design and simulation*”. Prentice Hall India, New Delhi.

[3] K. Ghousiya Begum, D. Mercy, H. Kiran Vedi, M. Ramathilagam, *An intelligent model based level control of Boiler Drum”,* Proc. of IJETAE, volume 3, Issue 1, January 2013.

[4] D. Krishna, K. Suryanarayan, G. Aparna, R. Padma Sree, “*Tuning of PID Controllers for Unstable Continuous Stirred Tank Reactors*”, Proc. of IJASE, 2012.10.1.

[5] P. Srinivas, K. Vijaya Lakshmi, V. Naveen Kumar, “*A Comparison of PID Controller tuning methods for three tank level process* ”, Proc. of IJAREEIE, Volume 3, Issue 1, January 2014.