Identification of an Effective Controller for a Stirred Tank Heater Process

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Abstract*— *Safety and the satisfaction of production specifications are the two primary operational or functional objectives for a chemical plant. Once these two important factors are achieved , the next goal is to make the plant more profitable. Given the fact the conditions which affect the operation of the plant do not remain the same, it is clear that it is mandatory that the operation of the plant is changed in order to maximize the economic objective . This task is undertaken by the controllers of the plant. These controllers are the subject of interest in this paper , where a chemical process like a stirred tank heater is controlled using the PID , the IMC based PID and the adaptive controller. A mathematical model of the stirred tank heater is developed and the different control mechanisms are applied to it. A simulation study is carried out using MATLAB to control the process system using the above mentioned control techniques. With the help of the simulation studies and the time integral performance criteria, we can deduce which controller is the most suitable for a stirred tank heater system

Index Terms*—*Adaptive Control, IMC based PID Control, PID Control, Stirred Tank Heater

I. INTRODUCTION

A chemical plant is an arrangement of processing units like reactors, heat exchangers, pumps, distillation columns, absorbers, tanks, etc. integrated with one another in systematic manner. The primary objective of the plant is to convert certain raw materials into desired products using available sources of energy, in the most economical and productive way. During its operation , a chemical plant must satisfy several requirements imposed by its designer and the general technical, economic, and social condition in the presence of dynamic external disturbances. Among such important requirements are the safety, production specification ,environmental regulation, operational constraints and economics. All the requirements listed above demonstrate the need for continuous monitoring of the operation of a chemical plant and external intervention to guarantee the satisfaction of the operational objectives. This is accomplished through a meticulous and careful arrangement of various equipments like the measuring devices, valves, controllers ,computers and human intervention (plant designer, plant operator) which together constitute the control system.

In order to analyse the behavior of a chemical process and to study about its control, a mathematical representation of the physical and chemical phenomenon taking place in the process has to be developed. Such a mathematical representation constitutes the model of the system, while the activities leading to the construction of the model will be referred to as modeling.

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Modeling a chemical process is a very important activity, requiring the use of all the basic principles of the chemical engineering , such as thermodynamics, kinetics, transport phenomenon, etc. For design of controllers for chemical processes, modeling is a very critical step. It should be dealt with great amount of care.

A. Need for mathematical modeling

It is quite difficult and very much expensive to construct a physical equipment for a chemical process as it can be costly. Consequently the experiment to determine how the process reacts to various inputs cannot be designed and thus also the appropriate control system. And even if the process equipment is available for experimentation , the procedure is usually very costly. Thus a mathematical model gives a simple description of how the process reacts to various inputs and this description is very useful for the control designer.

II. THE STIRRED TANK HEATER

Mixing vessels are used in many chemical processes. These mixing vessels are heated by either a coil or a jacket surrounding the vessel. For example, a mixing vessel may function as a reactor, where two or more components are reacted to produce one or more products. These reactions have to occur at a certain temperature in order to attain the desired output. The temperature in the process vessel is maintained by varying the flow rate of a fluid through the jacket or coil.

Consider a stirred tank heater as mentioned in the Figure below , where the tank inlet stream is received from another process unit. The objective is to maintain the tank temperature at a desired point. A heat transfer fluid is circulated through a jacket to heat the fluid in the tank. But in some processes steam is used as the heat transfer fluid and most of the energy transported is due to the phase change of steam to water. Some other processes use a heat transfer fluid . Here an assumption is made that no change of phase occurs in either the tank fluid or the jacket fluid. If the phase change occurs , then it would lead into a different concept and thus the heating of the liquid inside the tank might not be good. As a result of this the design consideration may change , because the jacket input flow rate needs to be increased.

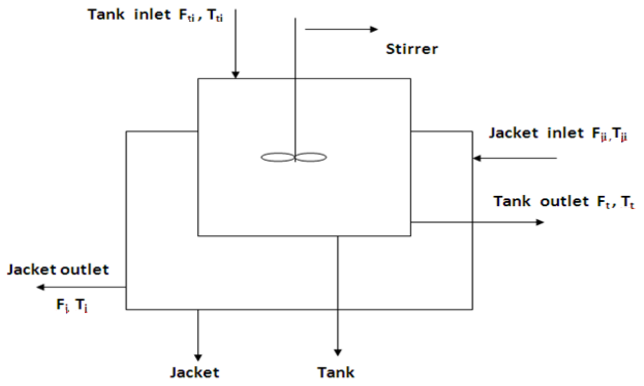

Fig : The Stirred Tank Heater

A. *Developing the mathematical model*

Here the dynamic modeling equations are written to find the tank and jacket temperatures. In order to write these equations we make the following assumptions :

a) The volume ,density and heat capacity of the liquid remains constant.

b) Perfect mixing is assumed in both the tank and the jacket.

c) The tank inlet flowrate, jacket flowrate, tank inlet temperature, and jacket inlet temperature may change (these are the inputs).

d) The rate of heat transfer from the jacket to the tank is governed by the equation $Q = U.A(Tj - Tt)$, where U is the overall heat transfer coefficient and A is the area for heat transfer where Tj and Tt are jacket and tank temperature respectively.

The operational objectives of the heater are :

a) To keep the effluent temperature at a desired value Tt.

b) To keep the volume of the liquid in the tank at a desired value V.

Table : Notations used in the process

VARIABLES	SUBSCRIPTS
A	Area of heat transfer
C_p	Specific heatcapacity(energy/mass*temp)
F	Volumetric flowrate (volume/time)
ρ	Density (mass/volume)
T	Temperature
t	Time
Q	Rate of heat transfer (energy/time)
U	Heat transfer coefficient(energy/time*area*temp)
V	Volume
i	Inlet
j	Jacket
ji	Jacket inlet
ref	Reference state
s	Steady state

The derivation of the mathematical model requires the following :

a) Material balance around the tank

b) Energy balance around the tank

c) Material balance around the jacket

d) Energy balance around the jacket

Using the conditions and the assumptions we can derive the mathematical model of the stirred tank heater. It is given as follows

$$\frac{dT_t}{dt} = \frac{F_t}{V_t}\left(T_j - T_t\right) + U.A.\frac{(T_j - T_t)}{V_t \rho_t C_{Pt}} \quad \text{------ (1)}$$

$$\frac{dT_j}{dt} = \frac{F_j}{V_j}\left(T_{ji} - T_j\right) - U.A.\frac{(T_j - T)}{V_j \rho_j C_{Pj}} \quad \text{------ (2)}$$

B. *Steady state conditions*

Before linearizing the non linear model to find the state-space form, the state variable values at steady-state are found . The steady-state is obtained by solving the dynamic equations for dx/dt = 0. The steady-space values of the system variables and some parameters for this process are given in the Table

Table : Parameters and steady state values

$F_s = 1\ ft^3/min$	$\rho C_p = 61.3$ Btu/°F- ft^3	$\rho_j C_{Pj} = 61.3$ Btu/°F- ft^3
$T_{is} = 50°F$	$T_s = 125°F$	$V = 10\ ft^3$
$T_{jis} = 200°F$	$T_{js} = 150°F$	$V_j = 1\ ft^3$

C. *State space model*

Here the modeling equations are linearized to find the state space form .The linearization of the non linear modeling equations requires the Taylor series expansion theorem. Thus here in the Taylor series ,only the linear terms are retained and the other terms are neglected Recall that our two dynamic functional equations are:

$$\frac{dT_t}{dt} = f_1\left(T_t, T_j, F_t, F_j, T_{ti}, T_{ji}\right) = \frac{F_t}{V_t}\left(T_j - T_t\right) + U.A.\frac{(T_j - T_t)}{V_t \rho_t C_{Pt}}$$

----- (1)

$$\frac{dT_j}{dt} = f_2\left(T_t, T_j, F_t, F_j, T_{ti}, T_{ji}\right) = \frac{F_j}{V_j}\left(T_{ji} - T_j\right) - U.A.\frac{(T_j - T)}{V_j \rho_j C_{Pj}}$$

----- (2)

Now the Taylor series expansion is performed ,retaining only the linear terms and neglecting other terms

$$\dot{T}_t = f_1 + \frac{\partial f_1}{\partial T_t} T_t + \frac{\partial f_1}{\partial T_j} T_j + \frac{\partial f_1}{\partial F_t} F_t + \frac{\partial f_1}{\partial F_j} F_j + \frac{\partial f_1}{\partial T_{ti}} T_{ti} + \frac{\partial f_1}{\partial T_{ji}} T_{ji} \quad \text{------ (3)}$$

$$\dot{T}_j = f_2 + \frac{\partial f_2}{\partial T_t} T_t + \frac{\partial f_2}{\partial T_j} T_j + \frac{\partial f_2}{\partial F_t} F_t + \frac{\partial f_2}{\partial F_j} F_j + \frac{\partial f_2}{\partial T_{ti}} T_{ti} + \frac{\partial f_2}{\partial T_{ji}} T_{ji} \quad \text{------- (4)}$$

on solving the equations by substituting the steady state values we get

$$\dot{T}_t = 0 + [0.4]\,T_t + [0.3]\,T_j + [-7.5]\,F_t + [0]\,F_j + [0.1]\,T_{ti} + [0]\,T_{ji} \quad \text{------ (5)}$$

$$\dot{T}_j = 0 + [3]\,T_t + [-4.5]\,T_j + [0]\,F_t + [50]\,F_j + [0]\,T_{ti} + [1.5]\,T_{ji} \quad \text{------ (6)}$$

Now the state space representation is used in which the values obtained are represented in the matrix form.

Thus the state space representation is given by:

$\dot{x}$ = A x + B u
y = C x + D u

Where 'x' is the state variables, 'u' is the input variable , 'y' is the output variables

'A' is the state matrix

$$A = \begin{bmatrix} -0.4 & 0.3 \\ 3 & -4.5 \end{bmatrix}$$

'B' is the input matrix

$$B = \begin{bmatrix} 0 & -7.5 & 0.1 & 0 \\ 50 & 0 & 0 & 1.5 \end{bmatrix}$$

'C' is the output matrix

$$C = \begin{bmatrix} 1 & 0 \\ 0 & 1 \end{bmatrix}$$

'D' is the translational matrix

$$D = \begin{bmatrix} 0 & 0 & 0 & 0 \\ 0 & 0 & 0 & 0 \end{bmatrix}$$

Now substitute the matrices in the state space representation and perform the transfer conversion we get eight transfer functions

The different transfer functions obtained for tank temperature with variation in different input parameters are given below

$$\frac{T_t(s)}{F_t(s)} = \frac{-7.5-33.75}{s^2+4.9s+0.9} \qquad \frac{T_t(s)}{T_{ti}(s)} = \frac{0.1s+0.45}{s^2+4.9s+0.9}$$

$$\frac{T_t(s)}{F_t(s)} = \frac{15}{s^2+4.9s+0.9} \qquad \frac{T_t(s)}{T_{ji}(s)} \frac{0.45}{s^2+4.9s+0.9}$$

The different transfer functions obtained for jacket temperature with variation in different input parameters are given below

$$\frac{T_j(s)}{F_t(s)} = \frac{-22.5}{s^2+4.9s+0.9} \qquad \frac{T_j(s)}{T_{ti}(s)} = \frac{0.3}{s^2+4.9s+0.9}$$

$$\frac{T_j(s)}{F_j(s)} = \frac{50s+20}{s^2+4.9s+0.9} \qquad \frac{T_j(s)}{T_{ji}(s)} = \frac{1.5s+0.6}{s^2+4.9s+0.9}$$

But the most important is the following transfer which relates the tank temperature to the jacket in flowrate

$$\frac{T_t(s)}{F_j(s)} = \frac{15}{s^2+4.9s+0.9}$$

which can be factored as,

$$\frac{T_t(s)}{F_j(s)} = \frac{16.6667}{(0.2123s+1)(5.23207s+1)}$$

which is gain – time constant form

The transfer function relating the jacket temperature to the jacket flowrate is also considered and is given by

$$\frac{T_j(s)}{F_j(s)} = \frac{50s+20}{s^2+4.9s+0.9}$$

which can be factored as,

$$\frac{T_j(s)}{F_j(s)} = \frac{22.2(2.5s+1)}{(0.2123s+1)(5.2307s+1)}$$

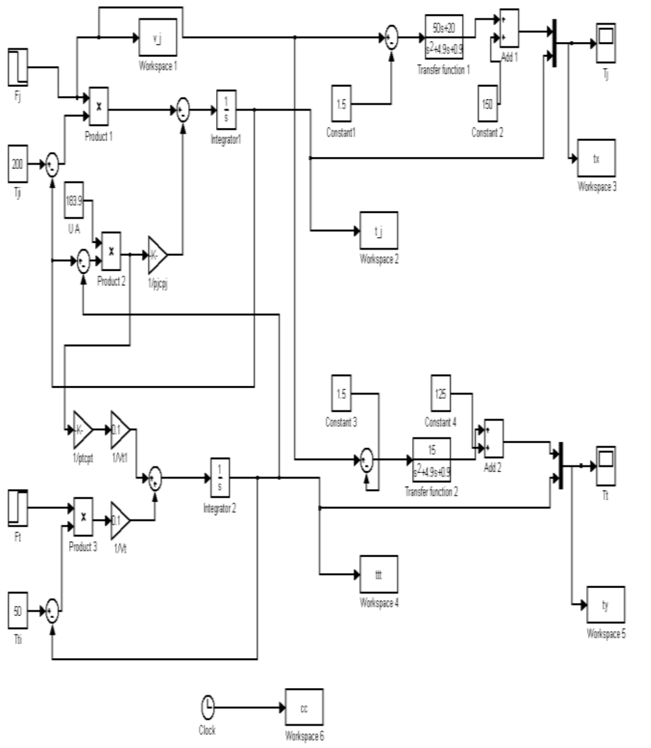

Fig : Simulink diagram for the stirred tank heater

III. LINEAR VERSUS NON LINEAR MODELS FOR THE STEP RESPONSE

A. Small increase in jacket flowrate

Consider a step change of 0.1 (from the value of 1.5 to 1.6 ft^3/min) in the jacket flowrate. The simulation responses of the nonlinear and linear models are shown in Fig (a) and Fig (b) for jacket temperature and tank temperature respectively. From Fig (a), the linear model attains set point in smooth way

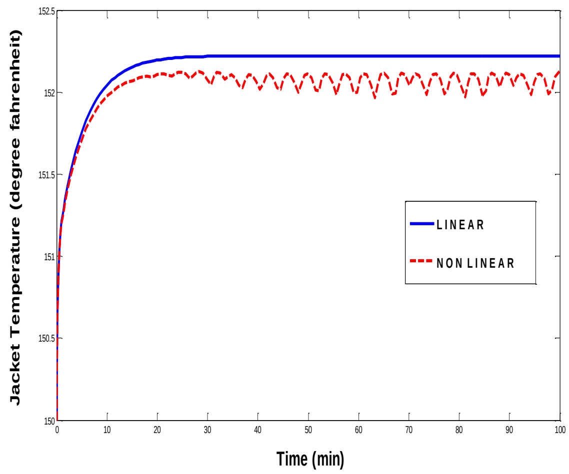

Fig (a) : Variation of the jacket temperature with $0.1ft^3$/min increase in jacket flowrate

From Fig (b) , the linear model shows good performance compared to non linear model

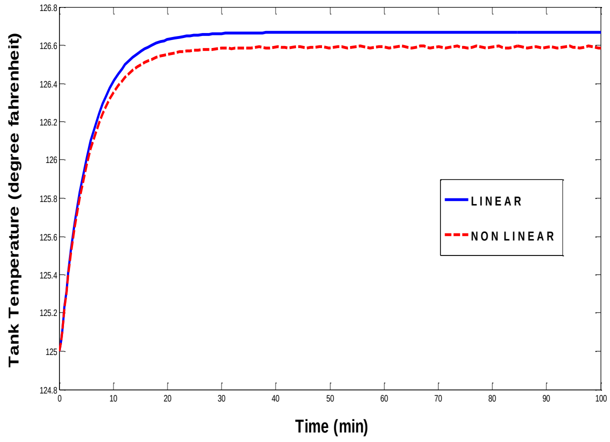

Figure (b) : Variation of the tank temperature with $0.1ft^3$/min increase in jacket flowrate

B. Large increase in jacket flowrate

Consider a step change of 1.0 (from 1.5 to 2.5 ft^3/min) in jacket flowrate. The responses of the linear and nonlinear models are shown in Fig (a) and Fig (b) for jacket and tank temperature . The gain of the linear model is greater than that of the non linear model.

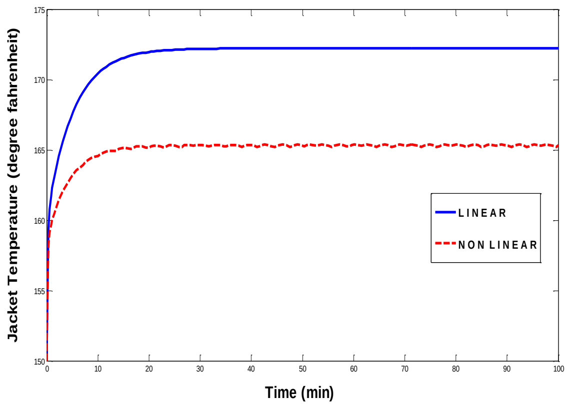

Fig (a) : Variation of the jacket temperature with respect to increase of $1ft^3$/min in jacket flowrate.

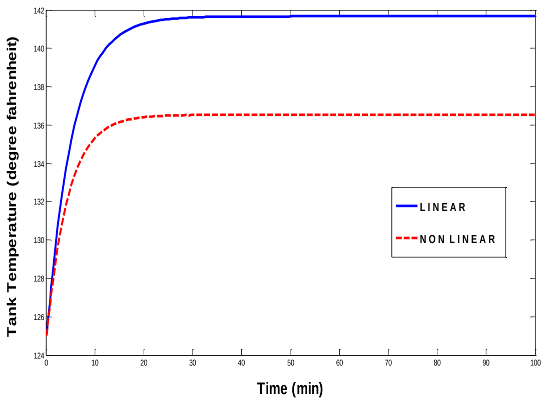

Fig (b) : Variation of the tank temperature with respect to increase of $1ft^3$/min in jacket flowrate.

C. Small decrease in jacket flowrate

Consider a step change of 0.1 (from 1.5 to 1.4 ft^3/min) in jacket flowrate. The responses of the linear and non linear models are shown in Fig (a) and Fig (b) for jacket and tank temperature respectively.

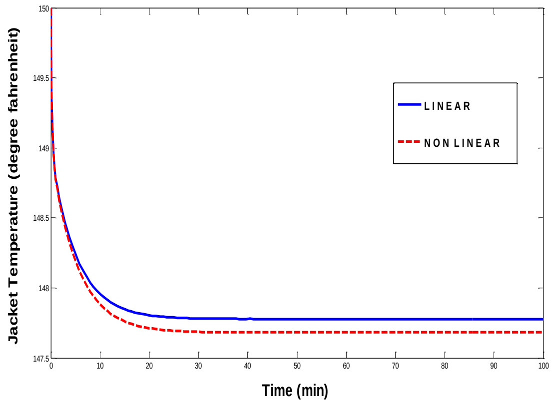

Fig (a) : Variation of the jacket temperature with respect to decrease of $0.1ft^3$/min in jacket flowrate

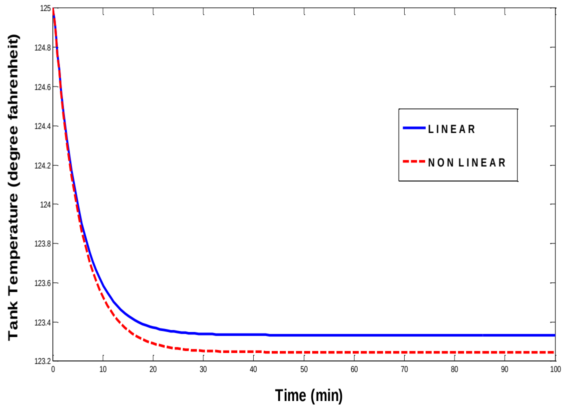

Fig (b) : Variation of tank temperature with $0.1ft^3$/min decrease in jacket flowrate

D. Large decrease in jacket flowrate

Consider a step decrease of 1.0(from 1.5 to 0.5 ft^3/min **)** in jacket flowrate. The responses of the linear and non linear models are shown in Fig (a) and Fig (b) for jacket and tank temperature respectively. The gain of the linear model is greater than gain of the non linear model showing good performance.

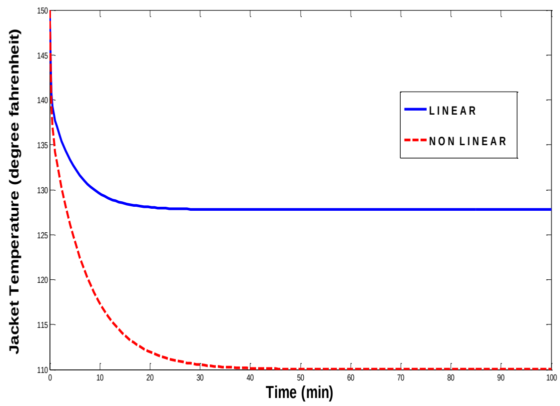

Fig (a) : Variation of jacket temperature with $1ft^3$/min decrease in jacket flowrate

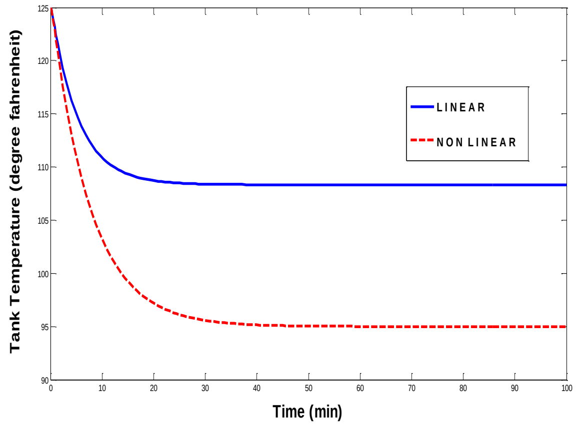

Fig (b) : Variation of tank temperature with $1ft^3$/min decrease in jacket flowrate

IV. THE PID CONTROL FOR THE STIRRED TANK HEATER

The PID controller is the most common control algorithm for feedback loops . It is implemented in many different forms, as a part of a DDC (Direct Digital Control) package or a hierarchical distributed process control system. Many control engineers worldwide are using such controllers in their daily work. The PID algorithm can seen from many different directions. It can be looked at as a device that can be operated with a few rules, but it can also be approached analytically.

There are three types of controller modes. They are

a) Discontinuous controller modes

This is a type of a controller mode that shows discontinuous changes in controller output as controlled variable error occurs. The two position control mode, multi position mode and the floating control mode are the discontinuous controllers modes

b) Continuous controller modes

The most common controller action used in process control is one or a combination of continuous controller modes. In these modes , the output of the controller changes smoothly in response to the error or the rate of change of error. These modes are an extension of the discontinuous types.

The proportional , integral , derivative modes are the continuous controller modes

c) Composite controller modes

It is very common in the complex of industrial processes to find control requirements that do not fit the application norms of any of the previously considered controller modes. It is both possible and expedient to combine several basic modes, thereby combining the advantages of each mode. In some cases , an added advantage is that the modes tend to eliminate some limitations they individually posses. The PI , PD and PID are examples.

A. Proportional – Integral – Derivative mode (PID)

One of the most powerful but complex controller mode operations combines the proportional , integral and derivative modes. This system can be used for virtually any process condition. The analytic expression is

$$p = K_P\, e_p + K_P\, K_I \int_0^t e_p(t)dt + K_P\, K_D \frac{d_{e_p}}{dt} + p_I(0)$$

where ,

K_P is the proportional gain between error and controller output(% per%)

K_I is the constant relating the rate to the error ((% / s) / %)

K_D is derivative gain constant (% - s / %)

$\frac{d_{e_p}}{dt}$ is the rate of change of error (% / s)

$p_I(0)$ is the integral term value at t = 0 (initial value)

B. Ziegler Nichols Tuning Technique

It is a trial- and-error loop tuning technique that is still widely used today. The automatic mode (closed-loop) procedure is as follows:

Unlike the process reaction curve method which uses data from the open-loop response of the system , the Ziegler-Nichols tuning technique is a closed loop procedure.

a) The controller is set to P-Only mode and switched to automatic when the process is at the design level of operation.

b) An initial controller gain, KC , is guessed (assumed) that is good enough to keep the loop stable.

c) The set point is bumped a small amount and the response behavior observed

d) If the controller is not making the measured process variable (PV) to sustain oscillations, the KC is increased(or decrease the proportional band which is the PB) .

e) By trial and error method the KC value is found out. . These oscillations should neither be growing nor dying out , and the controller output (CO) should remain unconstrained.

f) The controller gain at this condition is called the ultimate gain , KU. The period of the PV oscillation pattern at the ultimate gain is called the ultimate period , PU.

g) Using the values of KU and PU ,Ziegler and Nichols recommended the following settings for feedback controllers to find the controller parameters.

Table : Zeigler Nichols tuning parameter settings

Controller modes	K_C	τ_I	τ_d
Proportional	$K_U/2$	-	-
Proportional-integral	$K_U/2.2$	P_U /1.2	-
Proportional-integral-derivative	$K_U/1.7$	P_U /2	P_U /8

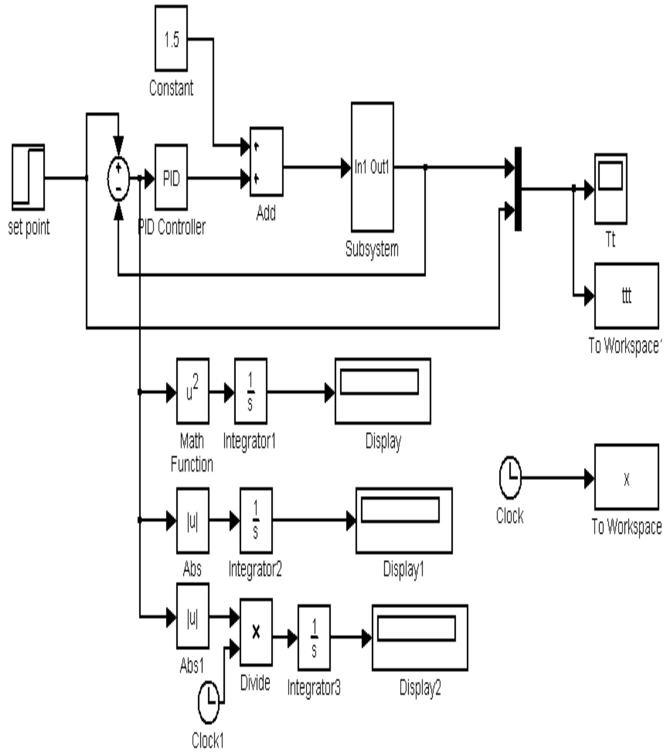

Fig : Simulink diagram for PID control of stirred tank heater

A set point of 130°F , which has to be maintained in the tank is given as an input to the process. Fig shows that the response has settled at the set point , indicating that the temperature in the tank is maintained .

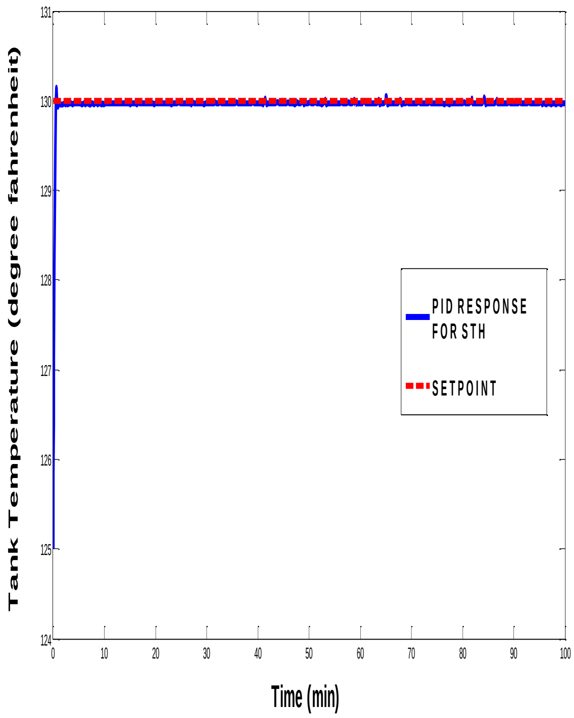

Fig : Variation of tank temperature with time for the PID control

A set point of 145°F is given to the process. Figure 3.5 shows that the response has approximately settled at the set point

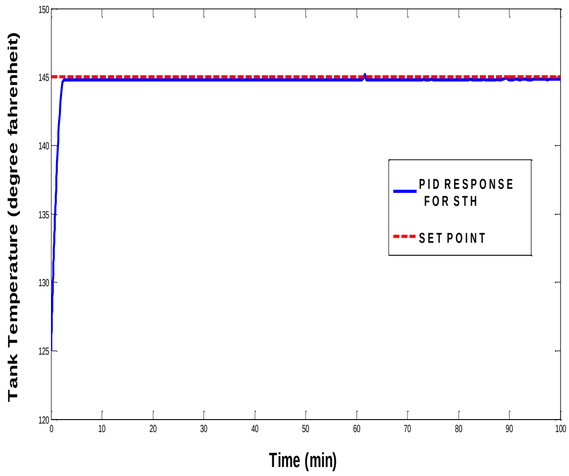

Fig: Variation of tank temperature with time for the PID control

V. INTERNAL MODEL CONTROLLER (IMC) BASED PID FOR THE STIRRED TANK HEATER

The closed-loop oscillation technique developed by Ziegler and Nichols did not require a model of the process. Direct synthesis, however, was based on the use of a process model and a desired closed-loop response to synthesize a control law; often this resulted in a controller with a PID structure. Here a model-based procedure is developed, where a process model is "embedded" in the controller. By explicitly using process knowledge and by virtue of the process model, better performance is obtained.

The main advantage to IMC is that it provides a transparent framework for control-system design and tuning. This is pleasing because standard equipment and algorithms (i.e., PID controllers) are used to implement an "advanced" control concept . If the controller and the process is stable then the overall controlled system is also stable. This is a good result because in a standard feedback control formulation, the controller and the process can each be stable, yet the feedback system may be unstable. IMC is able to compensate for disturbances and model uncertainty, while open-loop control is not capable of doing so..

Table : List of transfer function variables

Transfer function variables	Abbreviations
d(s)	Disturbance
$\tilde{d}(s)$	estimated disturbance
$g_p(s)$	process
$\tilde{g}_p(s)$	process model
q(s)	internal model controller
r(s)	Set point
$\tilde{r}(s)$	modified setpoint (corrects for model error and disturbances)
u(s)	manipulated input (controller output)
y(s)	measured process output
$\tilde{y}(s)$	model output

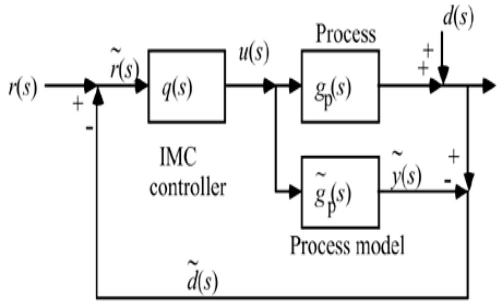

Fig : The IMC structure

A. The implementation of the IMC based PID for the stirred tank heater process

The transfer functions which are obtained for the tank temperature and for the jacket temperature is used in designing the IMC based PID technique. This IMC based PID process has specific formulae for different model process equations and with the help of these equations the controller can be designed.

The transfer functions of stirred tank heater are as follows

$$\frac{T_t(s)}{F_j(s)} = \frac{15}{s^2+4.9s+0.9} \quad \text{----- 1}$$

$$\frac{T_j(s)}{F_j(s)} = \frac{50s+20}{s^2+4.9s+0.9} \quad \text{----- 2}$$

The most important among these two transfer functions is equation 1 as it helps in controlling the tank temperature with the help of the jacket input flowrate

Thus a set of formulae to calculate the PID values for the controller is given below.

$$g_p(s) = \frac{k_p}{(\tau_1 s+1)(\tau_2 s+1)} \quad , \quad k_c = \frac{\tau_1+\tau_2}{k_p \lambda}$$

$$\tau_I = \tau_1 + \tau_2 \ , \quad \tau_D = \frac{\tau_1 \tau_2}{\tau_1+\tau_2}$$

Where $g_p(s)$ is the process model , k_c is proportional constant, τ_I is integral time constant, τ_D is derivative time constant.From the above equations the value for k_p is 16.6667 , τ_1 is 0.2123 and τ_2 is 5.23207.For finding the value for 'λ' the following formula is used : $\lambda = \frac{T_s}{4}$, where T_s is known as the settling time which is the time needed for the response to reach $\pm$5% of the final value and to stay there. From the output response graphs from chapter 1, the settling time is found and substituted in the equation given above to find the value of λAfter finding λ , value of k_c is found with the available values of k_p , τ_1 , τ_2 and λ . Similiarly the values for τ_I and τ_D is found out using τ_1 and τ_2 After finding the values for k_p , τ_I , τ_D the values for the integral constant k_I and for the derivative constant k_D has to be found. This can be done with the help of k_c where , $k_I = \frac{k_c}{\tau_I}$ and $k_D = k_c . \tau_D$ Thus now the values for the proportional constant k_p , the derivative constant k_D and the integral constant k_I are found These values have then to be used in SIMULINK block diagram for finding the response for a given set point

The values for the proportional ,integral and derivative constants are $k_p = 16.6667$,$k_I = 0.0057$, $k_D = 0.1526$

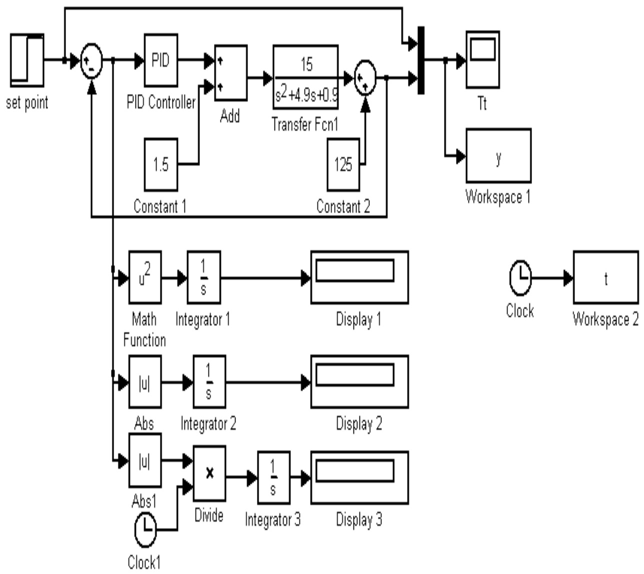

Fig :Simulink for IMC based PID for the stirred tank heater

In the Fig above the set point is given as 135°F and the IMC controller ensures that the response settles at the set point. It can be seen from the Fig below that the system response has settled approximately at 135°F . Thus the IMC based PID controller shows good response for the given set point

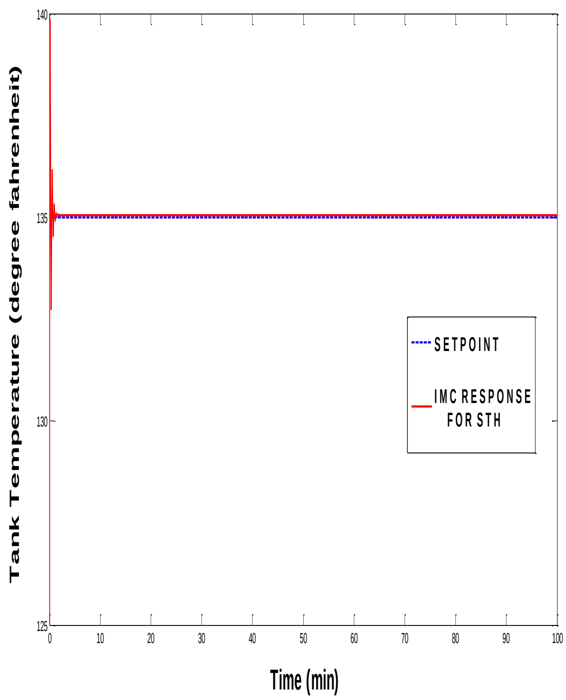

Fig : Variation of tank temperature with time for the IMC based PID controller

A set point of 150°F is given to the controller and it can be clearly seen that the IMC response settles at the desired setpoint

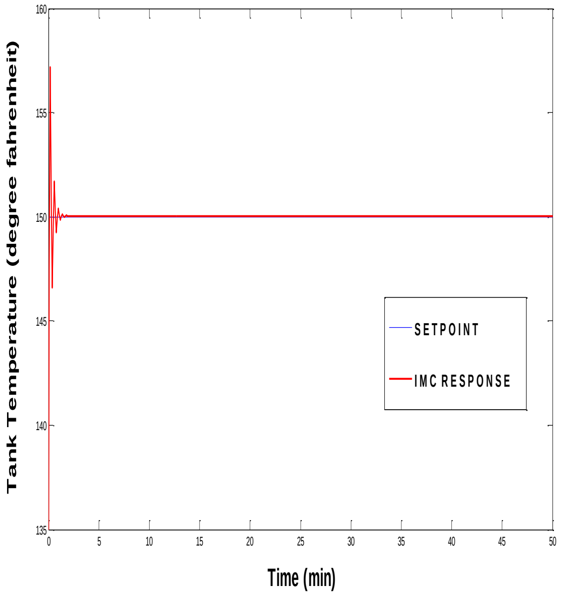

Fig : Variation of tank temperature with time for the IMC based PID controller

VI. THE ADAPTIVE CONTROLLER FOR THE STIRRED TANK HEATER

A. Gain scheduling technique

Adaptive control is a control system, which can adjust its parameters and internal settings automatically in such a way as to compensate for variations in the characteristics of the process it controls. The various types of adaptive control systems differ only in the way the parameters of the controller are adjusted. There are two important reasons why we choose adaptive control for a chemical or thermal process. First, most chemical processes are non-linear . Therefore , the linearized models that are used to design linear controllers depend on the particular steady state (around which the process is linearized). It is clear, then, that as the desired steady-state operation of a process changes, the controller parameters change. This explicitly asks for the need for controller adaptation. Second , most of the chemical processes are non stationary (that is their characteristics change with time).

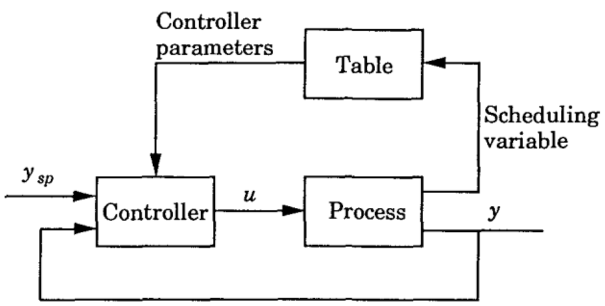

Fig : Gain scheduling technique

B. The adaptive control design procedure

The adaptive control is control technique , which controls a process in spite of the dynamic environment. Here , the adaptive control of the stirred tank heater is performed. It is done with the help of the gain scheduling technique. The main aim of the process is to control the tank temperature with variation in jacket flowrate. Thus gain scheduling is a principle in which a scheduling variable is found and its range is quantitized into number of discrete operating conditions. The controller parameters are then determined by Ziegler Nichols tuning . The procedure is repeated until all operating conditions are covered.

One of the major parameter that affect the tank temperature is the inflow rate and hence it is chosen as the scheduling variable . Now the range of this scheduling variable is divided into smaller segments and the controller parameters ie the proportional , integral and the derivative constants are found out for each of the smaller segments using the Ziegler Nichols method.

The values are then given to a feedback loop having a PID block. The PID values are tuned to get the optimized parameter values.

An embedded MATLAB function is used where the smaller ranges and its PID values are enclosed . If the range of the scheduling variable is exceeded , then the adaptive controller fails to work and the tank temperature be controlled

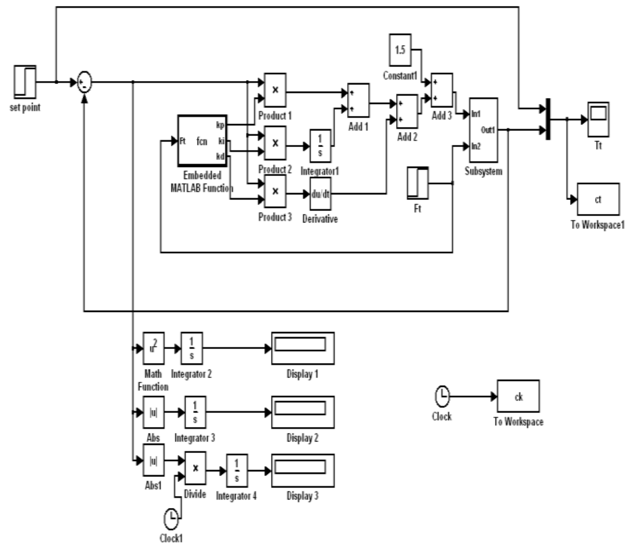

Fig : Simulink diagram for the Adaptive control for stirred tank heater

A set point of 135°F is given. The Fig below shows that the adaptive control response settles at the set point. Thus this proves that the controller has been designed perfectly

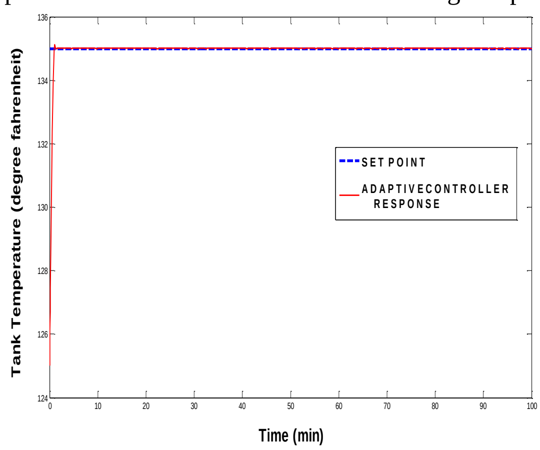

Fig : Variation of tank temperature with time for the adaptive controller

A set point of 140°F is given to the process. The Fig above shows adaptive controller response settling at the set point indicating the fact that the controller is designed perfectly

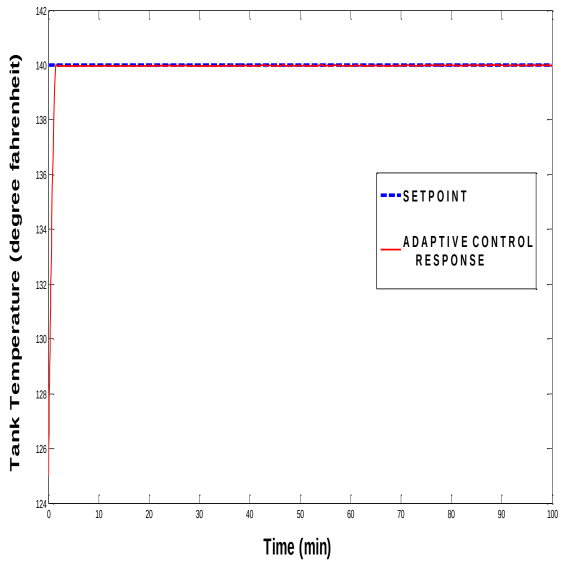

Figure 5.5 : Variation of tank temperature with time for the adaptive controller

VII. RESULTS

When small input changes were made in jacket liquid flow rate, the gain did not change much from the linear model case. For large input changes, the gain of the non linear system was less than the linear system for increases in jacket flow rate, but higher than the linear system for decrease in jacket flow rate. The time integral performance can be used to evaluate which controller is the most suitable for the stirred tank heater

Table : Time integral performance of the controllers

Time integral performance criteria	PID	IMC	ADAPTIVE
ISE	27.6	8.6	24.99
IAE	11.31	7.091	4.286
ITAE	342.4	264.5	18.64

The PID control for the stirred tank heater process has given a high performance as the offset is negligible. The Ziegler Nichols closed loop principle has been very useful in designing the PID controller. But on rise of the temperatures, the offset is increased which is undesirable. Thus its errors are on a higher side as shown in Table .This leads to the design of some other controller where the errors can be less.

The IMC based PID control was designed perfectly as the response graph settles at the set point. Thus its performance is highly satisfactory when compared to that of the PID tuning. The main advantage of IMC is that it provides a transparent framework for control-system design and tuning. Clearly it has ISE, IAE and the ITAE values lesser than the PID, showing that it is the better than PID

The adaptive controller is by far the most useful controller for a chemical process . Thus it is noted that for any value of tank input flowrate within the specified range, the controller works effectively. For a value exceeding the specified range , it is proved that the controller cannot control such a process. The IAE and the ITAE values for the adaptive controller are lesser than the PID and the IMC based PID controller. The stirred tank heater being a non stationary process , the adaptive controller finds its maximum use here. The gain scheduling technique is used in the controller design. It has the advantage that it can follow rapid changes in the operating condition

VIII. CONCLUSION AND FUTURE SCOPE

Thus by calculating the time integral performance criteria for the three controllers used in the paper , we can conclude that the gain scheduling adaptive control is the best controller among the three.By incorporating this control technique, productivity as well as cost of operation of a stirred tank heater can well be reduced.

a) In designing the adaptive controller , apart from the gain scheduling technique , the model reference adaptive control technique and the self tuning regulator technique can be implemented . The main advantage is that the comparison of the various control techniques can be done.

b) Here the adaptive control can be used only if the tank inflow rate is within the limits of 1 ft^3. Thus the adaptive control has to be designed to accommodate other important operating conditions apart from the inflow rate.

c) A stirred tank heater can be considered where , the volume of the fluid in the tank and the jacket can vary . Formulating the mathematical model incorporating the above condition can be challenging.

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