

Self-tuning Regulator Based Cascade Control for Temperature of Exothermic Stirred Tank Reactor

Tesfabirhan Shoga

Lecturer
Jimma Institute of Technology
School of Electrical and Computer
Engineering
Ethiopia

Amruth Ramesh Thelkar

Assistant Professor
Jimma Institute of Technology
School of Electrical and Computer
Engineering
Ethiopia

C Bharatiraja

Associate Professor
SRM Institute of Science and Technology
Department of Electrical and Electronics
and Visiting Researcher, Engineering
Department of Electrical, University of
South Africa

Sisay Mitiku

Lecturer
Jimma Institute of Technology
School of Electrical and Computer
Engineering
Ethiopia

Yusuff Adedayo

Associate Professor
Department of Electrical, University of
South Africa

In the process of chemical industries production, an exothermic stirred tank reactor permits the reaction to release the heat due to mixing of substances. The release of heat due to the exothermic process in the stirred tank makes the system nonlinear and uncertain unless it is not regulated properly. In order to regulate the heat release, the self-regulated cascade control model of reactor temperature plant is developed with reference to material and energy balance model of exothermic stirred tank reactor. In this paper the recursive least square algorithm is used for self-tuning regulator cascade control online temperature plant and its controller parameters estimation and also the temperature plant output has been simulated in the presence and absence of a primary self-tuning controller. The use of the secondary loop PI controller will respond rapidly to the flow rate of cooling jacket plant that in turn cascade with the disturbance variables like inlet feed concentration and cooling water temperature. The reactor temperature is monitored to the desired steady state level and becomes stable for further steady state value of parameters. The parameters of the temperature plant will also be stable with online parameter estimating loop of the control system. The designing of master controller and desired plant model using minimum degree pole placement algorithm is highlighted. With the developed SIMULINK design of self-tuning controller, the better comparative results have been achieved over Internal Model Control (IMC), Model Predictive Control (MPC) and conventional based temperature controller plant.

Keywords Exothermic Stirred Tank reactor, Cascade control, RLS Algorithm, Self-tuning controller, Pole placement Algorithm.

1. INTRODUCTION

The development of chemical production industries has become the expanding environment-friendly technology for a country to enhance its national income through the desired manufactured products. A chemical reactor is one which plays an important role in the reaction of chemical compounds of different composition to increase the production process. It allows the reactants to be mixed effectively so as to generate a well manufactured desirable product at a specified production rate [1,2]. The stirred tank reactor allows an exothermic reaction which releases heat energy so that the reactor temperature will increase, so fast [3,4]. Due to the production of large increment in temperature, the heat generated on the reactor highly affects the proper operation and the life span of the reactor. This may result in unexpected loss of the product. Since the temperature of exothermic stirred tank reactor exhibits non-linear properties. The rising of temperature should be controlled well by applying certain appropriate control mechanism a for offering more stable output.

Self-tuning regulator advanced control strategy is used to resolve the problem on the reactor due to the time varying fluctuation of temperature. Associated with the cascade model of stirred tank reactor, self-tuning regulator will be able to make better control of the non-linear reactor temperature.

This advanced control technique has the capability to provide a response for the plants having non-linear dynamic characteristics. In such control process, the reactor temperature is controlled properly in such a way that the plant parameters are adjusted using RLS estimation algorithm and the desired control action can be determined online. On the other hand such control system provides good tracking of the plant output with the set point with minimum error and better transient characteristics when it is compared with other control techniques. This paper is tending to describe self-tuning regulator advanced control to provide a solution by enhancing the stability of the temperature of the chemical reactor through its intermediate process dynamics control. In conjunction with cascade configuration of the reactor temperature plant model, self-tuning regulator involves online monitoring and controlling of process parameters. The reactor temperature is considered as controlled variable, while the flow rate of liquid within cooling jacket will be manipulated variable. However, the inlet feed concen-

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Correspondence to: C. Bharatiraja, Associate Professor,
Department of Electrical & Electronics Eng. SRM
Institute of Science and Technology, India
E-mail: bharatiraja.c@ktr.srmuniv.ac.in
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tration and inlet temperature of the cooling jacket are disturbance variables of the system. In the control system, PI controller is used as a slave controller and self-tuning controller was used as the primary controller for the output. Self-tuning controller as a primary master controlling element is used for monitoring the temperature of exothermic stirred tank reactor in the cascade control system of the reactor. In such control scheme, due to the process exhibiting non-linear behavior, the entire temperature plant of the reactor tends to be uncertain. Such uncertainty can be controlled by using two major control loops. The inner loop has a plant with feedback controller, while the other loop consists of a process parameter adjustment network [5,6]. The system can be considered as an automation of the model and process design with the corresponding estimation of both plant and controller parameters for every sampling interval. The parameters of the controller need to be re-tuned after the process parameters have been adjusted effectively for each interval to increase the system performance [7,8].

The entire cascade control system configuration with master self-tuning controller is shown in figure1 with RLS estimator and Stirred Tank Reactor control law.

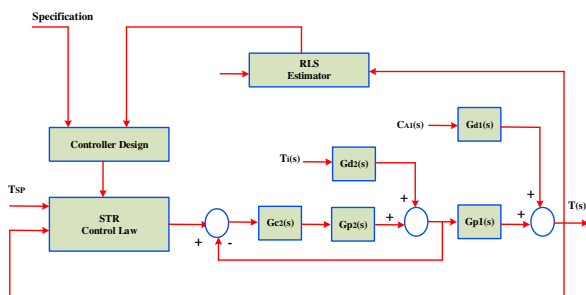


Figure 1. Block diagram of cascade control master self-tuning controller

The control strategy is based on the fact that the temperature of the reactor expressed in terms of inlet feed temperature, inlet feed concentration and flow rate of cooling jacket around the reactor. The other task was computing the stability of the desired plant model under each sub-system and the entire control process dynamics which was followed by designing the reference plant model, the primary self-tuning controller and parameter estimator for both controlled outputs. For the sake of having better control of process variables, this paper provides the ways to perform Continuous regulation of controlled variables, to determine updates or estimation of the past and present time slot plant parameters adjustment, to control both linear and non-linear plant dynamics having uncertain characteristic under different circumstances, to make online or continuous update of parameter of both plant dynamics and controller, to obtain an accurate result in finding the extent that the plant model is to be stable or not [9,10].

Besides achieving the above purposes of this work allows the chemical reactor to be effective in its operation by reducing the heat generated on its surface. Enables for chemical factories to avoid unnecessary loss or wastage of products and to deliver good quality of products to the community at any time and place. Also how the chemical reacting system is more effective and

efficient in its operation of reacting two streams of mixtures having different concentration. Since a continuous stirred tank reactor has a constant height that can react two mixtures at constant volume. With the improvement of product quality, due to continuous updating process parameters adjustment, the factors influencing the inlet mixture of the reactor will be easily adjusted to enhance the stability of the system.

The remainder of this paper is organized as follows. In Section 2, the system design procedure and methodology are outlined. Section 3 addresses the Dynamic model of exothermic stirred tank reactor. Section 4 establishes the Simulink Design of the Proposed System. Section 5 elucidates the Temperature of reactor plant output, Self-tuning controller parameters, Reactor temperature estimation error and Reactor temperature estimated parameters. Finally, the concluding remarks are given in Section 6 to summarize the contribution of this paper.

2. SYSTEM DESIGN PROCEDURES AND METHODS

The analysis of the system designs includes some systematic standard approaches such as a non-linear material and energy balance model of exothermic continuous stirred tank reactor, Development of transfer function model related with controlled variables, Representation of block diagram for cascade continuous stirred tank reactor system having secondary slave controller and primary self-tuning controller as master controller, Designing the master controller and desired plant model using minimum degree pole placement algorithm, Development of Proposed Simulink design of the entire system and comparative analysis with reference to existing and conventional model.

The method which is going to be used for controlling feed concentration and temperature of chemical reactor is adaptive self-tuning regulator. It adjusts the parameters of the dynamic plant model under the control system of cascade feedback loops having slave controllers.

2.1 Self-tuning regulators (STR)

Self-tuning regulator can be thought of as being composed of two loops. The inner loop consists of the process and an ordinary feedback controller and the other loop is a parameter adjustment loop. The parameters of the controller are adjusted by the outer loop, which is composed of recursive parameter estimator and design calculation [11,12].

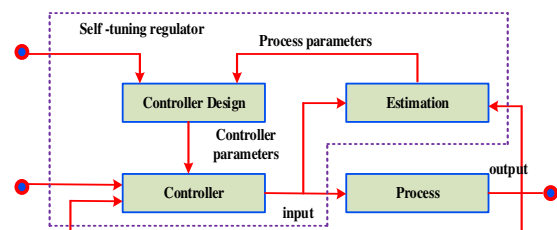


Figure 2. Self-tuning regulator block diagram

2.2 Estimation algorithms

It is important to estimate the process parameters online in adaptive control. For an adaptive control system,

the adaptive mechanism is based on identifying the system first. A self-tuning regulator in Figure.2 explicitly includes a recursive parameter estimator. The process parameters estimation is a part of system identification. In a broader sense, system identification is selection of model structure, experiment design, parameter estimation, and validation [13,14].

2.3 Least-squares estimation algorithm

Its principle is that the unknown parameters of a mathematical model should be chosen by minimizing the sum of the square of the difference between the actually observed and the analytically predicted output values with possible weighting that measure the degree of precision. The least-squares criterion is quadratic, so an analytic solution to the least-squares problem exists as long as the measured variable is linear in the unknown parameters. In adaptive control system the observations are obtained sequentially in real time. Recursive estimation algorithm is desirable. It saves the computation time by using the results obtained at previous time to get the estimates at present time. Hence, the recursive least-square (RLS) estimation method is used in this section [15,16].

2.4 Exothermic stirred tank reactor process control

The reactor temperature and concentration are monitored on the basis of the cascade control of the cooling water flow rate with the inlet feed temperature and concentration. So that, the reactor is equipped with a heat transfer surface that contains a flow of cooling water. It is required to know how the output concentration and temperature may vary with time. Figure 3 illustrates the schematics of STR. A stirred tank reactor with an exothermic reaction requires a more elaborated system model, in terms of controlled variables like temperature and concentration. Furthermore, the model is nonlinear, which forces us to make a linear approximation to solve it [17,18].

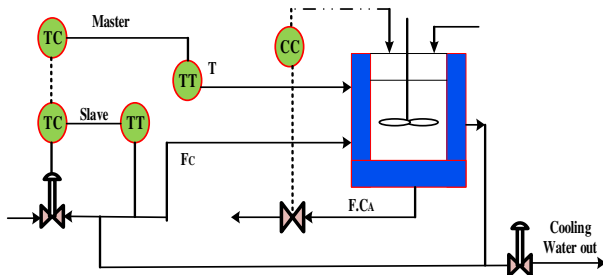


Figure 3. Schematic diagram of stirred tank chemical reactor

3. DYNAMIC MODEL OF EXOTHERMIC STIRRED TANK REACTOR

The dynamic model is developed using mole and energy balance of the given reactant A given as [19]:

$$V \frac{dC_A}{dt} = F C_{Ai} - F C_A - V(-r_A) \quad (1)$$

The equation takes a second-order kinetic rate expression for the rate of disappearance of reactant A, according to Arrhenius temperature dependence [20].

$$-r_A = -\frac{1}{V} \frac{dN_A}{dt} = k C_A^2 = k_0 e^{-E/RT} C_A^2 \quad (2)$$

The energy balance equation is applied for the reaction process and heat transfer.

$$V \rho C_p \frac{dT}{dt} = F \rho C_p (T_i - T_{ref}) - F \rho C_p (T - T_{ref}) - \Delta H_R V (-r_A) - Q \quad (3)$$

For heat transfer rate, Q is represented to eliminate in terms of thermal variables as:

$$Q = F_c \rho_c C_{pc} (T - T_{ci}) \left[1 - \exp \left(-\frac{A_0}{F_c \rho_c C_{pc}} \left(\frac{1}{h_0} + \frac{A_0 F_{cr}^n}{A_i h_{ir} F_{cr}^n} \right)^{-1} \right) \right] \quad (4)$$

Applying the Taylor series approach for linearizing the above non-linear balance equations (1) and (2). The two equations will be expressed in terms of ordinary first order differential equation as:

$$\tau_c \frac{dC_A}{dt} + C_A = \frac{\tau_c}{\tau_R} C_{Ai} + K_{CT} T, C_A(0) = 0 \quad (5)$$

$$\tau_T \frac{dT}{dt} + T = \frac{\tau_T}{\tau_R} T_i + K_{TC} C_A + K_{ht} F_c, T(0) = 0 \quad (6)$$

where

$$\tau_R = \frac{V}{F}, \tau_C = \frac{\tau_R}{1 + 2\tau_R k_r C_{Ar}}, K_{CT} = \frac{E C_{Ar}^2}{RT_r^2} k_r \tau_c, \tau_T = \frac{\tau_R}{1 + \frac{F_{cr} \rho_c C_{pc}}{F \rho C_p} (1 - \beta_r) + \tau_R \frac{\Delta H_r}{\rho C_p} \frac{E}{RT_r^2} k_r C_{Ar}^2} \quad (7)$$

$$\beta_r = \left(\frac{U_{or} A_0}{F_{cr} \rho_c C_{pc}} \right)^{-1}, K_{TC} = -2\tau_T k_r C_{Ar} \frac{\Delta H_r}{\rho C_p}$$

$$K_{ht} = \frac{\tau_T \rho_c C_{pc} (T_r - T_{ci})}{\tau_R F \rho C_p} \left[\frac{\beta_r U_{or} A_0}{F_{cr} \rho_c C_{pc}} \left(1 - \frac{n U_{or} A_0}{A_i h_{ir}} \right) - (1 - \beta_r) \right]$$

The Laplace transform of the mole balance equation (5) and energy balance equation (6) is:

$$(\tau_c s + 1) C_A(s) = \frac{\tau_c}{\tau_R} C_{Ai}(s) + K_{CT} T(s) \quad (8)$$

$$(\tau_T s + 1) T(s) = \frac{\tau_T}{\tau_R} T_i(s) + K_{TC} C_A(s) + K_{ht} F_c(s) \quad (9)$$

Combining equations (8) and (9) by taking the standard Reference steady state parameters values of table 1, the transfer function for reactor temperature can be expressed in terms of coolant flow rate (F_c), inlet coolant temperature (T_i) and inlet feed concentration (C_{Ai}) as:

$$T(s) = \frac{(0.0193s + 0.917)T_i(s)}{s^2 + 312s + 0.0635} + \frac{0.00185C_{Ai}(s)}{s^2 + 312s + 0.0635} + \frac{(32.22s + 1534.15)F_c(s)}{s^2 + 312s + 0.0635} \quad (10)$$

The slave PI controller in the secondary control loop is represented with a transfer function as[22]:

$$G_c(s) = K_p \left(1 + \frac{1}{\tau_i s} \right) \quad (11)$$

TABLE 1. List of Standard Reference steady state parameters value [21]

Variable	unit	Value	Variable	unit	Value
ρ	Kg/ m ³	800	F	m ³ /hr	1.13
ρ_c	Kg/ m ³	1000	T _{cr}	⁰ K	331.4
V	m ³	1.36	F _{cr}	m ³ /hr	1.41
C _p	KJ/Kg ⁰ K	3.13	T _r	⁰ K	335
E	KJ/K mol	69815	C _{AR}	K mol/m ³	4.031
R	KJ/K mol ⁰ K	8.314	C _{pc}	KJ/Kg ⁰ K	4.18
(-ΔH _R)	KJ/K mol	69815	T _{ci}	⁰ K	294.7
U _{0r}	KJ/hr m ² ⁰ K	3065	k ₀	s ⁻¹	1100

3.1 Design of secondary loop pi controller

The characteristic equation is obtained for the plant transfer function of reactor temperature, with the manipulated variable of coolant flow rate(Fc) by assuming the reference value of the peak overshoot of less than 5% and settling time of 10s [23]. From equation (10) the plant model in terms of manipulated variable is taken and the gain K will be computed by applying Routh's Stability analysis as:

$$1 + \frac{K(32.22s + 1534.15)}{s^2 + 312s + 0.0635} = 0 \quad (12)$$

$$s^2 + (32.22K + 312)s + 0.0635 + 1534.15K = 0$$

From the given transient parameters, the natural frequency (ω_n) and the damping ratio (ξ) become 0.614 rad/s and 0.652 respectively.

Using the given transient parameters value and equation (7), the gain K becomes:

$$K = 0.0242 \quad (13)$$

Using the Ziegler Nichols tuning topology, the critical period and gain can easily be obtained as [14]:

$$k_{cr} = K = 0.0242, K_p = 0.45 * k_{cr} = 0.011, \quad (14)$$

$$P_{cr} = \frac{2\pi}{\omega_n} = 10.23, \tau_i = \frac{P_{cr}}{1.2} = 8.525$$

Substituting all the above values of parameters, equation (11) become:

$$G_c(s) = 0.011 \left(1 + \frac{1}{8.525s} \right) \quad (15)$$

The closed loop transfer function for the entire secondary control loop can be written as:

$$G(s) = \frac{G_p(s)G_c(s)}{1 + G_p(s)G_c(s)} \quad (16)$$

where

$$G_p(s) = \frac{T(s)}{F_c(s)} = \frac{32.22s + 1534.15}{s^2 + 312s + 0.0635} \quad (17)$$

Thus, the closed loop transfer function of the secondary control loop become:

$$G(s) = \frac{0.7269s^2 + 32.24s + 230.2}{s^3 + 6.52s^2 + 114.4s + 80.88} \quad (18)$$

Representing the z transform equivalent for the temperature plant in equation (10), bilinear transformation technique is applied to obtain G(z) as:

$$G(z) = \frac{y}{u} = \frac{0.198z^2 - 0.012z - 0.0323}{z^3 - 1.727z^2 + 1.302z - 0.521} \quad (19)$$

Now the reference model should be specified so that it would track the plant output to the steady state level which is expressed as :

$$G_m(z) = \frac{B_m(z)}{A_m(z)} = \frac{b_{m0}z^2 + b_{m1}z + b_{m2}}{z^3 + a_{m1}z^2 + a_{m2}z + a_{m3}} \quad (20)$$

The parameter β is chosen between 0 and 1 so that it will expressed in terms of the ratio between $A_m(1)$ and $B(1)$ as:

$$\beta = \frac{A_m(1)}{B(1)} = 0.789 \quad (21)$$

The numerator polynomial $B_m(z)$ can be written as:

$$B_m(z) = 0.798(0.198z^2 - 0.012z - 0.0323) \quad (22)$$

$$= 0.158z^2 - 0.00943z - 0.0253$$

The denominator polynomial $A_m(z)$ is expressed as:

$$A_m(z) = z^3 + a_{m1}z^2 + a_{m2}z + a_{m3} \quad (23)$$

From equation (13), the denominator polynomial $A_m(1)$ is represented as

$$A_m(1) = a_{m1} + a_{m2} + a_{m3} \quad (24)$$

$$= \beta B(1) = 0.123$$

Since the the denominator polynomial $A(z)$ of the plant model $G(z)$ has two complex and one real zeroes so that for appropriate model following between both the reference and plant models, there should also be two complex and one real zeroes in the denominator of reference model $G_m(z)$.

$$r_1 = a + jb, r_2 = a - jb, r_3 = c, \quad (25)$$

$$A_m(z) = (z - (a + jb))(z - (a - jb))(z - c)$$

$$= z^3 + (c - 2a)z^2 + (a^2 + b^2 - 2ac)z + (a^2 + b^2)c$$

Choosing arbitrary values between 0 and 1 for the real and imaginary parts of the two roots as:

$$r_1 = 0.5 + j0.4, r_2 = 0.5 - j0.4 \quad (26)$$

Combining equation (24) and (25) with $z=1$

$$1 + (c - 2a) + (a^2 + b^2 - 2ac) + (a^2 + b^2)c = \beta B(1) \quad (27)$$

$$c = 0.7$$

Thus the coefficients of $A_m(z)$ can be computed by combining equation (23) and (25)

$$a_{m1} = (c - 2a) = -1.687$$

$$a_{m2} = (a^2 + b^2 - 2ac) = 1.102 \quad (28)$$

$$a_{m3} = (a^2 + b^2)c = -0.2873$$

Thus, the reference model for the reactor temperature plant will be:

$$G_m(z) = \frac{0.158z^2 - 0.00943z - 0.0253}{z^3 - 1.687z^2 + 1.102z - 0.2873} \quad (29)$$

3.2 Controller design of reactor temperature

The primary controller for output temperature y represented with the pole placement control law algorithm in terms of the desired parameters R , S and T as [24].

$$Ru = Tu_C - Sy \quad (30)$$

Parameters of the plant and its reference model can be related by using Diophantine equation as:

$$AR + B^-S = A_0A_m \quad (31)$$

From the compatibility condition, the degree of polynomials R , S and T will be determined as [25]:

$$\begin{aligned} \deg A_0 &= \deg A - \deg B - 1 \\ \deg R &= \deg A_{C2} - \deg A \\ \deg T &= \deg S \leq \deg R \end{aligned} \quad (32)$$

$$R(z) = z^2 + r_0z + r_1, \quad S(z) = s_0z^2 + s_1z + s_2$$

$$T(z) = t_0z^2 + t_1z + t_2$$

The observer polynomial $A_0(z)$ is chosen as arbitrary polynomial so that its zeroes should be negative and real and should satisfy jury stability criterion given as:

$$A_0(z) = (z + 0.4)(z + 0.8) = z^2 + 1.2z + 0.32 \quad (33)$$

Substituting all the parameters in the Diophantine equation (31) and equating coefficients with equal power of z , we have the following equations:

$$\begin{aligned} r_0 + 0r_1 + 0.198s_0 + 0s_1 + 0s_2 &= -0.687 \\ -1.627r_0 + r_1 - 0.012s_0 + 0.198s_1 + 0s_2 &= -0.425 \\ 1.302r_0 - 1.627r_1 - 0.0323s_0 - 0.012s_1 + 0.198s_2 &= 0.5448 \\ -0.521r_0 + 1.302r_1 + 0s_0 - 0.0323s_1 - 0.012s_2 &= -0.111 \\ 0r_0 - 0.521r_1 + 0s_0 + 0s_1 - 0.0323s_2 &= -0.0459 \end{aligned} \quad (34)$$

Solving the unknowns in (34) simultaneous, the controller parameters become:

$$\begin{aligned} r_0 &= 0.875, \quad r_1 = 0.172, \quad s_0 = 0.872, \\ s_1 &= -0.371, \quad s_2 = 1.231 \\ S(z) &= 0.872z^2 - 0.371z + 1.231, \\ R(z) &= z^2 + 0.875z + 0.172 \quad (35) \\ T(z) &= \beta_1 A_{01}(z) \\ &= 0.798(z^2 + 1.2z + 0.32) \\ &= 0.798z^2 + 0.958z + 0.255 \end{aligned}$$

Control law can be obtained using equation (30) which is computed as shown below

$$\begin{aligned} u(t) &= 1.482u(t-1) - 0.524u(t-2) + 0.798u_c(t) + \\ &+ 0.798u_c(t-1) + 0.128u_c(t-2) \\ &- 1.687y(t) - 0.643y(t-1) + 0.123y(t-2) \end{aligned} \quad (36)$$

The overall transfer function of reactor temperature become:

$$\frac{y}{u_c} = \frac{BF}{AR+BS} = \frac{0.158z^4 + 0.1484z^3 - 0.036z^2 - 0.02731z - 0.004131}{z^5 - 0.687z^4 - 0.425z^3 + 0.5448z^2 - 0.111z - 0.04597} \quad (37)$$

3.3 Parameter estimation

The reactor temperature Process model (19) can also be rewritten as:

$$\begin{aligned} y(t) &= 1.727y(t-1) - 1.302y(t-2) + \\ &+ 0.521y(t-3) + 0.198u(t-1) - \\ &- 0.012u(t-2) - 0.0323u(t-3) \end{aligned} \quad (38)$$

The plant output is represented in terms of estimated parameters and regression vector as:

$$y(t) = \phi^T(t)\theta \quad (39)$$

where

$$\phi^T(t) = [-y(t-1) \quad -y(t-2) \quad -y(t-3) \quad u(t-1) \quad u(t-2) \quad u(t-3)] \quad (40)$$

$$\theta^T = [0.198 \quad -0.012 \quad -0.0323 \quad -1.727 \quad 1.302 \quad -0.521] \quad (41)$$

The new parameter estimate for the desired internal plant model can be obtained as

$$\theta(t) = \theta(t-1) + K(t)\varepsilon(t)a \quad (42)$$

where

$$\varepsilon(t) = y(t) - \phi^T(t)\theta(t-1) \quad (43)$$

$$K(t) = P(t-1)\phi(t)\left(\lambda + \phi^T(t)P(t-1)\phi(t)\right)^{-1} \quad (44)$$

where λ is an exponential forgetting factor for the temperature plant which is chosen between 0 and 1.

The initial covariance matrix $P(t-1)$ is necessarily chosen as large value for further computing the matrices $K(t)$ and $P(t)$ in the estimation algorithm.

4. SIMULINK DESIGN OF THE PROPOSED SYSTEM

Simulink design of figure4 shows the entire control system of stirred tank reactor with controller, plant model with estimator block for temperature outputs. The parameter estimator block uses an RLS estimation algorithm. On the other hand, the controller block uses the pole placement algorithm for estimating coefficients of the controller parameters.

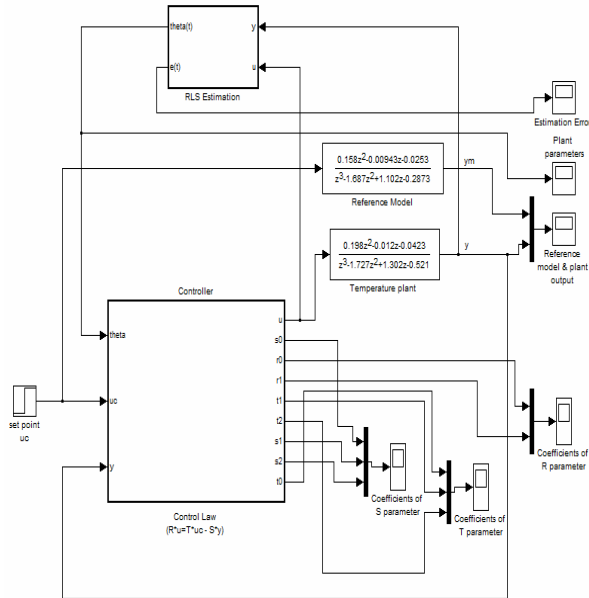


Figure 4. Simulink design of the proposed system

4.1 Recursive least squares estimator

The RLS estimator updates the plant parameters for achieving a stable plant model output. It estimates the desired temperature plant parameters of the reactor system. It updates the parameters a_1, a_2, a_3, b_0, b_1 and b_2 as shown in the figure 5. With the change in the parameter λ , the entire parameter of the system will also be varied.

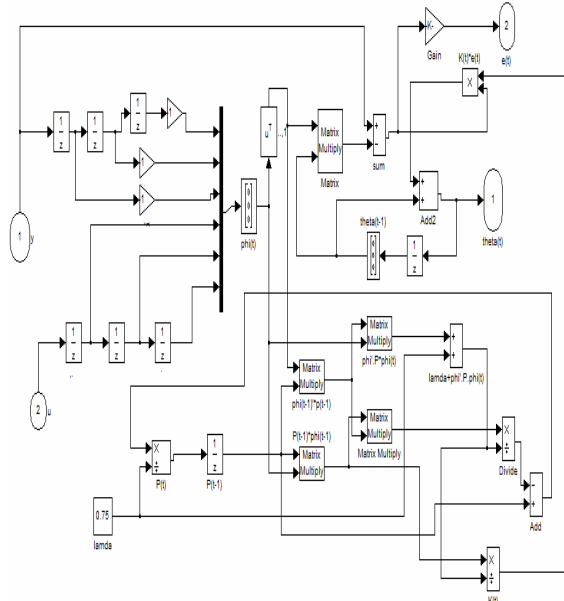


Figure 5. Simulink design of Parameter estimation block for Temperature Plant

4.2 Controller Block

The master self-tuning controller will estimate the coefficients of each controller parameters S, T and R with the corresponding update in the plant parameters. It satisfies the pole placement control law algorithm in order to estimate the controller parameters

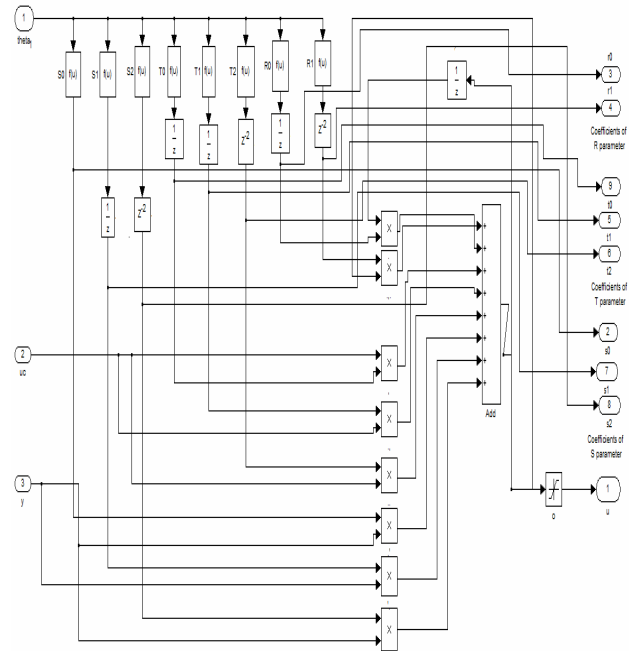


Figure 6. Simulink design of Primary Self-tuning controller of Temperature Plant

5. RESULTS AND DISCUSSION

5.1 Temperature of reactor plant output

In this case the output of temperature plant dynamics followed the reference model as the step reference input to the entire system and the simulation result of the temperature of reactor plant with reference to reference model is shown in figure 7. From the simulation graph, temperature of the plant output without controller exceeds the required steady state temperature value 335k. But in the case of the plant with controller, the plant output approaches to the steady state value.

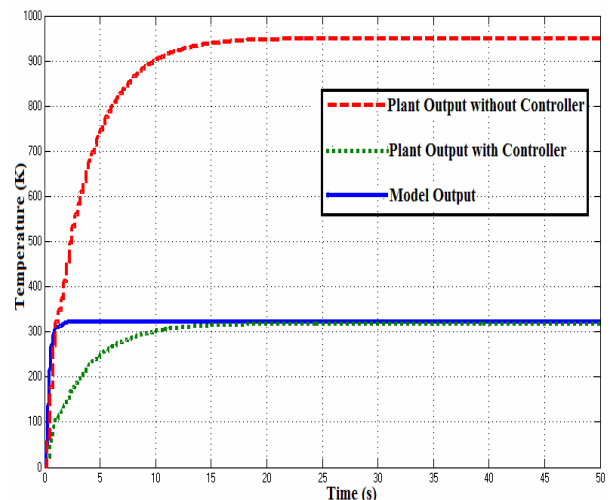


Figure 7. Simulation result of temperature of reactor plant with reference to reference model

Table 2. Outputs of Temperature plant and its controller at different period of time

Time (sec)	Temperature plant output y without controller (K)	Temperature plant output y with controller (K)
t	y(t)	y(t)
0	0	0
2.5	480	260
5	750	250
7.5	800	300
10	900	316
15	950	334

5.2 Self-tuning controller parameters

Figure 8 shows the simulation results of Temperature Plant Master Controller Parameters, the controller parameters S and R shows slight variation in their coefficient values before 13s compared with the coefficients of the parameter T.

With reference to figure 8 graphs the following data has been collected as shown in the table 2.

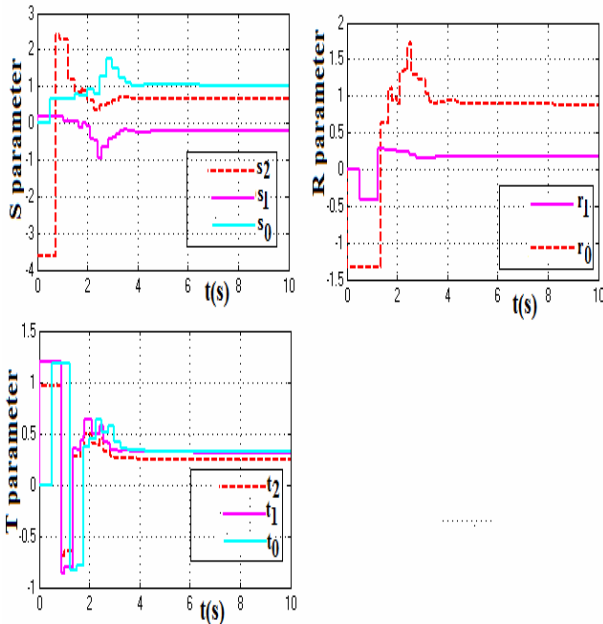


Figure 8. Simulation result of control parameters

Table 3. Controller parameters value at different period of time

Time (sec)	Controller Parameters								
	S parameters			T parameters			R parameters		
	q ₀	q ₁	q ₂	t ₀	t ₁	t ₂	p ₀	p ₁	
0	0	0.15	-3.6	0	1.25	1	-1.45	0	
1.5	0.5	0.2	2.4	1.25	-0.25	1	-1.45	-0.4	
2	1	-0.9	0.85	0.45	0.6	0.4	0.9	0.35	
4	1.05	-0.85	0.75	0.35	0.45	0.3	0.95	0.3	
6	1.2	-0.9	0.8	0.35	0.45	0.3	0.95	0.3	

As it can be seen from figure 8 the value of each controller parameters shows slight variation up to 2s and will be stable beyond such time. This will ensure that the stability of the temperature plant output will be maintained with the stability of both the plant and controller parameters.

As it can be seen from table 3 all controller parameters in the pole placement algorithm tends to be stable after 2s so that the stability of such parameter makes the plant output to be stable.

5.3 Reactor temperature estimation error

As it can be observed in the following simulation diagram of figure 9, the appropriate model following between the output of temperature plant dynamics estimation error become zero for time $t > 5s$.

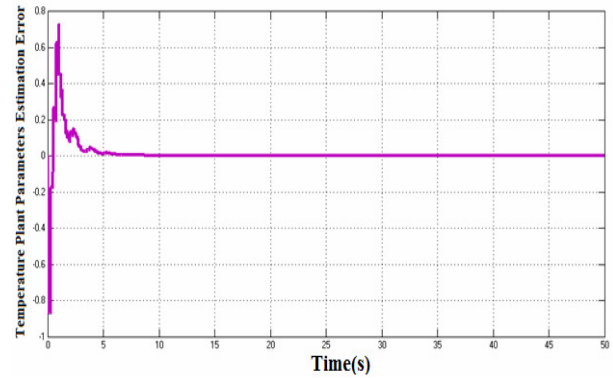


Figure 9. Simulation result for parameter estimation error of temperature plant

With reference to figure. 9, the following data has been collected as shown in the table 4.

5.3 Reactor temperature estimation error

In this section there is also an estimator block that estimate or adjust the plant parameters of the reactor temperature using a recursive least square algorithm the simulation results are obtained as shown in figure 10.

Table 4. Parameters Estimation Error of Temperature plant at different times

Time(s)	Estimation error(ϵ_1)
t	$\epsilon(t)$
0	-0.83
2.5	0.17
5	0
7.5	0
10	0
15	0

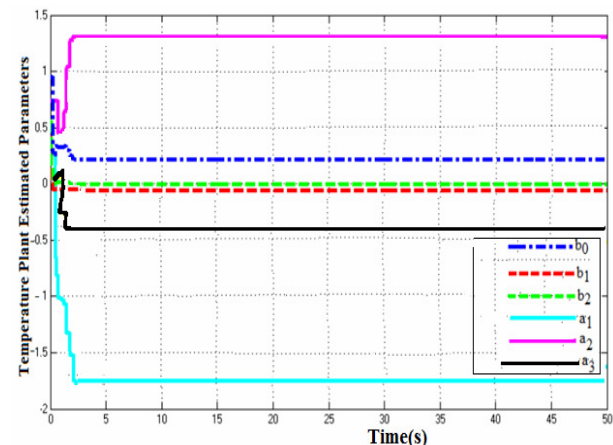


Figure 10. Simulation result for parameters of temperature plant dynamics

Table 5. Comparative Results of Proposed Self-Tuning Regulator based design over the existing controller design

Plant	Peak Overshoot	Settling Time	Rise time	Peak Time
Proposed Self-tuning controller based Temperature plant	0.042	0.22	0.204	0.822
IMC based Temperature plant[21]	0.12	0.512	0.248	0.95
MPC based Temperature plant[21]	0.74	0.621	0.345	0.94
Conventional based Temperature plant [22]	0.08	21	9.85	50

As it can be seen from the simulation graph, until the time $t=2s$ the response of the estimated parameters of temperature plant shows the variation and become stable from this time. Table 5 Enumerates the comparative results of the proposed system with the other design system.

6. CONCLUSIONS

An exothermic chemical reactor allows, releasing of heat energy in the production of desired chemical products from the entire chemical reaction. The feed steams are reacted with the same flow rate and the product of the reactor is withdrawn with its own unique composition. The PI controller in the secondary control loop of the cascade process minimizes the steady state error of its plant so that, the effect of disturbance will be reduced. On the other hand, the controller of the primary control loop will monitor its plant after the secondary plant has been control. An adaptive self-tuning controller is used as a master that allows online and continuous update of the parameters of un-certain reactor temperature and outlet concentration plants. The temperature of the entire reactor system is properly monitored so that its steady state value is kept around 335K. For optimizing the performance of un-certain system, the self-tuning regulator based temperature control approach is an important tool rather than the conventional PID controller based control and other controlling approaches.

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All the authors are contributed equally. The corresponding author is C Bharatiraja, Associate Professor SRM Institute of Science and Technology Department of Electrical and Electronics and Visiting Researcher, Engineering Department of Electrical, University of South Africa, South Africa.

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NOMENCLATURE

CSTR	Continuous stirred tank reactor
PI	Proportional and integral
PID	Proportional, integral and derivative
STR	Self-tuning regulator
MPC	Model predictive control
RLS	Recursive least squares
k₀	Frequency factor
k_r	Arrhenius law
τ_R	Residence time
τ_C	Composition time constant
τ_T	Thermal time constant
K_{CT}	Gain for the effect of temperature on concentration
K_{ht}	Gain of heat transfer
ξ	Damping ratio
C_A	Concentration of reactant A in reactor tank
C_{Ai}	Concentration of reactant A in inlet feed stream
r_A	Rate of reaction for reactant A
Q	Heat energy of reaction
ρ	Product density
ρ_c	Water density
V	Reactor volume
C_p	Thermal capacity
E	Activation energy
R	Perfect gas constant
ΔH_R	Reaction heat
h_i	Inner surface heat transfer barrier
h_{ir}	Reference Inner surface heat transfer barrier
h₀	Outer surface heat transfer barrier
A₀	Outer heat transmission surface
U_{0r}	Heat transmission coefficient
F	Feed flow rate
T	Reactor Temperature
T_{cr}	Reference temperature of cooling jacket
T_{ir}	Reference inlet feed temperature
T_{ci}	Steady state cooling water input temperature
F_{cr}	Reference flow rate of water in cooling jacket
T_r	Reference reactor temperature
T_i	Inlet feed stream temperature
C_{AR}	Reference feed concentration
C_{pc}	Water thermal capacity
β_r	Number of transfer units
θ₁	Parameter of temperature plant
θ₂	Parameter of concentration plant
Φ₁	Regression matrix for temperature plant parameters
Φ₂	Regression matrix for concentration plant parameters
P₁	Covariance Matrix for temperature plant
λ₁	Exponential forgetting factor for temperature plant parameters estimator
λ₂	Exponential forgetting factor for concentration plant parameters estimator
ε₁	Temperature plant parameters estimation error
ε₂	Concentration plant parameters estimation error
P₂	Covariance Matrix for concentration plant

САМОПОДЕСИВИ РЕГУЛАТОР БАЗИРАН НА КАСКАДНОМ УПРАВЉАЊУ ТЕМПЕРАТУРОМ КОД ЕГЗОТЕРМНЕ РЕАКЦИЈЕ МЕШАНОГ СРЕДСТВА У РЕЗЕРВОАРУ РЕАКТОРА

Т.Схога, А.Р.Тхелкар,Ц.Бхаратираја, С.Митику,
Ј.Адедајо

У процесу производње у хемијској индустрији резервоар реактора омогућава ослобађање топлоте код реакције која настаје мешањем супстанци. Ослобађање топлоте егзотермном реакцијом чини систем нелинеарним и неизвесним уколико нема адекватне регулације. У циљу регулације ослобађања топлоте развијен је модел самоподесивог каскадног управљања који је укључио и модел уравнотежења материјала и енергије. Користи се рекурзивни алгоритам најмањих квадрата за онлајн каскадно управљање температуром самоподесивог регулатора и процену параметара контролера као и

излазне температуре у објекту, чија је симулација изведена у присуству и одсуству примарног самоподесивог контролера. Применом PI контролера секундарног кола долази до брзог одзива на брзину протока у расхладном омотачу објекта који каскадно опада код променљивих вредности поремећаја као што су концентрација улазне супстанце и температура расхладне воде. Температура реактора се прати до жељеног нивоа стабилности и постаје стабилна код наредних вредности параметара стабилног стања. Параметри температуре објекта ће такође бити стабилни приликом онлајн процене параметара кола управљачког система. Пројектовање главног контролера и жељеног модела објекта применом алгоритма за подешавање полова је у раду истакнуто. Коришћењем SIMULINK окружења за пројектовање самоподесивог контролера постижу се бољи резултати у поређењу са IMC, MPC и конвенционалним објектом применом контролера температуре објекта.