Journal of
Control and Instrumentation Engineering
e-ISSN: 2582-3000
Volume-7, Issue-1 (January-April, 2021)
www.matjournals.com
https://doi.org/10.46610/JOCIE.2021.v07i01.004

Adaptive MPC Controller based Cascade Control of Distillation Column Parameters Estimation and Optimization

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ABSTRACT

One of the key element of many industrial process plants is a crude oil distillation system. This system requires heat for the vaporization of a mixture of feed to the distillation column in vapor form. In the existing process of distillation, an abnormal change of heat is exhibited due to improper control and monitoring of disturbances affecting the plant. This would result for undesired loss of product and product purity to be reduced. The parameters that are expected to be considered in the analysis, model and control of a distillation system described under the proposed study are inlet feed temperature, distillation column temperature, feed composition, internal liquid and vapor composition, feed flow rate, reboiler temperature and an external reflux temperature to the distillation tower.

Controlling distillation column parameters with Adaptive model predictive control allows for the determination of the predicted future instant values of the plant outputs. Using such control technique, the controlled plant outputs such as the temperature of distillation column, feed preheater, re boiler and the upper reflux is properly controlled and their parameters are estimated using recursive least squares approach in the entire process adaptation mechanism. From the analysis and optimization work made on the proposed system, the efficiency of the plant outputs in tracking their corresponding set point has improved based on the value of the transient system parameters as well as the value of relative volatility of feed mixture to the column. As per the finding in the analysis of the process, 95.4% and 93.5% improvement on the set point tracking and to the amount of evaporation liquid feed has been obtained respectively. On the other hand an improvement on transient parameters has been achieved to all plant outputs. As per the result obtained from the analysis, the peak overshoot, settling time and peak time of the system response has found to be less than 40% including the effect of measured disturbance to the plant. Hence, entire process variable optimization has been performed using the parameters of the model predictive controller to provide the proper degree of stability. Finally the proposed method of study has compared with other control strategies through which the performance of the proposed design has been ensured.

Keywords-- Relative Volitility, MPC Controller, Plant Order Reduction, Optimization, RLS Adaptation Algorithm

INTRODUCTION

Crude oil distillation is a technique used to purify solvents, chemicals, natural products, petroleum, biodiesel, crude oil and other materials in refining process. From such chemical compounds, crude oil is the one that will be processed or refined to produce useable products. The method is exceptionally complex and includes both chemical responses and physical separations. It is composed of thousands of distinctive molecules[1]. Essentially, unrefined oil contains distinctive small-sized solid substances that can be recognized by their weight as well as their bubbling temperature. The division of various components and extraction of the specified fluid is made by considering bubbling focuses of the division of chemical fluids in a distillation tower. The more productive the refining column, the way better would be the

e-ISSN: 2582-3000
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partition of rough oil compounds. With the successful handle of fragmentary refining in acolumn. Unrefined oil would be chemically prepared to alter one fraction into another. At last, divisions of undesired distillates are permitted to be maintained a strategic distance from the refining column as residue [2]. The objectives of the refining framework are to preserve an ideal generation rate and meet determinations that are set for its item. Within the proposed think about the different components that must be controlled in case a refining framework meet its the objective is inspected. For a distillation system the term material balance means some of the materials that leave the distillation column or tower must be equal to the feed that enters the tower. If the flows of materials are not balanced problem could develop in the process. A refining framework fabric adjust can be controlled by controlling the flows of materials. The streams that make up the tower's fabric adjust, are the bolster stream into the tower. The overhead item stream of the overhead framework. The foot item stream of within the tower within the off-gas stream. The off-gas stream is the stream of gasses drawn out of the collector. Control circles comprise of rebellious and gadgets that work together to screen and control the values of prepare factors such as weight, level, and stream. Control circles of a portion of the refining framework is the control system. The flow of materials are controlled by the feed flow control loop, a bottom level control loop, an overhead product level control loop and an off gas stream pressure control loop. The feed flow control loop provides the means to control the weighted which feed flow into the tower. The bottom level control loop controls the level in the bottom of the tower by controlling the bottom product flow. The overhead item level control circle directs the level within the overhead recipient by controlling the overhead product flow. The off-gas stream weight control circle controls the discharge of gas from the overhead receiver which controls the pressure within the refining tower. [3]

Heat Transmission in Distillation Column

For distillation system, the term energy balance is the heat that goes into distillation tower must be equal to the heat goes out of the tower. The energy flows for a distillation tower can be divided into primary energy flow and secondary energy flow as it is shown in Fig.1. The primary energy flow associated with heat transfer into or out of the system. The primary energy flow of the system are the heat input to the reboiler and the feed preheater and the heat that transferred out of the overhead vapor and condenser. Most of the heat input to the tower from the reboiler in the feed preheater is removed by the condenser.

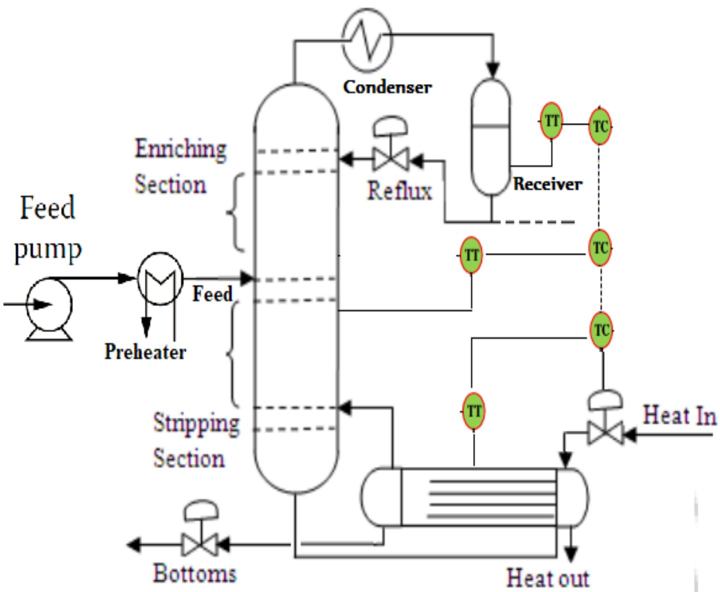

Figure 1: *Crude Oil Distillation Process.* [4]

e-ISSN: 2582-3000
Volume-7, Issue-1 (January-April, 2021)
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Heat also leaves the tower in the products, in the gases that is invented in the system and by heat transfer through equipment piping. Secondary energy flow is associated with heat transfer in which the heat remains within the distillation system. The amount of secondary energy flows is determined by external reflux flow rate. Two major factors in achieving in an energy balances are the weighted which the reboiler vaporizes liquid and the weight of vapor condensation in the condenser. If the system has the feed preheater, the heat of the preheater supply must also be removed to achieve an energy balance. The energy flows is controlled by the energy balance control loop.

Each control loop consists of instrument and devices that work together to monitor and control the value of the process variables. The term steady state operation describes conditions in the distillation system when the process variables are changing in small amount within prescribed limit. During steady state operation changes in the flows of material and energy is minimal under handle by the control system. However, major changes called process disturbances can affect the material balance and energy balance. Many process disturbances results from changes in feed composition or concentration. [5].

Materials Flow in Distillation Column

The liquid feed comming from the preheater block is dicomposed into liquid having a mixture of lighter and heavier componets which are specified with their concentration. If the concentration of lighter component become large, there would be a lower amount of heavier component in the feed that flow down the distillation tower. The decrease in heavier component causes the tower bottom temperature to initially increase. The bottom temperature sensor that is part of the reboiler temperature control loop senses the temperature increase and the control loop automatically decreases the reboiler steam flow to adapt the new state of vapor in the tower. With less heat added by reboiler and since there is no more lighter low boiling vapors the tower top temperature decreases. When the amount of vapor raised in the tower, the condensed liquid level in the overhead receiver will also be raised. The overhead product control loop senses the increase and increases the flow of the overhead product. The change in feed composition also affects the level in the bottom of the tower. Since there are more lighter components in the feed the bottom level decreases, the bottom level control loop senses the decrease and reduces the bottom product flow. A distillation tower's bottom temperature makes change during the course of evaporation or it may be changed to the outer product composition. [6]. The system's control loop reacts to change the bottom temperature to maintain material balance and energy balance. In the entire system the bottom temperature is controlled by adjusting the steam flow to the re boiler which is accomplished by the bottom temperature control loop. If the bottom temperature set point is increased, the bottom temperature control loop respond by increasing steam flow.

Relative Volitility

In a distillation tower the vapor at some pressure is generated through the process of evaporation made by the supply of heat on the feed mixure. At the give temperature of heat the change in vapor pressure due to the lighter and heavier components examine the measure for the relative volitility of these two substances. With an increase in vapor flow, large amount of vapor is condensed in the liquid in the overhead condenser in center of overhead receiver. The variation of level and flow of condensed liquid from the upper liquid receiver would have a further change on the vapor pressure of liquid in the tower. The control loop would adjust the lubber in the parts of forced air from the fan of the condenser to cool the increased amount of vapors. In addition the external reflux control loop senses the temperature increase in the tower and reactesd by increasing the flow of external reflux to the tower. When the external reflux control loop senses the decrease in temperature, it decreases the flow of reflux to the tower. The amount of condensation is varied by changing the temperature of the reflux [7]. An increase in the bottom temperature is an decrease in the liquid level in the bottom of the tower. Since the bottom level has decreased the bottom level control loop decreases the flow of the bottom's product. An increase in the bottom temperature means

e-ISSN: 2582-3000
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there is greater amount of vapor flow. As the additional vapor is condense in the condenser and flow is enter the receiver the temperature of liquid in the receiver increases. The condenser temperature control loop senses the increase in temperature in the receiver[8].

Distillation Column Process Control

Fractional distillation column is a multiple input and Multiple output system in which several parameters are going to be controlled. The system has three process variables control loops such as the reboiler temperature control loop, a feed preheater temperature control loop and a condenser temperature control loop. The reboiler temperature control loop controls the amount of heat that sends to the tower. The feed preheater temperature control loop controls the temperature of the feed by regulating the flow of steam through the preheater [9]. The condenser temperature control loop regulates the amount of cooling that takes place in the condenser.

Cascade Control

A distillation system contains multiple number input and output parameters including disturbances to the plant. Therefore, there would be multiple control loops through which the parameters of the given are monitored. Due to the multiple number of manipulated and controlled variables in the distillation system, the entire control system is expected to have multiple control loops. Thus, plant outputs are expressed in terms of both manipulated and disturbance input variables in cascade configuration [10].

Model Predictive Controller(MPC)

An MPC is an advanced control technique required to control and optimize a system exhibiting uncertain behavior by providing a predicted future values of the plant output. For a given reference signal r(ki) at sample time ki, within a prediction horizon the objective of the predictive control system is to bring the predicted output as close as possible to the set-point signal, where it is assumed that the set point signal remains constant in the optimization window. The objective is then goes to a system design to compute the proper control parameter vector ΔU such that an error between the set-point and the predicted output is minimized [11-12].

LITERATURE REVIEW

The distillation system design and analysis has been studied in vourious research works. The research study described in [13] has presented the control of distillation column parameters using fuzzy inference approach. In this research the parameters expected to be monitored are only the feed flow rate and its temperature in the column. The research work mentioned in [14] described a case study on the implementation of advanced process control for crude distillation unit. In such study the research concern was on the way to control the distillation system process by the techniques of site survey, draft design, acceptance test and trial operation. The study described under [15] dealt with the optimization of crude distillation unit in providing maximum rate of concentration product from the column product by varying of the column temperature. The research study [16] presented the design of general predictive controller for distillation system using order reduction techniques. In this research study, the materials flow control analysis has been modeled and comparatively the performance of the system has determined. In the research work [17] described tuning of PI controller parameters for lab scale distillation column using extended predictive control. In such research the column temperature and overhead liquid flow was considered for estimating the plant parameters. The research work presented in [18] discussed the effects of corrosion on distillation column and its qualitative method of controlling it with the prevention performed on the overhead product flow. The other research work carried on distillation column control was the study described in [19] that dealt with fuzzy logic control technique for monitoring the upper product and vapor concentration control in side a distillation column.

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https://doi.org/10.46610/JOCIE.2021.v07i01.004

METHODOLGY

Proposed System Design

In the proposed system design, the following assuptions has been taken to analze the process parameters.

1. Presence of simple steady state on the vapour and liquid in the distillation tower.
2. The flow rate of liquids through the entire trays of the column is same and time varying.
3. Vapor flow rate on each tray is same and is varied with time.

Process Variables Estimation Algorithm

For the estimation of parameters of the plant a Recursive least squares estimation techniques has been used within the estimator block of the system. From various research work it is esured that an a justifiable claim is given for the least-squares problem on the measured variable to determine the unknown parameters. [20]

Dynamic Modelof Distillation Column

The distillation process control contains manipulated input, disturbance, controlled and uncontrolled variables. The entire contro system has four manipulated variables are flow rate of feed to the distillation column, flow rate of liquid component in the column, flow rate the upper reflux from the overhead receiver and flow rate of steam from the bottom reboiler part to stripping section of the column. The control system also has four controlled outputs from the didtillation column.

The dynamic model for distillation column is developed using material and energy balance of fractional distillation column.[21]

Material and Energy Balance

Material Balance

Material balance for composion of feed to the distillaton column from the preheater section.

$$V\frac{dC_F}{dt} = F_{iF}C_{Ai} - (F_V + F_L)C_F \qquad (1)$$

The material balance for the composition liquid Reflux from the upper receiver:

$$V\frac{dC_R}{dt} = F_L C_{Ri} - (F_R + F_L)C_R \qquad (2)$$

Material balance for composition of liquid inside the destillation colum :

$$V\frac{dC_L}{dt} = F_F C_{Li} + F_L C_L \qquad (3)$$

Material balance for composition of vapor inside the destillation colum :

$$V\frac{dC_V}{dt} = F_V C_V \qquad (4)$$

Energy Balance

The fractional distillation system energy balance is expressed in terms of variables related with heat on that would result for the variation of the column temperature. Such variables are the temperature of feed entering to the tower, the bottom reboiler temperature and the temperature of reflux liquid from the overhead receiver block. The variables are related as it is described in [22].

$$\begin{aligned}\frac{dT_d}{dt} &= \frac{F_v}{V_v}\left(T_{vi} - T_d\right) + \frac{F_L}{V_L}\left(T_{Li} - T_d\right) + \frac{F_F}{V_F}\left(T_{Fi} - T_d\right)\\ &+ \frac{F_R}{V_R}\left(T_{Ri} - T_d\right) + \frac{F_S}{V_S}\left(T_{Si} - T_d\right) - \frac{A_d U_d}{V_d \rho_f C_{pf}}\left(T_d - T_f\right)\\ &- \frac{A_d U_d}{V_R \rho_R C_{Rf}}\left(T_d - T_R\right) - \frac{A_d U_d}{V_r \rho_r C_{pr}}\left(T_d - T_r\right)\end{aligned} \qquad (5)$$

e-ISSN: 2582-3000
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https://doi.org/10.46610/JOCIE.2021.v07i01.004

$$\frac{dT_f}{dt} = \frac{F_v}{V_v}\left(T_{if} - T_f\right) + \frac{A_f U_f}{V_d \rho_f C_{pf}}\left(T_d - T_f\right) \quad (6)$$

$$V_R \rho_c C_{pc} \frac{dT_R}{dt} = F_R \rho_c C_{pc}(T_R - T_{Ri}) - F_c \rho_c C_{pc}(T_d - T_R) \quad (7)$$

$$\frac{dT_r}{dt} = \frac{F_v}{V_v}\left(T_{ir} - T_f\right) + \frac{A_r U_r}{V_d \rho_f C_{pf}}\left(T_d - T_r\right) \quad (8)$$

The heat dissipated from the distillation column due to the stated heat variables as a temperature on the tower is uniform in the space between thermal capacitance of the column wall material negligible and convective thermal resistance of the vapour side inside the column. The losen heat from the column can be mathematically expressed as it has been described as: [23]

$$Q = \frac{T_d - T_{amb}}{1/h_i A_I + \frac{\ln\left(r_{owal} / r_{i\,wal}\right)}{K_p . A_{wall}} + \frac{\ln\left(r_{ins} / r_{owalll}\right)}{K_p . A_{wall}} + 1/h_o A_o} \quad (9)$$

Table 1: *List of steady state values of parameters in distillation system.* [24]

Variable	unit	Value	Variable	unit	Value
ρ	Kg/ m^3	900	C_{fr}	K mol/m^3	4.031
ρ_w	Kg/ m^3	1000	C_{Rr}	KJ/Kg ^{0}K	5.82
Vd	m^3	1.56	C_{vr}	KJ/Kg ^{0}K	5.82
C_p	KJ/Kg ^{0}K	3.13	C_{Lr}	KJ/Kg ^{0}K	5.82
A	KJ/K mol	7900	V_L	m^3	4.11
R	KJ/K mol ^{0}K	8.314	V_V	m^3	3.12
T_{dr}	^{0}K	315			
T_{fr}	^{0}K	290			
T_{Rr}	^{0}K	52			
T_{rr}	^{0}K	285			
F_{Fr}	m^3/hr	1.13			
F_{Lr}	m^3/hr	1.41			
F_{Vr}	m^3/hr	1.61			
F_{Sr}	m^3/hr	1.72			
α		1.5			

Applying the Taylor series linearizing approach for the above non-linear material and energy balance model equations by take the steady state value of the parameters in Table 1, The state space model representation of the entire system dynamic become:

e-ISSN: 2582-3000
Volume-7, Issue-1 (January-April, 2021)

www.matjournals.com https://doi.org/10.46610/JOCIE.2021.v07i01.004

$$\begin{bmatrix} \dot{T}_f \\ \dot{T}_d \\ \dot{T}_R \\ \dot{T}_r \\ \dot{C}_F \\ \dot{C}_R \\ \dot{C}_L \\ \dot{C}_V \end{bmatrix} = \begin{bmatrix} -4.32 & 4.32 & 0 & 0 & 0 & 0 & 0 & 0 \\ 4.32 & -5.321 & 3.65 & 6.32 & 0 & 0 & 0 & 0 \\ 0 & 3.65 & -3.65 & 0 & 0 & 0 & 0 & 0 \\ 0 & 6.32 & 0 & -6.32 & 0 & 0 & 0 & 0 \\ 0 & 0 & 0 & 0 & 11.72 & 13.8 & 0 & 0 \\ 0 & 0 & 0 & 0 & 0 & 7.321 & 0 & 0 \\ 0 & 0 & 0 & 0 & 0 & 0 & 7.61 & 0 \\ 0 & 0 & 0 & 0 & 0 & 0 & 0 & 9.4 \end{bmatrix} \begin{bmatrix} T_f \\ T_d \\ T_R \\ T_r \\ C_F \\ C_R \\ C_L \\ C_V \end{bmatrix}$$

$$+ \begin{bmatrix} 0 & 0 & 43.6 & 0 \\ 14.3 & 10.8 & 4.32 & 7.36 \\ 0 & 0 & 0 & 8.32 \\ 0 & 0 & 0 & 0 \\ 17.3 & 5.16 & 0 & 8.4 \\ 0 & 0 & 0 & 4.83 \\ 5.2 & 0 & 0 & 0 \\ 0 & 0 & 0 & 0 \end{bmatrix} \begin{bmatrix} F_L \\ F_F \\ F_S \\ F_R \end{bmatrix} + \begin{bmatrix} 0 \\ 2.74 \\ 0 \\ 0 \\ 3.41 \\ 0 \\ 4.63 \end{bmatrix} [F_V] \tag{10}$$

From the above state space model the transfer function of the system variables have a characteristic polynomial having a degree of four.

$$\frac{T_f(s)}{F_L(s)} = \frac{88.04s^2 + 1045s + 3050}{s^4 + 28.51s^3 + 177.9s^2 - 137.6s - 2319} \tag{11}$$

$$\frac{T_d(s)}{F_L(s)} = \frac{12.4s^3 + 235.2s^2 + 1475s + 3050}{s^4 + 28.51s^3 + 177.9s^2 - 137.6s - 2319} \quad \frac{T_R(s)}{F_L(s)} = \frac{64.11s^2 + 884.7s + 3050}{s^4 + 28.51s^3 + 177.9s^2 - 137.6s - 2319}$$

$$\frac{T_r(s)}{F_L(s)} = \frac{30.82s^2 + 378.2s + 1131}{s^4 + 28.51s^3 + 177.9s^2 - 137.6s - 2319} \quad \frac{T_f(s)}{F_F(s)} = \frac{32.66s^2 + 387.7s + 1131}{s^4 + 28.51s^3 + 177.9s^2 - 137.6s - 2319}$$

$$\frac{T_f(s)}{F_F(s)} = \frac{32.66s^2 + 387.7s + 1131}{s^4 + 28.51s^3 + 177.9s^2 - 137.6s - 2319} \quad \frac{T_d(s)}{F_F(s)} = \frac{4.6s^3 + 87.26s^2 + 547s + 1131}{s^4 + 28.51s^3 + 177.9s^2 - 137.6s - 2319}$$

$$\frac{T_R(s)}{F_F(s)} = \frac{59.97s^2 + 1074s + 4503}{s^4 + 28.51s^3 + 177.9s^2 - 137.6s - 2319} \quad \frac{T_r(s)}{F_F(s)} = \frac{77.72s^2 + 1273s + 4503}{s^4 + 28.51s^3 + 177.9s^2 - 137.6s - 2319}$$

$$\frac{T_f(s)}{F_s(s)} = \frac{65.11s^2 + 876s + 2789}{s^4 + 28.51s^3 + 177.9s^2 - 137.6s - 2319}$$

e-ISSN: 2582-3000
Volume-7, Issue-1 (January-April, 2021)
https://doi.org/10.46610/JOCIE.2021.v07i01.004

$$\frac{T_d(s)}{F_s(s)} = \frac{9.17s^3 + 188.5s^2 + 1269s + 2789}{s^4 + 28.51s^3 + 177.9s^2 - 137.6s - 2319} \quad \frac{T_R(s)}{F_s(s)} = \frac{47.41s^2 + 729.4s + 2789}{s^4 + 28.51s^3 + 177.9s^2 - 137.6s - 2319}$$

$$\frac{T_f(s)}{F_R(s)} = \frac{6.71s^3 + 226s^2 + 1489s + 2311}{s^4 + 28.51s^3 + 177.9s^2 - 137.6s - 2319} \quad \frac{T_d(s)}{F_R(s)} = \frac{11.6s^3 + 267.7s^2 + 1945s + 403}{s^4 + 28.51s^3 + 177.9s^2 - 137.6s - 2319}$$

$$\frac{T_R(s)}{F_R(s)} = \frac{1.4s^3 + 70.62s^2 + 641.2s + 1521}{s^4 + 28.51s^3 + 177.9s^2 - 137.6s - 2319}$$

$$\frac{T_r(s)}{F_R(s)} = \frac{49.18s^2 + 651.9s + 2149}{s^4 + 28.51s^3 + 177.9s^2 - 137.6s - 2319}$$

Order Reduction

The entire system transfer function has the order of four in its characteristic equation. By using **Skogestad's half rule** approximation technique, the second order transferfunction equivalent is represented in terms of a function having time delay parameter. [25]

$$Q(s) = \frac{Ke^{-Ts}}{(\tau_1 s + 1)(\tau_2 s + 1)} \tag{12}$$

The entire system plant transfer function is made to be represented simplified form as:

$$G(s) = \frac{(b_0 s + 1)(b_1 s + 1)\ldots(b_m s + 1)}{(a_0 s + 1)(a_1 s + 1)\ldots(a_n s + 1)} \tag{13}$$

As per the rule τ_1 chosen to be the largest dominant time constant to be retained. Hence, τ_1 and T holds the maximum value from all values of a_n and b_m in the numerator and denominator of G(s) respectively. While τ_2 woul chosec as

$$\tau_2 = Second \;\; Max\,(a_n) + \sum_{i=1}^{n-1} a_i \tag{14}$$

$$T = Second \;\; Max\,(b_m) + \sum_{i=1}^{m-1} b_i \tag{15}$$

e-ISSN: 2582-3000
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Table 2: *List of approximated plant transfer functions.*

The approximated Transfer functions	
$\frac{T_f(s)}{F_L(s)} = \frac{(0.226s+0.036)e^{-3.34s}}{s^2+0.442s+0.036}$	$\frac{T_f(s)}{F_F(s)} = \frac{(0.27s+0.027)e^{-1.63s}}{s^2+0.406s+0.027}$
$\frac{T_d(s)}{F_L(s)} = \frac{(0.17s+0.044)e^{-1.52s}}{s^2+0.422s+0.044}$	$\frac{T_d(s)}{F_F(s)} = \frac{(0.137s+0.022)e^{-2.13s}}{s^2+0.291s+0.022}$
$\frac{T_R(s)}{F_L(s)} = \frac{(0.145s+0.0504)e^{-2.36s}}{s^2+0.4486s+0.0504}$	$\frac{T_R(s)}{F_F(s)} = \frac{(0.176s+0.0189)e^{-3.54s}}{s^2+0.288s+0.0189}$
$\frac{T_r(s)}{F_L(s)} = \frac{(0.185s+0.018)e^{-4.033s}}{s^2+0.272s+0.018}$	$\frac{T_r(s)}{F_F(s)} = \frac{(0.1924s+0.046)e^{-5.19s}}{s^2+0.695s+0.046}$
$\frac{T_f(s)}{F_s(s)} = \frac{(0.058s+0.028)e^{-3.72s}}{s^2+0.403s+0.0276}$	$\frac{T_f(s)}{F_R(s)} = \frac{(0.2688s+0.027)e^{-6.11s}}{s^2+0.4059s+0.027}$
$\frac{T_d(s)}{F_s(s)} = \frac{(0.141s+0.0378)e^{-4.21s}}{s^2+0.389s+0.0378}$	$\frac{T_d(s)}{F_R(s)} = \frac{(0.103s+0.0249)e^{-5.32s}}{s^2+0.329s+0.0249}$
$\frac{T_R(s)}{F_s(s)} = \frac{(0.155s+0.023)e^{-2.04s}}{s^2+0.32s+0.023}$	$\frac{T_R(s)}{F_R(s)} = \frac{(0.1566s+0.019)e^{-2.65s}}{s^2+0.289s+0.019}$
$\frac{T_r(s)}{F_s(s)} = \frac{(0.146s+0.0289)e^{-1.35s}}{s^2+0.43s+0.0289}$	$\frac{T_r(s)}{F_R(s)} = \frac{(0.176s+0.036)e^{-4.03s}}{s^2+0.3923s+0.036}$

For the proper design of an MPC controller, the continuous time transfer functions of a proposed system listed under Table 2 should be converted into discrete time in terms of z transfor. The Bilinear

e-ISSN: 2582-3000
Volume-7, Issue-1 (January-April, 2021)

https://doi.org/10.46610/JOCIE.2021.v07i01.004

transform method has been chosed by the proposed system analysis to convert continuous-time transfer functions into discrete form as show below: [26]

$$G(z) = G\left(\frac{2z-1}{Tz+1}\right) \qquad (16)$$

where T is the time delay value in the exponential part of a transfer function.

Predictive Control System Optimization

The plant of entire distillation system which has explained before have four outputs and four states. Using the general formulation of predictive control

problem analsis, the proposed system has described by the following discrete state space equation [27].

$$x_m(k+1) = A_m x_m(k) + B_m u(k) + B_d w(k)$$
$$y(k) = C_m x_m(k), \qquad (17)$$

where

w(k) is an input measured disturbance or integrated white noise so it is based on the state-space model (A, B, C). Hence, the future state variables and the predicted output variables can be calculated sequentially in terms of the MPC controller parameters as:

$$x(k_i + N_p \mid k_i) = A^{Np}x(k_i) + A^{(Np-1)}B\Delta u(k_i) + A^{(Np-2)}B\Delta u(k_i+1) + \ldots + AN_p{-}N_cB\Delta u(k_i + N_c - 1). \qquad (15)$$

$$y(k_i + N_p \mid k_i) = CA^{Np}x(k_i) + CA^{(Np-1)}B\Delta u(k_i) + CA^{(Np-2)}B\Delta u(k_i+1) + \ldots + CA^{(Np-Nc)}B\Delta u(k_i + N_c - 1) \qquad (18)$$

where N_c is the control horizon dictating the number of parameters used to capture the future control trajectory while N_p is the prediction horizon that represent the length of the optimization window. Since all predicted variables are determined using the present state of the variable information $x(k_i)$ and the future control signal trajectory $\Delta u(k_i + j)$,

where $j = 0, 1, \ldots N_c - 1$.

Adaptation Mechanism

It is well known that, most industrial process have non-linear and time varying characteristics.Therefore, modelling of it the most challegic work especially when the process operating conditions frequently vary. In most cases control strategies with constant parameters cannot adapt to changes in the operating conditions. The non-linear and time varying property of industrial distillation column is the main motivating factor for the development of adaptive control technique. The plant output and its parameters can be described as [28].

$$y(k) = \theta^T(k)\phi(k) \qquad (19)$$

But the proposed system has four plant outputs with respect to the four manipulated variables. Hence,

$$\theta^T(k) = [\theta_1(k),\ \theta_2(k),\ \theta_3(k),\ \theta_4(k)] \qquad (20)$$

where

$$\theta_1(k) = [a(k),\ b(k),\ c(k),\ d(k)]^T$$
$$\theta_2(k) = [e(k),\ f(k),\ g(k),\ h(k)]^T \quad \theta_3(k) = [i(k),\ j(k),\ k(k),\ l(k)]^T \qquad (21)$$
$$\theta_4(k) = [m(k),\ n(k),\ p(k),\ q(k)]^T$$

The regression vector relating the four plant output is given by:

$$\phi(k) = [\phi_1(k),\ \phi_2(k),\ \phi_3(k),\ \phi_4(k)]^T \qquad (22)$$

Where,

$$\phi_1(k) = [y_1(k-1)\ldots y_1(k-n), u(k-1)\ldots u(k-1-m),$$
$$w(k-1)\ldots w(k-1-p)]^T$$

e-ISSN: 2582-3000
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$$\begin{aligned}\phi_2(k) = [&y_2(k-1)\ldots y_2(k-n), u(k-1)\ldots u(k-1-m), \\ &w(k-1)\ldots w(k-1-p)]^T \\ \phi_3(k) = [&y_3(k-1)\ldots y_3(k-n), u(k-1)\ldots u(k-1-m), \\ &w(k-1)\ldots w(k-1-p)]^T \\ \phi_4(k) = [&y_4(k-1)\ldots y_4(k-n), u(k-1)\ldots u(k-1-m), \\ &w(k-1)\ldots w(k-1-p)]^T\end{aligned} \quad (23)$$

θ and ϕ are the vector of adaptive parameters for the inputs and outputs vector respectively. The prior estimation error:

$$e(k/k-1) = y(k) - \theta(k-1)^T \phi(k) \quad (24)$$

With the new parameter update to the existing plant model can be made by using the present state values the parameters which is given as:

$$\psi(k) = \theta(k) + K(k)e(k/k-1) \quad (25)$$

where, $e(k/k-1) = y(k) - \phi^T(k)\theta(k)$

$$K(k) = P(k-1)\varphi(k)\left(\lambda + \phi^T(k)P(k-1)\phi(k)\right)^{-1} \quad (26)$$

where λ is exponential forgetting factor for parameter adjustment block in the parameter estimator section.

RESULT AND DISCUSSION

The Simulink result of the entire distillation process is shown on the Fig. 2. The entire control system of the plant used an MPC controller with an adaptive parameter estimator block for online estimation of the parameters of the plant having four manipulated and controlled output variables. The parameter estimation block provides an estimate of the plant parameters in accordance with the four controlled outputs.

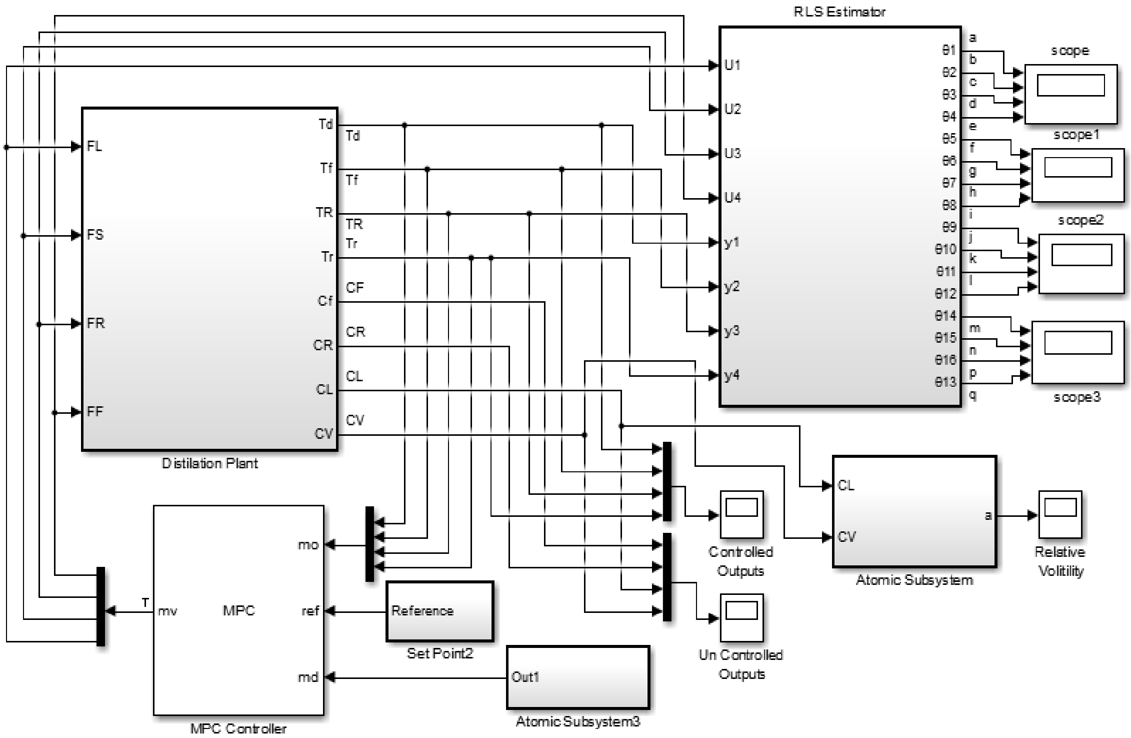

Figure 2: *Simulink design of the proposed system.*

e-ISSN: 2582-3000
Volume-7, Issue-1 (January-April, 2021)

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https://doi.org/10.46610/JOCIE.2021.v07i01.004

With an appropriate value exponential forgetting factor at 0.95 the estimator part would update plant parameters using recursive list square algorithm.With an MPC controller block the future behavior of the plant is predicted to obtain the correspondin stable outputs. Hence, the MPC controlled has been designed with its parameters such as prediction horizon of N_p = 4, control horizon of N_c = 2 and constraints of W=0.8 were selected depending on the number of parameters existed under the control signal U. The change in controlled signal U exprssed with the four maniplated mariables in the entire destillation process such as feed flow rate, liquid floow rate, reboiler steam flow rate and reflux flow rate. The relative volitility block of the system provides an information on the rate of liquid compent is going to convert into vapor form depending on concentration of liquid and vapor inside the destillation column.

Temperature Outputs of Distillation Column

In the proposed system the controllled variable are the temperature of reflux, reboiler, feed and distillation column. Since the reference value of for these have been obtained as 315^0K, 289^0K, 55^0K and 295^0K respectively. Thus by the implementation of adaptive MPC controller the four controlled plant outputs have tracked their set poit properly as shown on Fig 3.

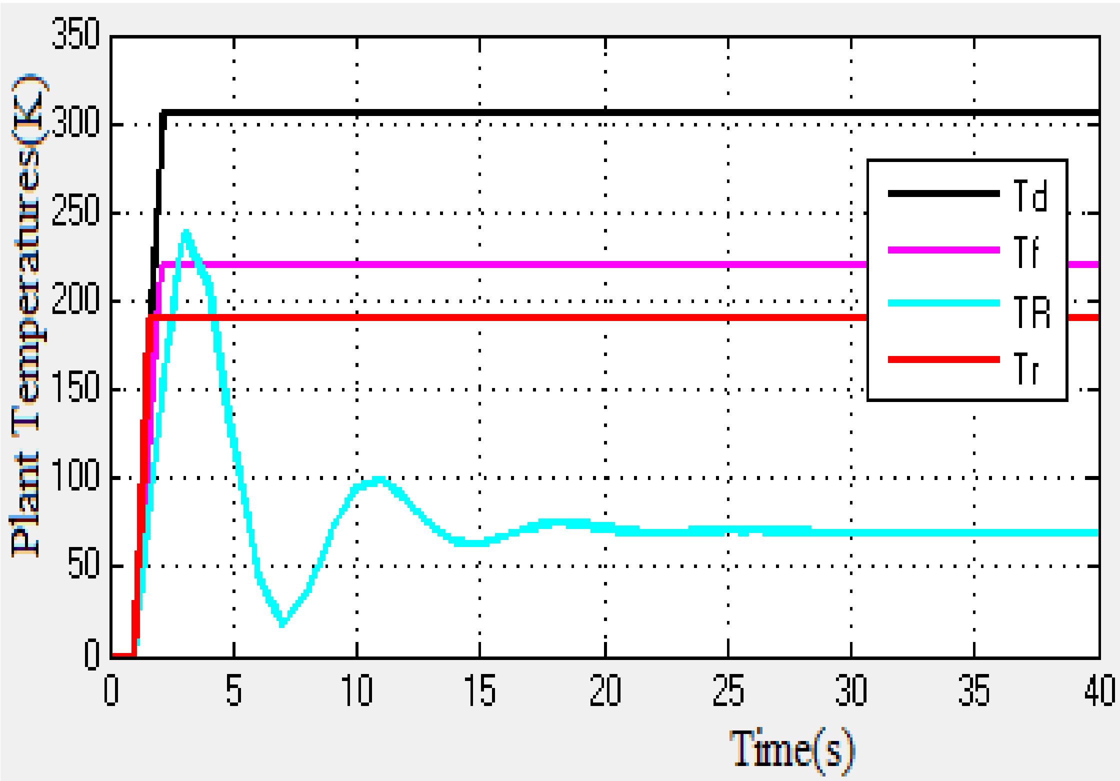

Figure 3: *Simulation result of temperature component of Distillation system.*

Distillation Column Estimated Parameters

In this section of the result, an estimator block provides an estimate of the plant parameters of the entire system using a recursive least square algorithm. The proposed design have estimated the parameters of distillation column plant with respect to the four manipulated and controlled variables as shown from Fig. 4-7. Therefore, the parameters which were described at equation (21) are going to be estimated properly for the respective plant outputs with respect to the four manipulated input variables.

e-ISSN: 2582-3000
Volume-7, Issue-1 (January-April, 2021)

https://doi.org/10.46610/JOCIE.2021.v07i01.004

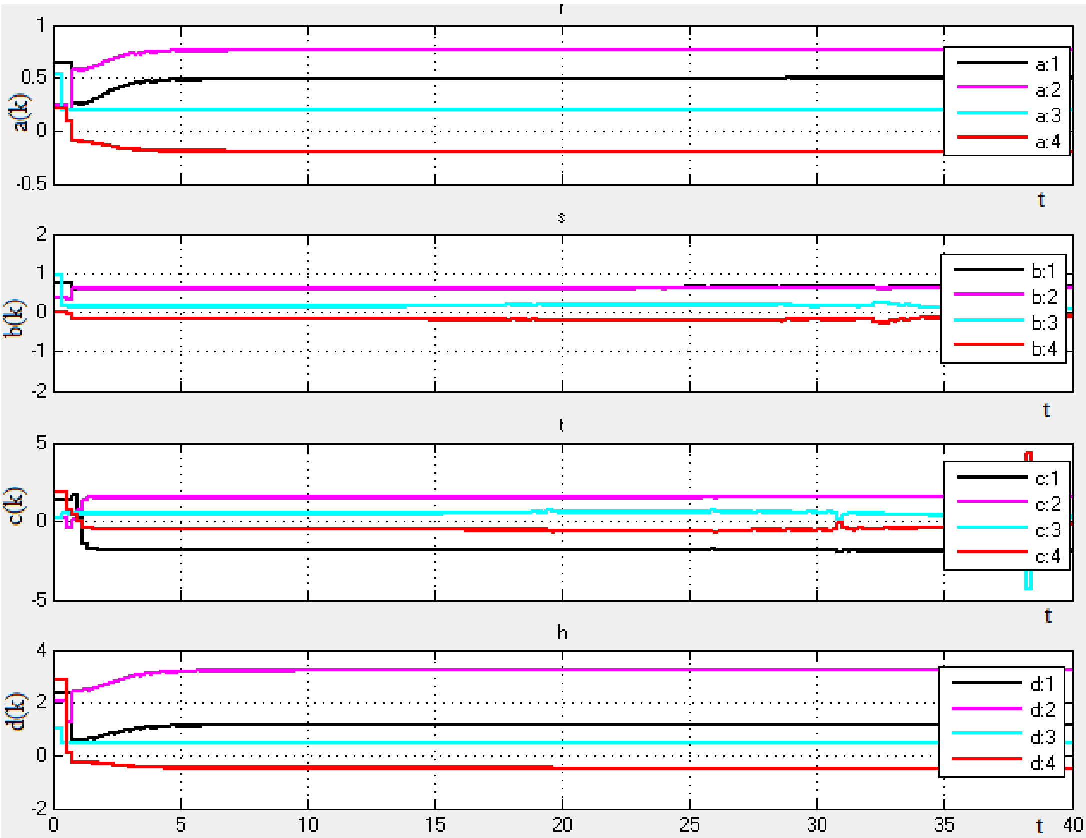

Figure 4: *Distillation Column Temerature Estimated Parameters*

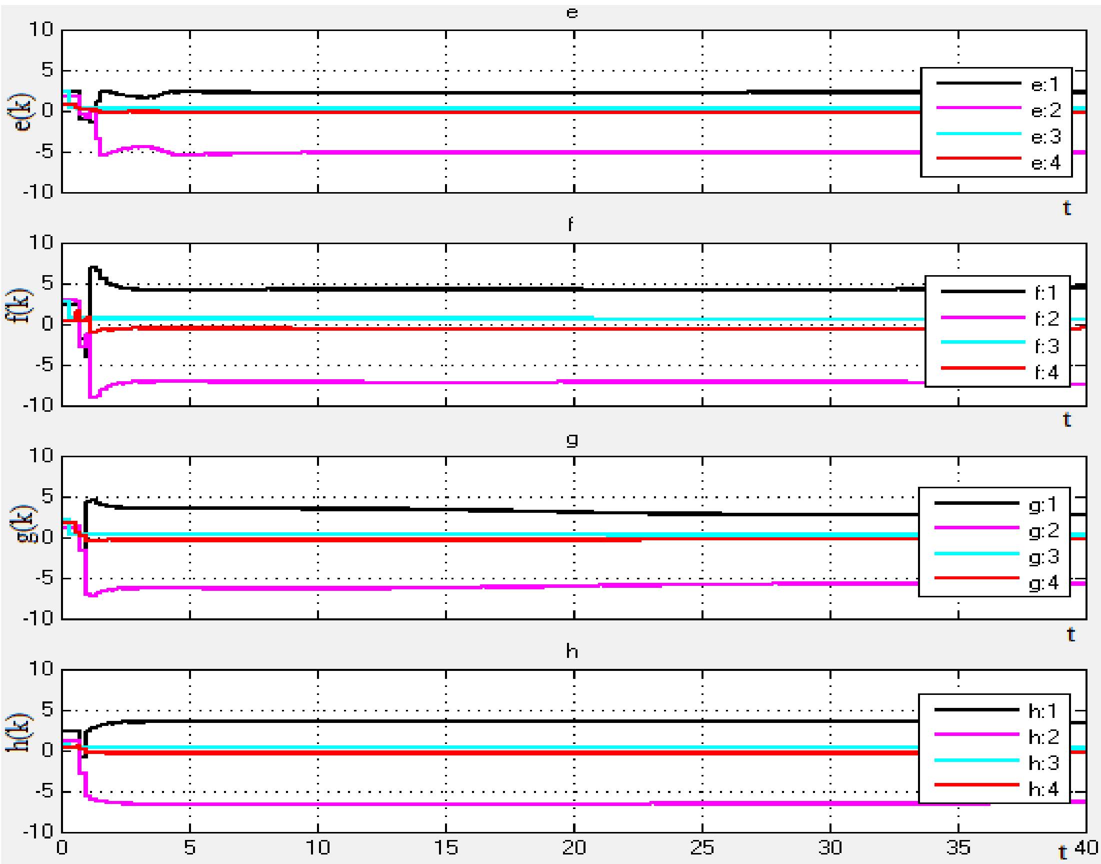

Figure 5: *Feed Temerature Estimated Parameters.*

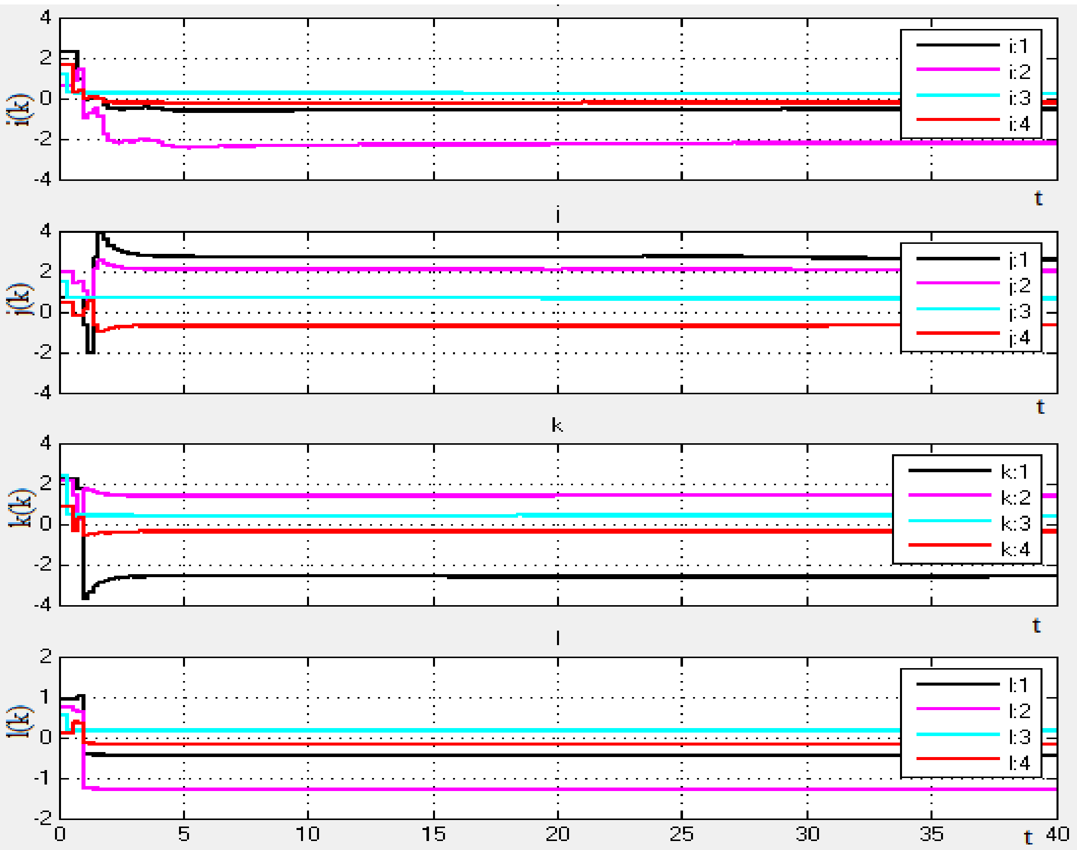

Figure 6: *Reflux Temerature Estimated Parameters.*

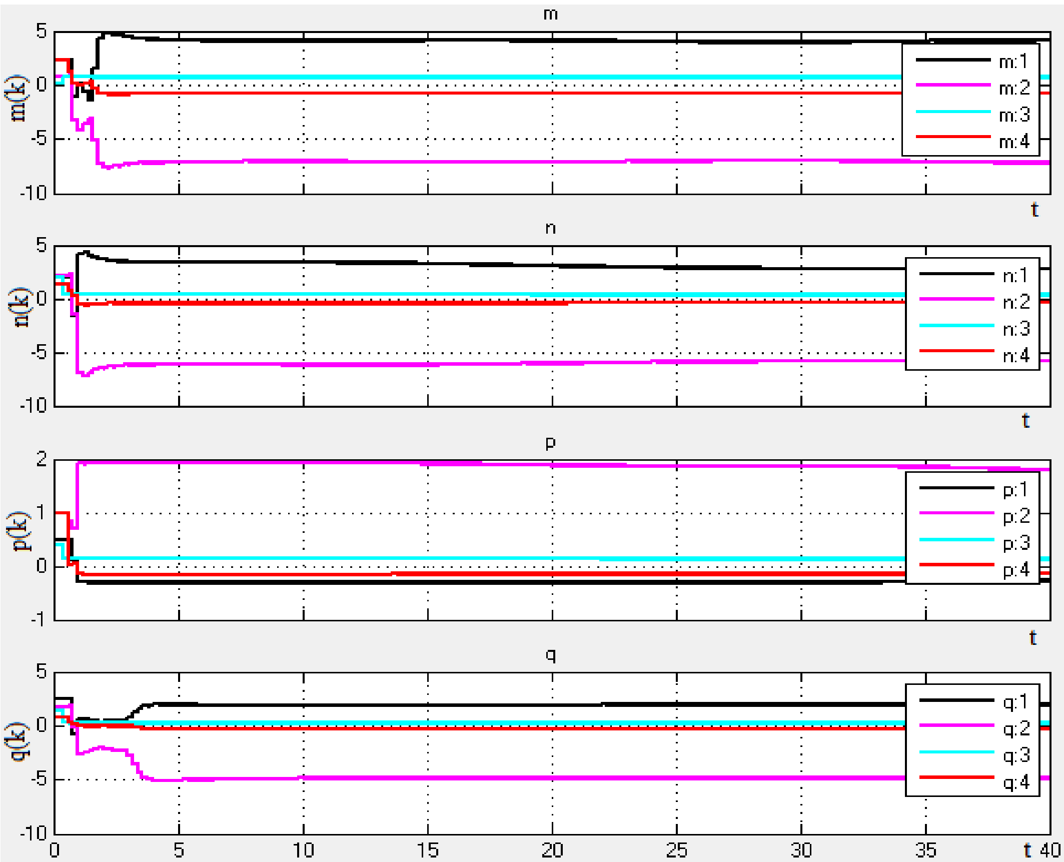

Figure 7: *Reboiler Temerature Estimated Parameters.*

Controlled Output of Relative Volitility

The relative valitility of the materials flowing in the tower has been determined by the effict of temperature on the vapor and liquid composition. As it can be see from Fig. 8 up to a time of 7 seconds relative volitility is tend to vary with time but beyod this period it would follow a constant or stable operational trajector in providing refernce value of 1.5 for the entire process.

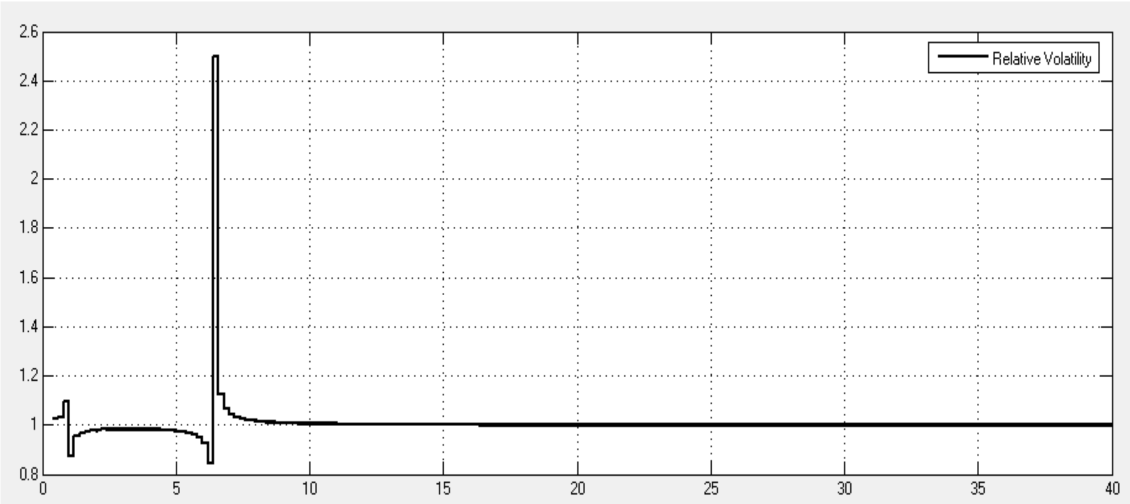

Figure 8: *Relative volitility of Liquid in distillation column.*

Table 3: *Comparative Results of Proposed system design with the existing control techniques.*

Control Tecnique	Plant	Peak Overshoot	Settling Time	Rise time	Peak Time
With PID type controller	T_d	0.957	0.82	0.752	0.691
	T_f	0.852	0.951	0.631	0.734
	T_d	0.634	0.634	0.712	0.684
	T_R	0.744	0.502	0.824	0.773
With IMC type controller	T_d	0.417	0.821	0.673	0.547
	T_f	0.651	0.551	0.634	0.634
	T_d	0.534	0.634	0.514	0.514
	T_R	0.743	0.611	0.677	0.707
With proposed MPC type controller	T_d	0.336	0.325	0.361	0.351
	T_f	0.381	0.302	0.353	0.307
	T_d	0.345	0.264	0.322	0.314
	T_R	0.273	0.342	0.327	0.373

Using the proposed system controller, the stability of the entire distillation column process become more better than the other control strategies as shown in Table 3.

CONCLUSION

The design of Adaptive model predictive control technique for the MIMO distillation system control is modelled to optimize the process having a set of constraints due to the dynamics behavior of the plant.

The constraint on the plant is due to the presence of uncertain distillation process variables. The process variables in the entire system are frequently needs proper design and optimization to obtain stable controlled plant output. Hence, with the imlementation an advanced adaptive MPC base model analysis, the uncertain parameters of the plan have been properly controlled optimized. The MPC controller used in the proposed study provides an estimate of future values of the plant outputs using the ordinary model of the plant. The future predicted values of controlled plant variables have been determined properly and they have tracked their set ponts effectively with the minimum appropriate values of transient parameters.

ACKNOWLEDGEMENT

This work was financially supported by Jimma Institute Technology, Jimma University Ethiopia for the research scholars.

CONFLICT OF INTEREST

The authors declare that there is no conflict of interest regarding the publication of this paper.

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NOMENCLATURE

MPC: Model predictive controller
IMC: Internal Model Control
PID: Proportional, integral and derivative
RLS: Recursive least squares
$C_{V:}$ Vapor composition in a tower
$C_{L:}$ Liquid composition in a tower
$C_{Rr:}$ Reference reflux liquid Composition
Q: The disspated heat
ρ: Density of feed
$\rho_{c:}$ Water density
V_d: Volume of distillation tower
V_f: Volume of feed
V_R: Volume of reflux liquid
$C_{p:}$ Thermal capacity
A: Activation energy for decomposition of feed
F : Feed flow rate
T_d: Distillation Column Temperature
$T_{iR:}$ Reference temperature value of reflux
$T_{ri:}$ Steady state temperature of reboiler
$T_{dr:}$ steady state value Distillation temperature
$N_{p:}$ prediction horizon
$N_{c:}$ control horizon
W : Constraints
θ: Estimated plant Parameter
ϕ : The Regression matrix
λ: Exponential forgetting factor plant outputs
P: Covariance Matrix
e: Plant parameters estimation error