International Journal of Scientific & Engineering Research, Volume 7, Issue 10, October-2016
ISSN 2229-5518

Petroleum Distillation Column Electronic Internal Temperature Controller

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Abstract—Petroleum fractional distillation columns are highly complex and non-linear, with variables changing at different rates. Efficient control of the product of the fractional distillation columns is one of the challenges faced by engineers in the distillation of petroleum raw feed in most refineries. An extreme and frustrating case of this occurs during rainfall and at night when the fractional distillation column external temperature drops, greatly affecting the column internal temperature. The variation in the internal temperature of a distillation column is the most difficult parameter to control in a fractional distillation column system as it exhibits high process and measurement lag and can also be ambivalent as a measure of composition. Consequently, it is difficult to maintain the stability and purity of the distillation column system products in relation to the internal temperature of the column. This research is aimed at designing an Electronic Internal Temperature Controller for a Petroleum Fractional Distillation Column. The material balance and energy balance models have been used to achieve the transfer function and state equation of the system. A suitable fine-tuned PID controller has been designed and integrated to the system. Simulation results obtained based on Root Locus and Step Response show a significant improvement on the performance of the system.

Index Terms— Distillation column energy balance, Electronic Internal Temperature Controller, Petroleum distillation, PID controller

1 Introduction

Distillation is the most commonly used separation technology employed in petroleum and chemical industries for the separation of two or more components from a homogeneous fluid mixture. It is not only use for separation but also used for enhancing mass transfer and transferring heat energy [2]. For instance, in a petroleum distillation process, different products are separated by using the phenomenon of different boiling points/temperatures of different components of the crude oil feed. The lighter products are collected from the top and the heavier products are collected from the bottom of the column.

In other words, distillation is an energy-separating-agent equilibrium process that uses the difference in relative volatility, or differences in boiling points (also known as flash points), of the components to be separated [3]. It is a process in which liquid or vapour mixtures of two or more substances are separated into different component fractions of desired purity by application and removal of heat [4]. The major equipment used for both binary distillation and multi-component distillation in modern times is the fractional distillation column. Fractional distillation columns are made up of several components, each of which is used either to transfer heat energy or to enhance mass transfer [1].

A typical fractional distillation column contains a vertical column where trays or plates are used to enhance the component separations, a reboiler to provide heat for the necessary vaporization from the bottom of the column, a condenser to cool and condense the vapour from the top of the column, and a reflux drum to hold the condensed vapour so that liquid reflux can be recycled back from the top of the column [1]

The vertical shell together with the condenser and reboiler constitute a fractional distillation column [5]. The liquid mixture that is to be processed in the fractional distillation process is known as the feed. The feed is usually introduced somewhere close to the middle of the column to a tray known as the feed tray. The feed flows down the column where it is collected at the bottom in the reboiler [6].

One of the major factors in influencing the overall heat balance of a distillation column system is the feed temperature. Feed temperature can be controlled in order to reduce the required energy input from the reboiler at the same degree of separation [7]. However, increasing feed temperature does not always improve the overall energy efficiency of a distillation unit. Most distillation columns operate with feed temperature controlled to some fixed value at or close to the boiling point of the feed, variations in this temperature being seen as unwanted disturbances to the column [8]. Careful review of feed temperature and phase is critical to minimize the distillation unit's overall energy consumption [7].

In petroleum refineries, fractional distillation is widely used to isolate and purify volatile materials. This is due to its suitability and robust nature. Thus, an efficient process control of the distillation process is vital to maximize the production of satisfactory purity end products [3]. This distillation process becomes challenging due to the presence of dynamic variables such as feed flow, temperature, pressure, level, and composition [9].

Furthermore, fractional distillation columns are highly complex systems characterized by nonlinear dynamics, multiple equilibrium points and operational modes [10]. This makes temperature the hardest parameter to control in a fractional distillation column system. It exhibits a high process and measurement lag and can also be ambivalent as a measure of composition. Therefore, adequate control of the internal temperature of a distillation column is required for a more efficient and high performance of the process. It is typical and required to provide three-mode controllers for all temperature conditions [11].

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ISSN 2229-5518

In order to achieve an effective control of the fractional distillation column temperature, the measurement of column temperature is very important. The three most important types of temperature measurement are the thermocouple measurement, resistive temperature and infrared measurement. In most cases the correcting device to which the controller outputs will be a control valve. Small valves can be pneumatic or electronic, larger valves are usually operated by air pressure, this may introduce additional dynamics for the correcting element, which could be approximated by a first-order time constant [11].

1.2 Literature Review

A brief review of some major researches carried out on petroleum fractional distillation columns are presented in this section. A graphical based software called LabVIEW was used to develop and model a fractional distillation column for basic crude oil plant control [12]. The use of a Distributed Control System (DCS) to control fractional distillation columns input parameters and overcome unstable furnace outlet temperature was presented [13]. A thorough research on well-instrumented pilot-scale continuous distillation column interfaced with an industrial distributed control system (DCS-ABB MOD 300) and a Vax cluster through the network was carried out [14]. The operation and control of crude oil distillation process was extensively studied [15]. This was a very sophisticated system involving modern technologies of utmost importance in any crude oil refinery. A research based on simulations, investigated how the operating conditions of distillation operation affected both product quality and energy efficiency of a distillation process [16]. The need to reduce the operating cost of existing units by optimization of the operating conditions of distillation led to the research. The statistical process control methodology to developing the process control of crude distillation unit (CDU) for reducing the production of specification oil products and maximizing the oil production rate was applied [17]. The process control system was established using the expertise of a practical CDU operating system provided by a group of process engineers.

Extensive research has been carried out over the years on Design or modelling, experimentation and simulations of the electronic internal temperature controller for a distillation column. However, the above review of previous research work and various evolutions in distillation column temperature control show the strengths and limitations of these technologies. For instance, with the current trend in PID (Proportional-Integral-Derivative) controller tuning for the control of various industrial processes, a more robust design of distillation column temperature controller is required to optimize the performance of the petroleum distillation process. This can be achieved using a more advanced distillation column temperature control model.

Most relevant of the reviews are the outstanding reasons behind the lingering challenge of controlling the internal temperature of a distillation column as the hardest parameter to control in a fractional distillation column system. Also, distillation columns have high process and measurement temperature control lags that can be ambivalent in relation to the measure of product composition. As a result, it is difficult to maintain the stability and purity of the distillation column system products in relation to the internal temperature of the column.

Therefore, it leaves enough evidence to believe that no suitable tool has so far been put in place to fully solve this problem. It is against this background that the study is based on the design of a more suitable electronic internal temperature controller to improve the effectiveness or performance of the control system for the petroleum fractional distillation column temperature. The distillation column temperature controller will be aimed at adequately controlling the internal temperature of the top section and the bottom section of the column.

2 DISTILLATION COLUMN TEMPERATURE CONTROL MODEL

The design of a good control system begins with the derivation of a suitable or typical process model. Control system modelling could be mathmatical or otherwise. This involves identifying the defining equations determined by the manipulated variables, load variables and the controlled variables, collectively known as the process variables. In the control / regulation of the internal temperature of a multi-component petroleum distillation column, these three categories of process variables are very essential to achieving a better performance.

The manipulated variables are those variables that can be altered or changed in order to maintain the controlled variables at their desired values. Some distillation column manipulated variables include: reflux flow, coolant flow, heating medium flow, and product flows. Load variables are those variables that provide disturbances to the column. Examples in this context are: feed flow rate and feed enthalpy, environmental conditions and coolant temperature. Controlled variables are those variables that must be maintained at a precise value to satisfy distillation column objectives. Some examples include product compositions, distillate rate, column temperatures, and column pressure.
Controlling the internal temperature of a typical multi-component petroleum distillation column requires the identification of the actual controlled, manipulated, and load variables. Some of the distillation column process variables considered in this research include: distillate rate, reflux rate, feed enthalpy, environmental conditions, coolant temperature, column temperatures, coolant flow, and heating medium flow. The main controlled variables discussed in this work is the column temperature. A simple diagram of a petroleum distillation column in figure 1 shows some of these variables.

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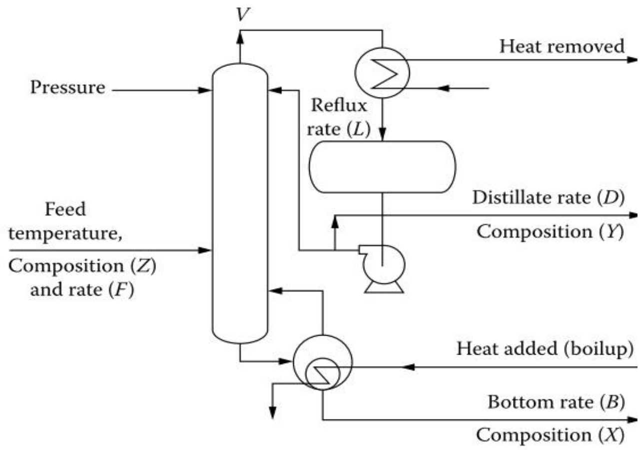

Figure 1 Diagram of a Petroleum Distillation Column [3]

Figure 1 demostrates the material and energy flow / balance in a fractional distillation column. The distillate rate, bottom rate, feed rate and composition represent the material flow while the heat added (boil-up), feed temperature, reflux rate and heat removed show the energy flow in the column.

Most petroleum distillation columns are modelled mathematically using the relationship between the material balance model and energy balance model. This is based on some assumptions.

2.1 Distillation Column Model Assumptions

The assumptions taken into consideration in developing both the material balance and energy balance models of a petroleum distillation column include the following [3]:

i) the tray temperature equals overhead vapor temperature
ii) the column operates in steady-state heat balance where the heat added to the distillation column Q_B equals the heat removed Q_T
iii) The overhead vapour is totally condensed
iv) The liquid holdups on each tray, the condenser, and the reboiler are constant and perfectly mixed.
v) The holdup of vapour is negligible throughout the system.
vi) The molar flow rates of the vapour and liquid through the stripping and rectifying sections are constant.
vii) There is constant relative volatility throughout the column as expressed by equation 1.0

$$y_n = \frac{\alpha x_n}{1 + (\alpha - 1)x_n} \quad (1.0)$$

Where, for the nth stage of the distillation process,
α = relative volatility.
y_n = vapour concentration.
x_n = liquid concentration

2.2 Energy Balance Model

In the energy balance model, the product composition of the distillation column is controlled by the rate of heat addition at the bottom of the column, the rate of heat removal at the top of the column and the heat input at the feed section. The input variables considered for energy balance are vapor rate V, boil-up rate V_B and the external reflux L. The vapor rate V above the feed tray equals the vapor boil-up rate plus the vapor entering with the feed.

$$V = V_B + F \times V_F \quad (2.0)$$

The vapor boil-up rate V_B equals the heat Q_B added by the reboiler divided by the heat of vaporization (ΔH) of the bottoms product:

$$V_B = Q_B/\Delta H \quad (3.0)$$

The liquid at the top tray of the column known as the internal reflux rate L_I, as well as the external reflux L is derived by the heat energy balance around the top of the distillation column. As in [3], assume a steady-state heat energy balance where the heat added to the distillation column Q_B equals the heat removed Q_T, that is;

$$Q_T = Q_B \quad (4.0)$$

Equation 4.0 can be expanded in terms of other heat energy balance parameters or variables as shown below [3].

$$Q_T = D \times (\Delta H_D + C_{PD} + T_t) + L_I \times (\Delta H_{L_I} + C_{PR_I} + T_t) + L \times (C_{PL} + T_r) \quad (5.0)$$

$$Q_B = D \times (\Delta H_D + C_{PD} + T_0) + L \times (\Delta H_L + C_{PL} + T_0) + L_I \times (C_{PL} + T_t) \quad (6.0)$$

Where:

ΔH_D = Heat of condensation of distillate

ΔH_L= Heat of vapourization of reflux

ΔH_{L_I} = Heat of condensation of internal reflux

C_{PD} = Specific heat of condensation of distillate

C_{PL} = Specific heat of vapourization of reflux

C_{PR_I} = Specific heat of condensation of internal reflux

T_0 = Overhead vapour temperature

T_r = External reflux temperature

T_t = Top tray temperature

L = External reflux

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L_I = Internal reflux

D = Distillate or overhead rate

Equating equation 5.0 to equations 6.0,

$$D \times (\Delta H_D + C_{PD} + T_t) + L_I \times \left(\Delta H_{L_I} + C_{PR_I} + T_t\right) + L \times (C_{PL} + T_r) = D \times (\Delta H_D + C_{PD} + T_0) + L \times (\Delta H_L + C_{PL} + T_0) + L_I \times (C_{PL} + T_t) \quad (7.0)$$

Eliminating like terms and rearranging equation 7.0,

$$D \times C_{PD}(T_t + T_0) + L_i \times \Delta H_{L_i} - \mathrm{L} \times \Delta H_L + L \times C_{PL} \times (T_r + T_0) = 0 \quad (8.0)$$

Making a simplifying assumption that the tray temperature T_0 equals overhead vapor temperature T_r (that is $T_r = T_0$),

$$L_1 \times \Delta H_{L_1} = \mathrm{L} \times \Delta H_L + \mathrm{L} \times C_{PL} \times (T_0 - T_r) \quad (9.0)$$

If a total condenser is employed, the composition of the internal reflux and external reflux are the same. This implies that $\Delta H_{L_i} = \Delta H_L$ and the external reflux is expressed as

$$L_i \Delta H_{L_i} = L \Delta H_L \left[1.0 + \frac{C_{PL}}{\Delta H_L} \times (T_0 - T_r)\right] \quad (10.0)$$

Equation 10.0 can be further reduced to

$$L = \frac{L_i \, \Delta H_{L_i}}{C_{PL} \, (T_0 - T_r)} \quad (11.0)$$

The vapor boil-up V generated by the heat input to the reboiler is calculated as in [3].

$$V = \frac{B c_B \, (t_B - t_f)}{\lambda} \quad (12.0)$$

Where:

B = the flow rate of bottom product

c_B = the specific heat capacity

t_F = the feed temperature

t_B = the distillate and bottom product temperatures

λ = the latent heat or the heat of vaporization

2.3 Material Balance Model

The material balance model involves the distillate and bottom product (residue) as manipulated variables, and the composition as the controlled variable. See figure 2. The overall material balance can be expressed mathematically as shown in equation 13.0 [3].

$$F = D + B \quad (13.0)$$

Where:

F = feed rate (The inflow).

B = bottom rate (The outflow).

D = distillate (overhead) rate (The outflow).

The Petroleum Distillation Column Heat Flowsheet showing the parameter for both energy and material balance is given in figure 2.

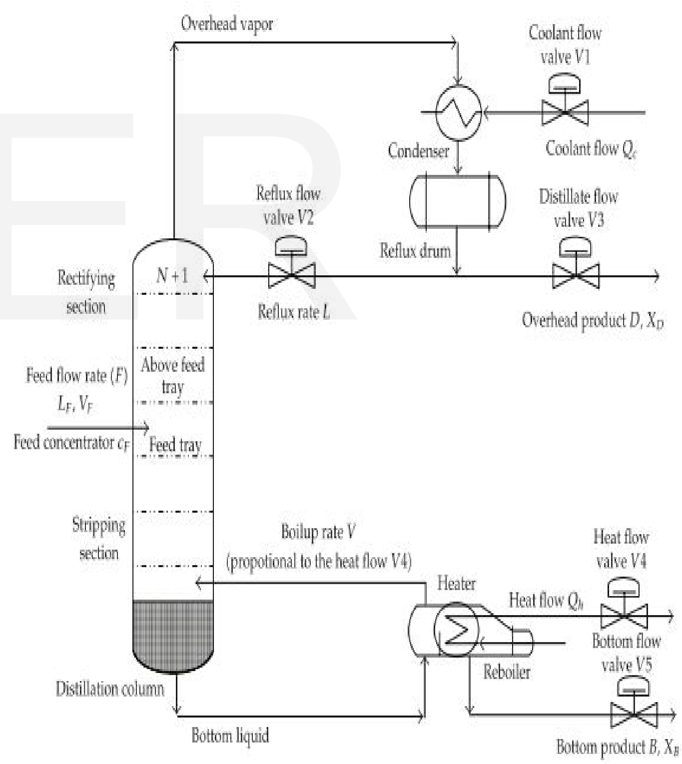

Figure 2. Distillation Column Heat Flow Diagram [1]

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From figure 1 and 2, based on the assumptions that there is a constant relative volatility throughout the column, the distillation column dynamic model can be expressed by the following equations [1]:

i) For the condenser (n = N+2):

$$M_D \frac{dx_n}{dt} = (V + V_F)y_{n-1} - Lx_n - Dx_n \quad (14.0)$$

ii) For the tray above the feed flow (n = f +1):

$$M \frac{dx_n}{dt} = V(y_{n-1} - y_n) + L(x_{n\pm1} - x_n) + V_F(y_F - y_n) \quad (15.0)$$

iii) For the tray below the feed flow (n = f):

$$M \frac{dx_n}{dt} = V(y_{n-1} - y_n) + L(x_{n+1} - x_n) + L_F(x_F - x_n) \quad (16.0)$$

iv) For the reboiler (n = 1):

$$M_B \frac{dx_1}{dt} = (L + L_F)x_2 - Vy_1 - Bx_1 \quad (17.0)$$

Where the different terms used are expressed as follows:

N = the number of trays

M = hold up rate in each of the trays

M_B = hold up rate in the column base

M_D = hold up rate in the reflux drum

n = number of stage

D = Column diameter

f = feed position

X_F= feed flow rate

Y_F = bottom product flow rate

L_F = reflux flow rate

V_F = distillate flow rate

V = vapor boil-up

L = External reflux

The dynamic model represented by equations from 14.0, 15.0, 16.0 and 17.0 is nonlinear due to the vapor-liquid equilibrium relationship between y_n and x_n. The control technique used in this research to effectively improve the performance of this process in terms of temperature control requires a linear model.

In order to obtain a linear control model for this nonlinear system or distillation process, it was assumed that the variables deviate only slightly from the normal operating conditions $(x_n - \bar{x}_n)$.

The full-order linear model which is represented by a two inputs-two outputs plant can be expressed as a reduced order linear model as in equations 18.0 [3].

$$\begin{bmatrix} x_D \\ x_B \end{bmatrix} = \frac{1}{1+\tau_c s} G(0) \begin{bmatrix} L \\ V \end{bmatrix} \quad (18.0)$$

Where $G(0)$ is the steady-state gain which can be expressed as follows:

$$G(0) = -CA^{-1}B = \begin{bmatrix} 0.0042 & -0.0062 \\ -0.0052 & -0.0072 \end{bmatrix} \quad (19.0)$$

τ_c is the standard time constant with a conventional value given by τ_c =1.9588.

The reduced-order linear model of the plant is a first-order system given by

$$\begin{bmatrix} x_D \\ x_B \end{bmatrix} = \frac{1}{1+1.9588\, s} \begin{bmatrix} 0.0042 & -0.0062 \\ -0.0052 & -0.0072 \end{bmatrix} \begin{bmatrix} L \\ V \end{bmatrix} \quad (20.0)$$

Where

x_D = **distillate temperature change rate**

x_B = **Bottom product temperature change rate**

The state space representation of equation 20.0 is given as follows:

$$\begin{bmatrix} \dot{x}_1 \\ \dot{x}_2 \end{bmatrix} = \begin{bmatrix} -0.5105 & 0 \\ 0 & -0.5105 \end{bmatrix} \begin{bmatrix} x_1 \\ x_2 \end{bmatrix} + \begin{bmatrix} 1 \\ 1 \end{bmatrix} \begin{bmatrix} L \\ V \end{bmatrix} \quad (21.0)$$

$$y(t) = \begin{bmatrix} 0.0021 & -0.0031 \\ -0.0026 & -0.0037 \end{bmatrix} \begin{bmatrix} x_1 \\ x_2 \end{bmatrix} \quad (22.0)$$

Where x_1 and x_1 are the state variables and system matrix A is assumed to be an indentical matrix (2x2). The transfer function, $G(s)$ of the distillation column derived from the state equation is presented in equation 23.0.

$$G(s) = \frac{0.001\,s + 0.0005}{s^2 + 1.021\,s + 0.2606} \quad (23.0)$$

The transfer function in equation 23.0 represents the Laplace transform of the ratio of the heat input into the column and the heat output from the column. Therefore the transfer function can also be represented by

$$\frac{Q_T}{Q_B} = \frac{0.001\,s + 0.0005}{s^2 + 1.021\,s + 0.2606} \quad (24.0)$$

Since this is aimed at designing an electronic internal temperature controller for a distillation column, the transfer function is converted from the continuous domain in equation 23.0 to a discrete domain shown in equation 25.0.

$$G(z) = \frac{0.006284\,z - 0.006252}{z^2 - 1.99\,z + 0.9898} \quad (25.0)$$

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Simulation was carried out using MATLAB Step Response and Root Locus to determine the initial performance of the system without a controller.

2.4 PID Controller Design

A Proportional Integral Derivative (PID) controller is a generic control loop feedback mechanism widely used in industrial control systems and regarded as the standard control structures. A PID controller, sometimes called three-term control, calculates an error value as the difference between a measured process variable and a desired setpoint. The controller attempts to minimize the error by adjusting the process through use of a manipulated variable. The PID controller has the optimum control dynamics including zero steady state error, fast response (short rise time), no oscillations and higher stability. The necessity of using a derivative gain component in addition to the PI controller is to eliminate the overshoot and the oscillations occurring in the output response of the system. One of the main advantages of the PID controller is that it can be used with higher order processes including more than single energy storage. Figure 3 shows the control system schematic model of a general PID controller.

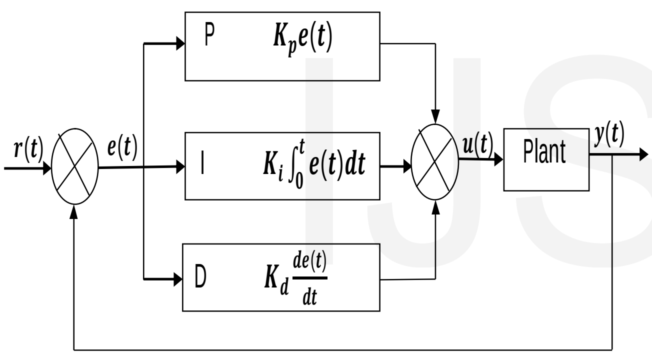

Figure 3: The Model of a PID Controller Configuration figure in the caption.

In figure 3, K_p = proportional gain, K_i = integral gain, and K_d = derivative gain. The overall mathematical description of linear relationship existing between the controller output, u(t) and the error, e(t) in figure 3 is expressed as in equation (26.0).

$$u(t) = K_p\left(e(t) + \frac{1}{T_i}\int e(t)dt + T_d\frac{de(t)}{dt}\right) \qquad (26.0)$$

Where T_i = integral time and T_d = derivative time.

Applying the PID controller to any control system involves adjusting the values of gain K_p, K_i and K_d in order to get the best response of the system. The selection of PID controller gain values causes the variation of observed response with respect to desired response. Equation 26.0 can be represented in discrete form as shown below:

$$\frac{u(z)}{e(t)} = K_p + K_i\frac{T_2(z+1)}{2(z-1)} + K_d\frac{z-1}{zT_2} \qquad (27.0)$$

Where T_2 = Sampling time.

2.5 Performance Criteria

The improved or optimized performance step response parameters to be achieved in this work are stated as follows:

a) Percentage overshoot less than 10% to a unit step input.
b) Settling time less than 10 seconds to a unit step input.
c) Rise time of less than 6 seconds to a unit step input
d) Steady-state error less than 0.2%

2.6 Distillation Column Model with PID Controller

The overall distillation column model integrated to the PID controller is shown in figure 4.

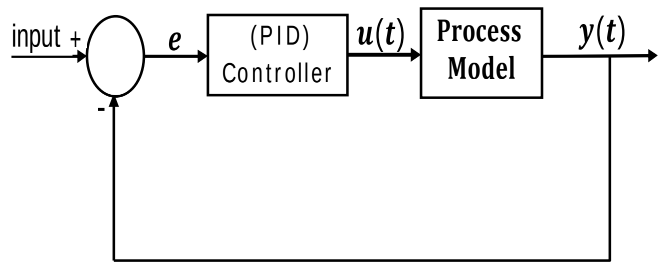

Figure 4 Distillation Column Temperature Control System

In Figure 4 the PID controller was fine-tuned using the MATLAB PID Tuner Application. The PID controller tuning was aimed at achieving an improved performance of the system. The controller gains of the fine-tuned PID controller obtained from the PID Tuner Application in a MATLAB Simulink modelling environment (As shown in figure 4) are K_p = 23.5042, K_i = 20.4505, K_d = 0.0001. Therefore, substituting these new PID controller gains into equation 3.44, the actual controller is presented in equation 28.0.

$$G_c(z) = \frac{u(z)}{e(t)} = 23.5042 + 20.4505\frac{T_2(z+1)}{2(z-1)} + 0.0001\frac{z-1}{zT_2} \qquad (28.0)$$

The forward path in figure 3 can be represented by the product of the transfer function G(z) (equation 25.0) and the discrete PID controller $G_c(z)$ (equation 28.0). The combined equation is derived as follows:

$$G_{cloop}(z) = G_c(z) \times G(z)$$

$$= \frac{0.001483z^2 + 0.002946\ z - 0.001463}{z^3 + 2.99z^2 - 2.98\ z + 0.9898} \qquad (29.0)$$

The above equation is simulated and the results are shown in section Three.

3 RESULTS AND DISCUSSION

The results obtained from the simulation of the petroleum distillation column system material balance and energy balance models as well as the PID (proportional-Integral-derivative) controller design and fine-tuning are presented in Section 2 are discussed in this section. The results include the MATLAB Simulink simulation Step Responses and Root Lo-

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cus of the petroleum distillation column mathematical model transfer function and the state equation. The linear control parameters values obtained from the simulation were compared to the performance criteria stated in section 2.5. This was to determine possible ways of improving the electronic internal temperature control / regulation of a petroleum distillation column.

Analysis of the MATLAB Simulink simulation results from the distillation column mathematical model integrated to a PID controller design was carried out to show the extent of optimizing of the system. In the analysis, the Step Response parameters such as settling time, steady state error, rise time as well as the Root Locus stability features are included. The corresponding practical positive effects of these linear control parameters to a typical distillation column temperature or heat control system are not left out.

3.1 Step Response of Transfer Function (Gs)

Figure 5 shows the initial Step Response plot of the petroleum distillation column mathematical model transfer function before a PID controller was designed, fine-tuned and integrated to it. The results in the plot represent the unstable and poor performance state of the electronic internal temperature control of the distillation column. The settling time is 7.5 seconds, indicating slow or poor response of the system to fluctuations in change in the internal column temperature due to cold weather. Also, the high rise time of 4.3 seconds shows the inadequate response in maintaining a regulated internal column temperature. The abnormal steady state error is an indication that there is a persistent uncontrolled temperature rise or temperature drop in the column.

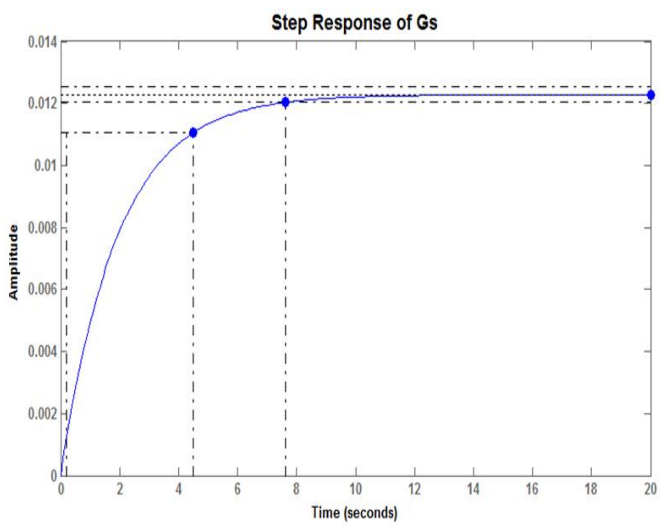

Figure 5: Initial Step response of Transfer Function, G(s)

3.2 Step **Response of State Equation**

The initial Step Response plot of the state equation representing the petroleum distillation column mathematical model before the addition of a PID controller is shown in Figure 6. The Step Response plot presents a clear picture of the actual state of the distillation column operating in steady state. The abnormal sharp or nature of the plot (inverted plot) further shows the unstable and low quality functionality of the system. The settling time is 7.5 seconds, indicating slow or poor response of the system to fluctuations in change in the internal column temperature due to cold weather. Moreover, the parameter values and scale of the plot are located in the reverse on negative axis far below the stability line (which is normally 1.0).

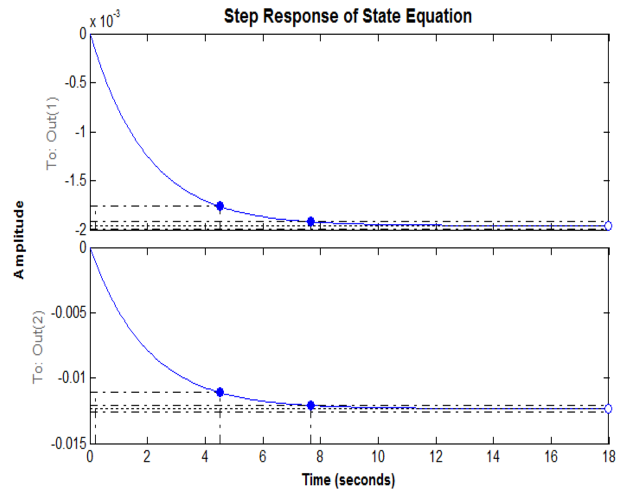

Figure 6: Initial Step response of State Equation.

3.3 Step Response of Closed-Loop Transfer Function

The unit step response plot of the closed-loop transfer function and controller representing the petroleum distillation column mathematical model after the addition of a PID controller is shown in figure 7. The control parameters obtained in figure 7 are: a settling time of 3.8 seconds, the rise time of 2.3 seconds, the percentage overshoot of 0% and a steady-state error less than 2%.

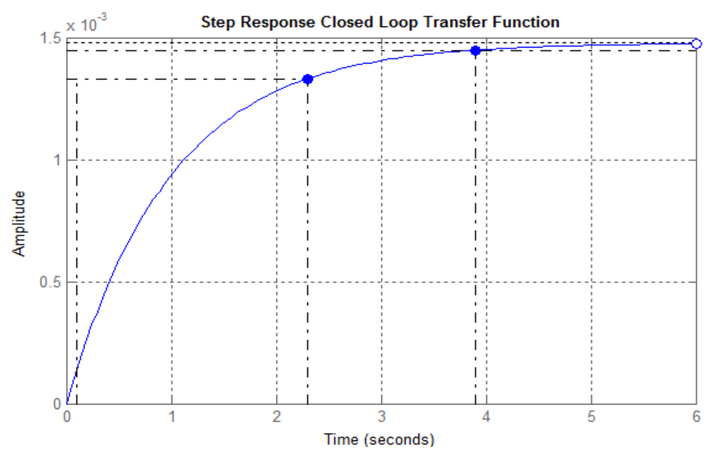

Figure 7: Unit Step response of Closed-Loop Transfer Function, G_{cloop}.

The settling time is 3.8 seconds, indicating faster response of the system to fluctuations in the internal column temperature due to cold weather. The step response plot shows that the

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manipulated variables as well as the controlled variable (internal temperature of the column) have been improved. Another improvement is a moderate and constant internal column temperature corresponding to the reduced rise time.

4 CONCLUSION

In conclusion, this research has so far presented an improved mathematical model of a Petroleum Distillation Column operating in steady state based on material balance and energy (heat) balance. The resulting combined equation was linearized to obtain a linear model transfer function. This allowed the application of linear control engineering techniques such as the step response. These control techniques help in ascertaining one of the best possible methods of improving the performance of the distillation column. An initial performance check was successfully carried out on the model state equation and transfer function using the linear control techniques. This proved the need for a fine-tuned classical controller.

The design of a classical electronic internal temperature controller that optimizes the performance of the column has been presented in this research in details. Due to the complex nature petroleum distillation column temperature control system, a PID controller has been design as a more suitable linear control strategy to optimize the system. The PID controller has been well tune using the MALAB PID Tuner to obtain excellent controller gains. The controller has been integrated to the column temperature control system and has achieved a better result in the system functionality.

In the simulation figures and values, it was possible to prove that the integrated PID controller provided the desired column temperature control with great speed and accuracy. The simulation results show that, PID controller integrated to overall closed-loop column temperature control system gives a better performance compared to previous system. Based on the analysis it was demonstrated that the settling time, percentage overshoot, steady state error and rise time were greatly improved by the by the new controller achieved in this research. Therefore, ultimate goal of the research which is to improve the internal temperature control of a petroleum distillation column for a better efficiency and best quality petroleum products was achieved to an acceptable extent.

4.1 Recommendation

The linear control system techniques employed in this work can be replaced with a few others to obtain similar results. There are many potential alternative techniques that can be employed for further research to achieve similar results as obtained in this research. Some of these techniques include: Advance three term (PID) control, Phase Lead Compensation, Linear Quadratic Regulator (LQR), Fuzzy Logic Control, Robust Control etc. All the above techniques can be applied to optimizing the petroleum distillation column to get an acceptable internal temperature control performance. Also, a continuous time method or a digital method based on Z-transforms can be used to implement each of the above listed techniques.

Generally, any particular technique could have advantages over the others in different areas of applications depending on the practical situation in which the column temperature controller is used. It has been recently discovered that special features of the different techniques have different effects on the distillation column temperature control under different control designs.

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International Journal of Scientific & Engineering Research Volume 7, Issue 10, October-2016
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