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Simulation Study in Control System Configuration of a Distillation Column

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Abstract

This research deals with the dynamic model and control of the distillation tower by applying a multi-loop control system in Matlab/Simulink for a dual mixture of benzene and toluene. The method of control in this research is PID multi - loop. The results showed a good performance of the benzene-toluene system using a single-ended temperature control structures with using bottom and top flow rate controlling bottom and top level respectively based on the calculation of the absolute error integration (IAE).

Keywords: Distillation; PID control; MATLAB/Simulink

Introduction

Distillation is probably the most popular and important process studied in chemical engineering literature. distillation is used in many chemical processes for separating feed streams and for purification of final and intermediate product streams. most distillation columns handle multicomponent feeds, but many can be approximated by binary or pseudo-binary mixtures. dynamic mathematical models are widely used in distillation process design and operation. dynamic modeling and simulation have proven to be an insightful and productive process engineering tool [1].

Rouzineau et al. [2,3] developed model unbalanced distillation Tower, multi-component, and this includes the mass transfer, specific, described Maxwell equation. It also tested the unbalanced model compared with balanced model for several component and reached a consensus to good results between the two models, but the difficulty of finding the Murphree efficiency and the sensitivity of the unbalanced model thickness of the boundary between the steam and the liquid.

Gutiérrez-Oppe et al. [4] studied the behavior of the three-phase distillation in a sieve tray column, through the effects of the process and geometrical variables. The experimental values were compared with predicted values obtained by simulation using the equilibrium and nonequilibrium models. Three-phase distillation has been used for glycerin dehydration using toluene as entertainer, to avoid the glycerol degradation by distillation at atmospheric pressure. Batista and Meirelles [5] studied continuous spirit distillation by computational simulation, presenting some strategies of process control to regulate the volatile content. The commercial simulator Aspen (Plus and dynamics) was selected. A standard solution containing ethanol, water and 10 minor components represented the wine to be distilled. A careful investigation of the vapor liquid equilibrium was performed for the simulation of two different industrial plants. The simulation procedure was validated against experimental results collected from an industrial plant for bioethanol distillation. The simulations were conducted with and without the presence of a degassing system, to evaluate the efficiency of this system in the control of the volatile content. To improve the efficiency of the degassing system, a control loop based on a feedback controller was developed.

Control plays a fundamental role in modern technological systems. The benefits of improved control in industry can be immense include improved product quality, reduced energy consumption, minimization of waste material, increased safety levels, and reduction of pollution. Kano et al. [6] applied modern method of control called (Predicative inferential control) through the predictability of the focus of the process control variables directly, rather than the current estimates to focus the results of the dynamic simulation showed that the outcome control system supplementing the cascade control to a large extent from other control systems. Fileti et al. [7] developed control algorithms obscurity on the computer and applied practically in the distillation device to separate a mixture of hexane and heptane. The controlling of the fuzzy weather got it by changing the gain and belonging to the variables inflows and outflows were compared to the results of this method with traditional dominant results showed that the fuzzy control system best performance of traditional control system (PID). Shin et al. [8] developed fuzzy controller of multi-component distillation tower to separate a mixture of toluene, benzene and xylene. This method focusses at the top of the tower toluene using the return rate and controls the concentration of toluene in the bottom using the amount of heat processed to boiler and xylene is controlled by the concentration of gasoline to control the flow rate of the tower Side stream and through virtualization results concluded that this method is appropriate high and serious performance.

Duraid and Mohammad [9] studied the dynamic behavior of continuous distillation column and implemented different type of control strategies for the separation of binary mixture composed of ethanol and water. The study includes the application of (PI and PID) controller to control the top and bottom concentration and tuning parameters of this controller by using (Cohen-coon) method and trial and error method. The study also includes designing fuzzy logic controller using MATLAB program and applying it on dynamics model for two control scheme and tuning it by trial and error method. The Comparison between fuzzy logic and (PID) is done and measure the controller performance by using mean square error (MSE) and integral square error (ISE) that the result performance showed that the fuzzy controller best than the conventional controller through fast access to the desired value and cancelling the disturbances.

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This study includes the dynamic behavior of distillation column by measuring the responses of temperature of each tray, level of top and bottom to step change in the amount of feed, amount of liquid reflux and concentration of feed flow. Also includes the deriving mathematical model for this process according to the basic principles and use this model in building the simulation case. The study includes design PID control system to control the top temperature, condenser temperature and level of top and bottom by MATLAB simulation and choose the best structure of PID controller at different conditions of feed flow, weight fraction of feed flow, reflux ratio and level through the simulation.

Material and Methods

Distillation column

The system of distillation is binary and nonideal liquid solution with ideal vapor system. The feed to the tower a mix at boiling point, saturated liquid and temperature it 92.3°C. The vapor hold-up on the tray is neglected compared to the liquid loaded on the tray. The flowrate of liquid to each tray is constant. The liquid and vapor on each tray in equilibrium state. The vapor output from the top of the distillation tower is intensive entirely for obtaining a fixed pressure of the tower and at 1 bar. The reflux is liquid at boiling point.

The material balance of the distillation can be derived as the following. The differential equations that represent mass balance of the distillation column can be found as the following:

Mass rate in-Mass rate out=Mass Accumulation ... (1)

For condenser and reflux drum the mathematical model can be represented as: -

$$\frac{dM_D}{dt} = V_{NT} - (R + D) \qquad (2)$$

$$\frac{dM_D X_D}{dt} = V_{NT} Y_{NT} - \left(R + L_{NT}\right) X_D - V Y_D \qquad (3)$$

Where,

R: Reflux flowrate to the tower (Kg/hr), D: Flow rate of top product (Kg/hr), V_{NT}: Vapor flow rate on tray (Kg/hr), M: Liquid holdup (Kg), L_{NT}: Liquid flow rate on tray (kg/hr) and X,Y: Weight fraction of vapor and liquid on each tray according to subscribe.

For top tray the mathematical model can be represented as:

$$\frac{dM_{NT}}{dt} = V_{NT-1} + R - L_{NT} - V_{NT} \qquad (4)$$

$$\frac{dM_{NT} X_{NT}}{dt} = V_{NT-1} Y_{NT-1} + X_D R - L_{NT} X_{NT}$$

$$- V_{NT} Y_{NT} \qquad (5)$$

For n^{th} tray the mathematical model is:

$$\frac{dM_n}{dt} = L_{n+1} - L_n + V_{n-1} - V_n \qquad (6)$$

$$\frac{dM_n X_n}{dt} = X_{n-1} L_{n+1} - X_n L_n + Y_{n-1} V_{n-1} - Y_n V_n \qquad (7)$$

Where: -

n: Number of trays.

For feed tray the mathematical model is:

$$\frac{dM_{NF}}{dt} = L_{NF+1} - L_{NF} + V_{NF-1} + F_L \qquad (8)$$

$$\frac{dM_{NF}}{dt} = X_{NF+1} L_{NF-1} - X_{NF} L_{NF} + Y_{NF-1} V_{NF-1}$$

$$- Y_{NF} V_{NF} + F_L Z_L \qquad (9)$$

Where, F: Feed flow rate (Kg/hr), Z_L: Concentration of feed flow rate for bottom tray the mathematical model is:

$$\frac{dM_1}{dt} = L_2 - L_1 - V_1 + V_B \qquad (10)$$

$$\frac{dM_1 X_1}{dt} = X_2 L_2 - X_1 L_1 - Y_1 V_1 + Y_B V_B \qquad (11)$$

For reboiler the mathematical model is:

$$\frac{dM_B}{dt} = L_1 - V_B - B \qquad (12)$$

$$\frac{dM_B X_B}{dt} = X_1 L_1 - Y_B V_B - B X_B \qquad (13)$$

Where, B: Bottom flow rate (Kg/hr). For condensation tank the mathematical model of mass balance is:

$$\rho A \frac{d\left(H_T\right)}{dt} = LT - (L + D) \qquad (14)$$

Where, ρ: Density of fluid (kg/m^3) and A: Area of storage tank (m^2) For reboiler tank the mathematical model of mass balance is:

$$\rho A \frac{d\left(H_b\right)}{dt} = L_1 - B \qquad (15)$$

The differential equations of energy balance which representing the distillation column after mass balance can be finds as following:

Heat rate in-Heat rate out=Heat Accumulation …. (16)

Energy balance on tray (n) depending on equation (16)

$$\frac{M_n C_{nd}\left(h_n\right)}{dt} = H^L_{n+1} L_{n+1} - H^L_n L_n + H^V_{n-1} V_{n-1}$$

$$- H^V_n V_n \qquad (17)$$

Where,

Hn=CnTn, H^L_n: Enthalpy of liquid on tray (n) (KJ/hr), H^V_n: Enthalpy of vapour on tray (n) (KJ/hr), Cn: Heat capacity on tray (n) (KJ/kg °C), and Tn: Temperature on tray (n)(°C).- Energy balance on reboiler

$$C_1 \frac{M_1 d\left(T_1\right)}{dt} = C_1\left(T_2 L_2 - T_1 B\right) - H^V_1 V_1 + QR \qquad (18)$$

Where: -

QR: Heat amount supply on reboiler (KJ/hr) and C_1: Heat capacity for residual bottom column (KJ/kg°C)

- Energy balance on condensation

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$$C_{NT}\frac{M_{NT}d(T_{NT})}{dt} = L_{NT}C_{NT}V_{NT-1}(T_{NT-1} - T_{NT})$$
$$- H_{NT-1}^{V}V_{NT-1} - QC \qquad (19)$$

Where: - QC: Heat amount reduced in condensation (KJ/hr) and NT: heat capacity for distilled condensation (KJ/kg°C).

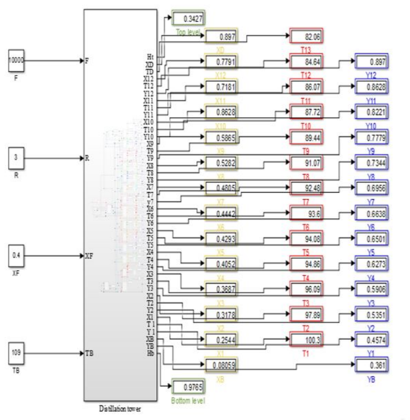

Figure 1: Open-loop control system.

Simulation work

A simulation program is built for the distillation column by using the program MATLAB/Simulink version (R2015a) from (Math works), which are a software modeling dynamical systems and simulation and analysis, whether linear or non-linear. By using Simulink, we can build models from scratch or amendment to existing models and of interest is the study of the characteristics of control and dynamic situation. The mathematical model is built for the distillation column in the form of a set of systems, and each system component with a set of subsystems which represents the mathematical model equations for distillation column. The model of the distillation column used throughout the paper is developed by Acharya et al. [10] composed by the mass, component mass and energy balance equations used as basis to implement the Simulink diagram. Tables 1 and 2 show the simulation runs that have been made using a simulation program to control methods system

Run No.	Type of Disturbance	Value
1	Step change in feed flow rate, Kmol/hr	500-525
2	Step change in reflux ratio	3-3.3
3	Step change in feed composition	0.4-0.45

Table 1: Simulation runs for dynamic systems.

Run No.	Type of Disturbance	Value	Structure No
1	Top flow rate, kmol/sec	1-1.4	1
2	Bottom flow rate, kmol/sec	05-Jul	1
3	Reflux ratio	03-Apr	1
4	Top flow rate, kmol/sec	1-1.4	2
5	Bottom flow rate, kmol/sec	1-1.4	2
6	Reflux ratio	03-Apr	2
7	Top flow rate, kmol/sec	1-1.4	3
8	Bottom flow rate, kmol/sec	1-1.4	3
9	Reflux ratio	03-Apr	3
10	Top flow rate, kmol/sec	1-1.4	4
11	Bottom flow rate, kmol/sec	1-1.4	4
12	Reflux ratio	03-Apr	4

Table 2: Simulation runs for control methods system.

Simulation of open-loop system: Open loop control is by far the simpler of the two types of control theory. In open loop control (Figure 1), there is some sort of input signal (digital or analog), which then passes through amplifiers to produce the proper output and is then passed out of the system. Open loop controls have no feedback and require the input to return to zero before the output will return to zero.

Simulation of the PID control system: After running the dynamic model that has been developed using Simulink and determining the extent of the system's response to some of the changes, as shown in Figures 2-5.

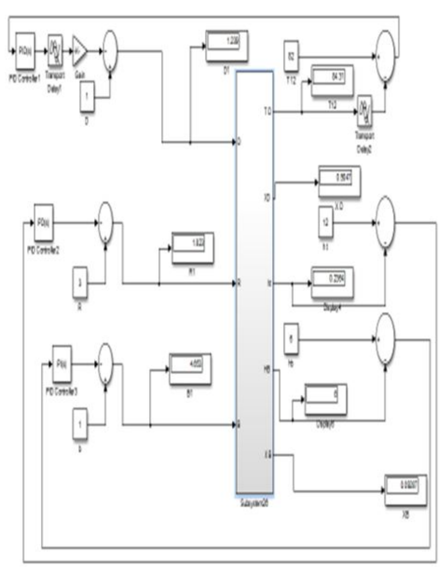

Figure 2: Multi control structure 1.

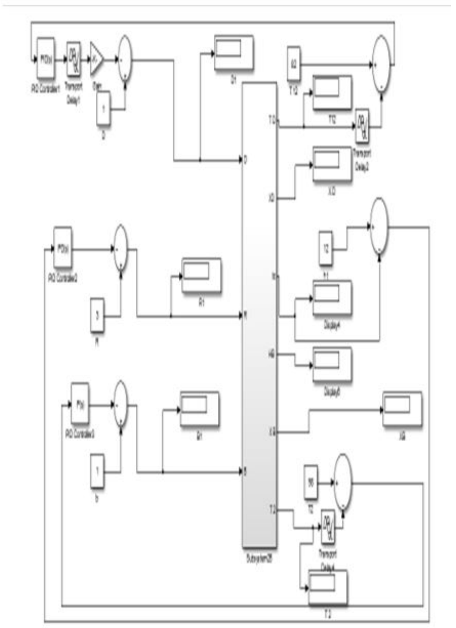

Figure 3: Multi control structure 2.

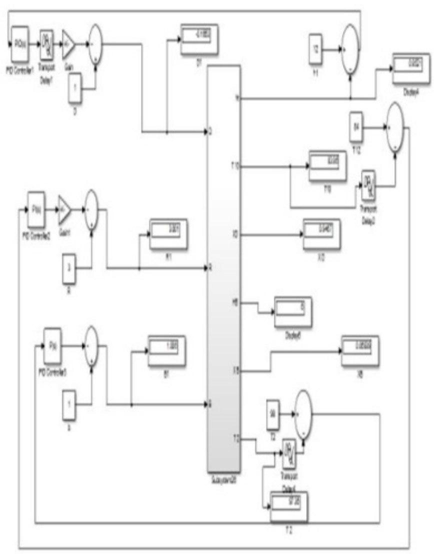

Figure 4: Multi control structure 3.

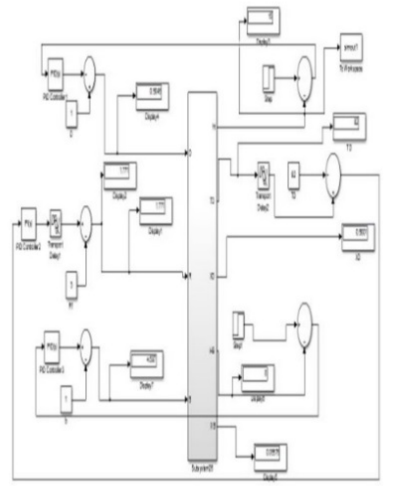

Figure 5: Multi control structure 4.

The system studied: The distillation column is used for the separation of a binary mixture of benzene and toluene which enters as a feed stream with flow rate and composition between the rectifying and the stripping section. The column consists of 14 trays. The overhead vapor is totally condensed in a water-cooled condenser. The reboiler is heated by steam, and the preheated feed stream enters the column at the feed tray as saturated liquid. The process inputs that are available for control purposes are the amount of distillate in the top and the reflux flow rate and amount of bottom product.

Results and Discussion

The results include the dynamic behavior of distillation column for separation the binary mixture (benzene-toluene) and comparing between two different control systems PID and open-loop on the modified dynamic model of distillation column are made. Reflux ratio is the basic element to control distillation column; it represents the amount of the feed flow, concentration of feed. Step changes in reflux ratio, concentration of feed theoretically by using MATLAB simulation program.

Effect of feed concentration

The initial steady-state feed composition is 40 moles % of the benzene. The feed composition is change by a step change from 40% to 45%. The responses of the top tray temperature, and condenser are shown in Figures 6 and 7. Since both the reflux and distillate product rates were held constant, the response of the top tray temperature shown in Figure 7 would be expected - a monotonic increase due to the increase in the higher - boiling components in the column. The increased benzene content in the top also affected the bottom of the column by initially increasing the bottoms temperature. This increased the temperature driving force in the reboiler, temporarily increasing the evaporation.

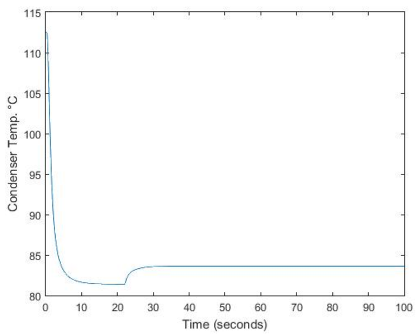

Figure 6: Condensing temperature responses to step change in feed composition of benzene from (0.40 to 0.45 mole%).

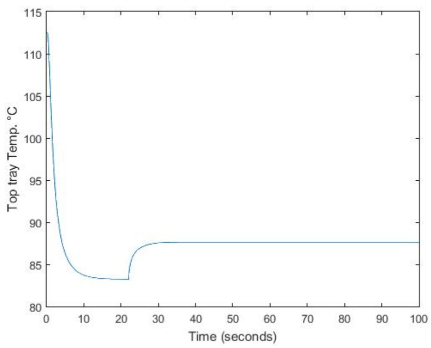

Figure 7: Top tray temperature responses to step change in feed composition of benzene from (0.40 to 0.45 mole%).

Effect of feed rate

The open-loop response to a change in feed rate was examined by imposing a 5% step increase to the feed rate, from 500 to 525 kmoles per hour. The feed composition, reflux ratio and temperature were held constant, to maintain the overall column material balance, the reboiler level controller was functional. The responses of the column top tray temperature and condensing temperature are shown in Figures 8 and 9, respectively. the distillate composition became purer because the distillate rate was held constant. Before the upset, there were 185 moles of benzene fed to the column and 194.4 moles of distillate product, so a pure distillate product was prevented by material balance considerations. with the additional feed, 194 moles of benzene were fed, making a very pure distillate product feasible. These represent the dew point and bubble point, respectively, of the vapor going overhead to the condenser. Clearly, as the purity of a single component product (such as benzene) increases, its dew point and bubble point will converge.

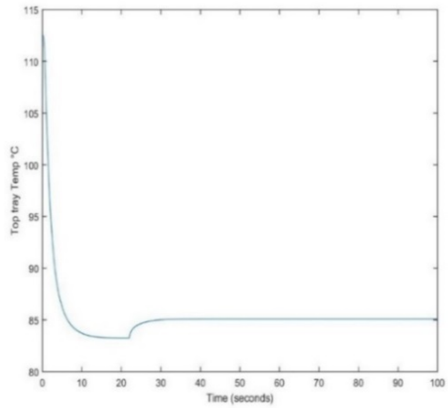

Figure 8: Top tray temperature responses to step change in feed rate from (500 to 525 kmoles/hr.).

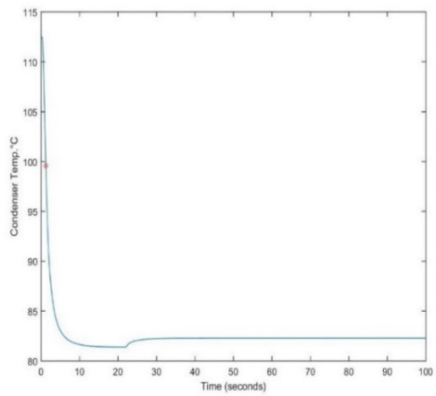

Figure 9: Condenser temperature responses to step change in feed rate from (500 to 525 kmoles/hr).

Effect of reflux rate

The response of the column top tray temperature, and condensation temperature, show in Figures 10 and 11 respectively, to a 10% step decrease in the reflux rate. The response of the top tray temperature is shown in Figure 11, a monotonic increase to the new steady state. However, the bubble point of the condensing vapor did not behave as one might expect. Ordinarily, the assumption might be made that the condensation temperature would also increase. the response to feed composition, and the response to feed rate exhibited similar behavior because of upsets in the material balance, unlike the response to the thermal upsets of the other open-loop examples. That is, the response to material balance upsets are monotonic.

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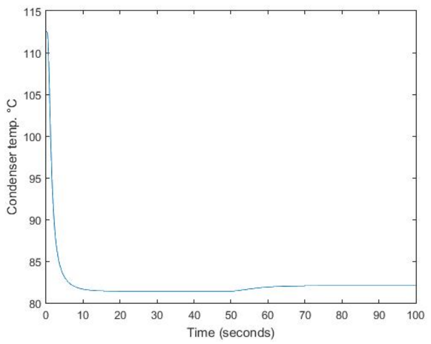

Figure 10: Condenser temperature responses to step change in reflux ratio from (3 to 3.3).

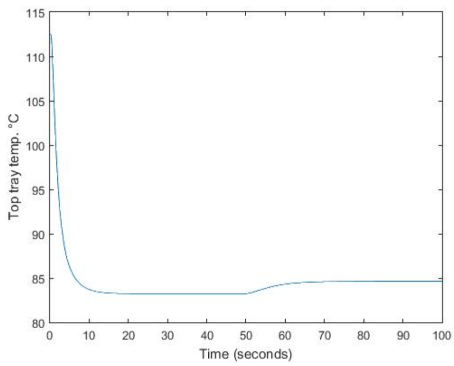

Figure 11: Top tray temperature responses to step change in reflux ratio from (3 to 3.3).

Controller	Structure	K_c	τ_I(minute)	τ_D(minute)
PID	1	0.224	1.27	1.87
PI	1	0.224	1.27	
PD	1	0.224		1.87
PID	2	0.224	1.27	1.87
PI	2	0.224	1.27	
P	2	0.224		
PID	3	0.224	1.27	1.87
PI	3	0.224	1.27	
P	3	0.224		
PID	4	0.224	1.27	1.87
PI	4	0.224	1.27	
PID	4	0.224	1.27	1.87

Table 3: Controller parameters by auto method in MATLAB Simulink.

Controller	Structure	K_c	τ_I(minute)	τ_D(minute)
PID	1	0.5	0.1	1
PI	1	0.1		0.1
PD	1	0.1	1	
PID	2	0.1	0.1	0.1
PI	2	0.1		0.1
P	2	-0.1		
PID	3	0.1	0.1	1
PI	3	-0.01	-0.01	
P	3	-0.01		
PID	4	1	0.04	5
PI	4	10	1	
PID	4	-1	-2.7	-4

Table 4: Controller parameters by trial and error.

Control of distillation column

The four structure of Multi-PID method is used to control level and temperature of distillation column and compared between those structure.

PID controller: The controller parameters for four control structures are given in Table 3. The controller's parameters for four control structures through auto method in MATLAB Simulink noted that these parameters don't give good results with parameters chose manually, so that done tuning the controllers by trial and error. Table 4 explain the controllers' parameters which done tuning manually by trial and error.

Level control: Level controllers are used to maintain the level in the accumulator, the reboiler and the intermediate accumulator of a distillation column. Loose level control on the accumulator and reboiler has been shown to worsen the composition control problem for material balance control configurations (when either bottom or top flow is used as a manipulated variable for composition control). When top or bottom flow is adjusted, the internal vapor/liquid traffic changes only after the corresponding level controller acts because of the change in top or bottom flow. On the other hand, if a level controller is tuned too aggressively, it can result in oscillations passed back to the column and contribute to erratic operation. When the reboiler, duty is set by the level controller on the reboiler. level control There will be two level loops on a distillation column as: the column base level (bottom level) must be maintained at an acceptable value. the reflux drum level (top level) must be maintained at an acceptable value and to test this controller we set step change to systems are built in top flow, bottom flow and reflux ratio and calculate the integral absolute error in Table 5 it shows best result and those run show in Figures 12-17.

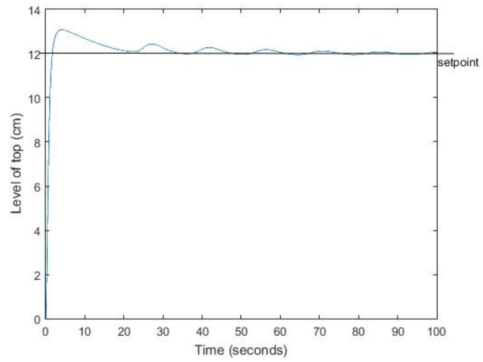

Figure 12: The response of the level of top for step change in bottom flow from (1 to 1.4 kmol/s) at setpoint 12 cm at PID of control structure 4.

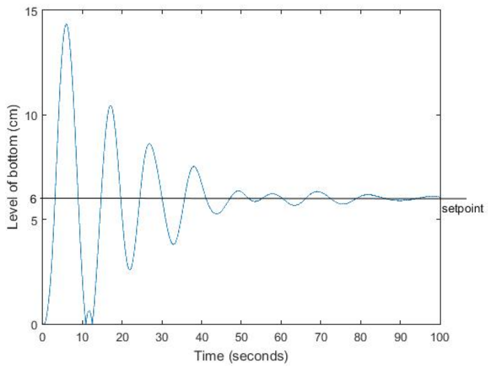

Figure 13: The response of the level of bottom for step change in bottom flow from (1 to 1.4 kmol/s) at setpoint 6 cm at PID of control structure 4.

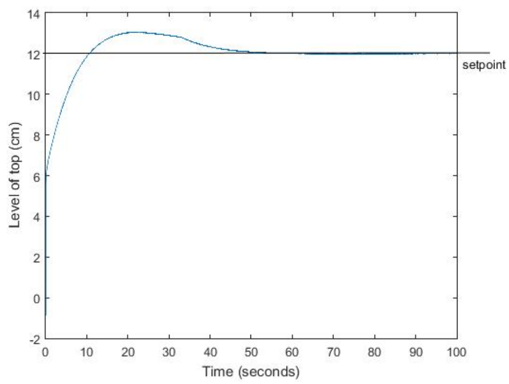

Figure 14: The response of the level of top for step change in top flow from (1 to 1.4 kmol/s) at setpoint 12 cm at PID of control structure 4.

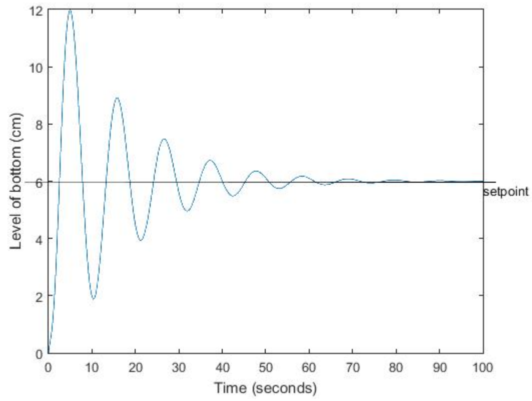

Figure 15: The response of the level of top for step change in top flow from (1 to 1.4 kmol/s) at setpoint 6 cm at PID of control structure 4.

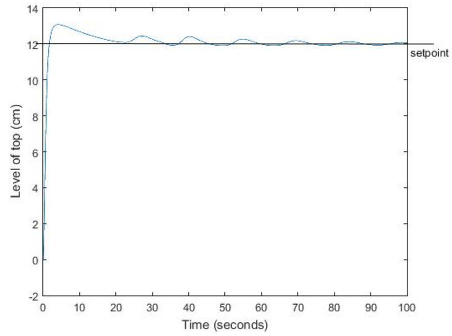

Figure 16: The response of the level of top for step change in reflux ratio from (3 to 4) at setpoint 12 cm at PID of control structure 4.

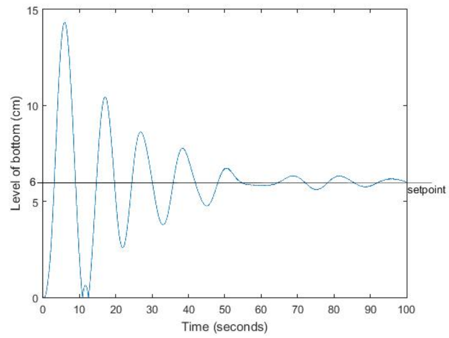

Figure 17: The response of the level of bottom for step change in reflux ratio from (3 to 4) at setpoint 6 cm at PID of control structure 4.

Run no.	Controlled variable	Variable of step change	Structure	Value of step change	IAE
1	Top level (Set point 12 cm)	Reflux ratio	4	03-Apr	30.7
2	Top level (Set point 12 cm)	Top flow rate, kmol/sec	4	1-1.4	30.07
3	Top level (Set point 12 cm)	Bottom flow rate, kmol/sec	4	1-1.4	27.77
4	Bottom level (Set point 6 cm)	Reflux ratio	4	03-Apr	136
5	Bottom level (Set point 6 cm)	Top flow rate, kmol/sec	4	1-1.4	127.7
6	Bottom level (Set point 6 cm)	Bottom flow rate, kmol/sec	4	1-1.4	127.6

Table 5: The integral absolute error (IAE) of level control.

Temperature control: In this simulation test, the top flow rate step increases from 1 kmol/sec. to 1.4 kmol/sec. In theory, the top tray temperature response should be a monotonic increase to some new steady state. However, the simulation result, shown in Figures 18-23 is at all monotonic and does appear to be reaching steady state after different time. Run 5 shown in Figure 22 after a dead time of three second, the condenser temperature rose extremely quickly to about 82.5°C, then curiously decreased and increased again, followed by another decrease as wave. Through the attempts which used to adjust the controller parameters, the most influential parameter was the integral time, whenever the integral time was increased the response and the controller performance was increased too. While the proportional gain was less influential compared to the integral time and it's found that the setting time of distillation column PID controller were less than 40 min, except run 5,6 those take taller time to arrive to steady state but shorter time to arrive to set point. Also, from the Integral absolute error (IAE) calculation for PID simulation that is given in Table 6, IAE for run 4, 5 and 6 was 24.5664, 25.1048, and 25.1543 respectively, and those give the desired performance because those take smaller time to arrive to steady state and does not appear effected in step change.

Run no.	Controlled variable	Variable of step change	Structure	Value of step change	IAE
1	Temperature of top tray (setpoint 82°C)	Bottom flow rate, kmol/sec	1	05-Jul	31.3638
2	Temperature of top tray (setpoint 82°C)	Top flow rate, kmol/sec	1	1-1.4	31.3682
3	Temperature of top tray (setpoint 82°C)	Reflux ratio	1	03-Apr	28.6983
4	Temperature of condenser (setpoint 82°C)	Reflux ratio	4	03-Apr	24.5664
5	Temperature of condenser (setpoint 82°C)	Top flow rate, kmol/sec	4	1-1.4	25.1048
6	Temperature of condenser (setpoint 82°C)	Bottom flow rate, kmol/sec	4	1-1.4	25.1543

Table 6: The integral absolute error (IAE) of temperature.

Comparison between control methods: quantitative performance IAE values for the PID in four control structures, level and temperature controllers are given in Tables 5 and 6. The comparison between control methods were made to show the close-loop performance of distillation column using the integral absolute error (IAE). The PID controller of structure 4 for level gave a good performance with an IAE range (27.7767-30.7048) for top level and (127.6333-136.0102) for bottom level. While the IAE for PID controller of structure 4 for temperature gave a good performance with an IAE rang (24.5664-25.15543) for that the best structure of control of the distillation column temperatures and level using PID is control structure 4 and that show in Figures 24 and 25 for temperature and level although we had a lower IAE in temperature in structure 2 but for level it bigger than other.

The best control structure is structure 4 it has three loop of controller, first loop controlling top level, second loop controlling temperature of condenser, and third loop controlling bottom level. In this section try to reconfiguration of structure 4 by testing of set one loop, two loops, and then three loops on distillation column.

In one loop bottom level controlling by top flow rate notice the top flow rate is still go up without control as show in Figures 24 and 26 that is unlike performance, same things happen for bottom level when control top level as show in Figures 27 and 28. The effect on temperature and mole fraction not appear when let without controller but taken more time to arrive to steady state as show in Figures 25 and 29-34 compare with those under controller. when testing put two variables control such as top level and bottom level under controller that give good performance for level as show in Figures 32 and 35 same effect explain above on temperature and mole fraction. when testing to put controller on temperature that given shorter time for temperature and mole fraction to arrive steady state.

To summarize the above results of close-loop tests, it must be noted that top and bottom level must be control, temperature response appeared shorter time to arrive steady state under control than without for that the best to controlling distillation column its single ended temperature control with use top and bottom flow rate to control top and bottom level.

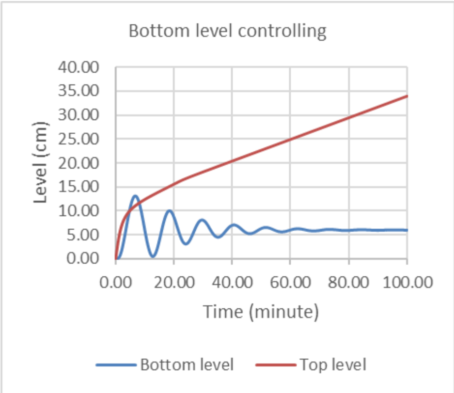

Figure 18: Comparison of the response of the level if setting by one loop controlling bottom level of PID of control structure 4.

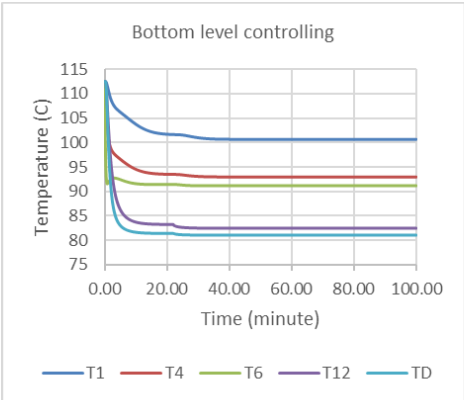

Figure 19: Comparison of the response of the temperature if setting by one loop controlling bottom level of PID of control structure 4.

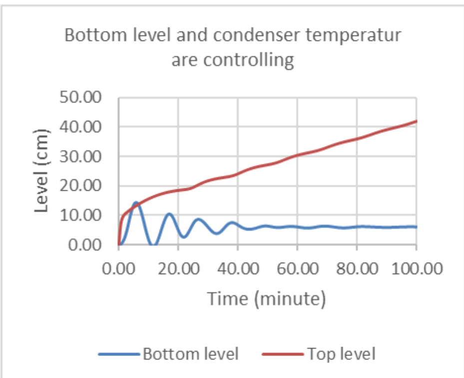

Figure 20: Comparison of the response of the level if setting by two loop controlling bottom level and condenser temperature of PID of control structure 4.

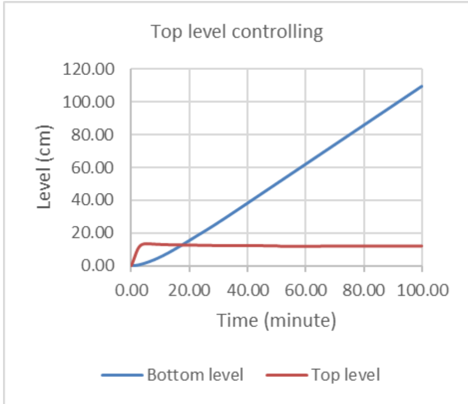

Figure 21: Comparison of the response of the level if setting by one loop controlling top level of PID of control structure 4.

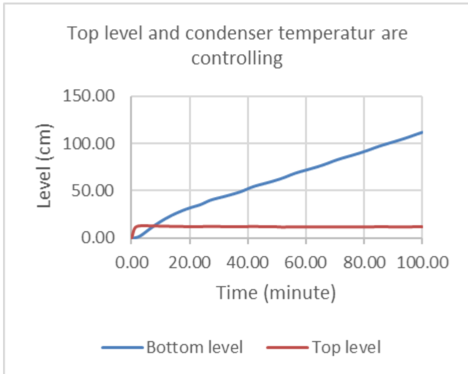

Figure 22: Comparison of the response of the level if setting by two loop controlling top level and condenser temperature of PID of control structure 4.

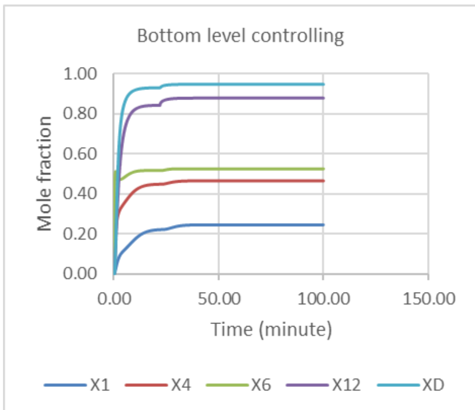

Figure 23: Comparison of the response of the mole fraction if setting by one loop controlling bottom level of PID of control structure 4.

Citation: Ahmed DF, Nawaf MY (2018) Simulation Study in Control System Configuration of a Distillation Column. J Chem Eng Process Technol 9: 388. doi:10.4172/2157-7048.1000388

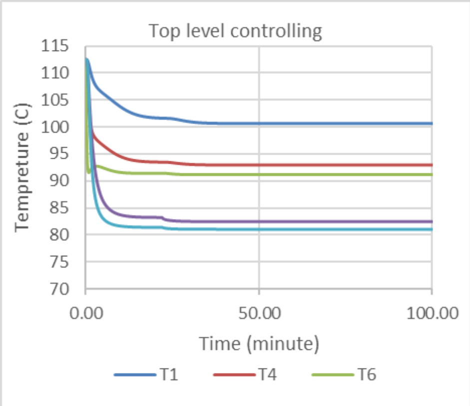

Figure 24: Comparison of the response of the temperature if setting by one loop controlling top level of PID of control structure 4.

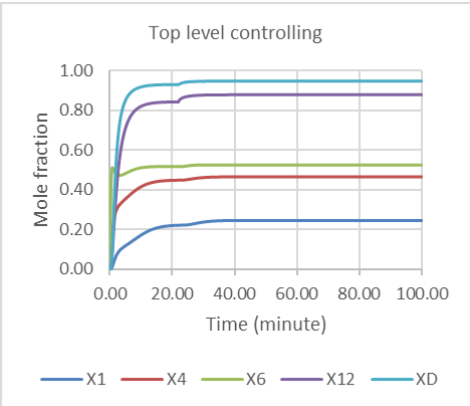

Figure 25: Comparison of the response of the mole fraction if setting by one loop controlling top level of PID of control structure 4.

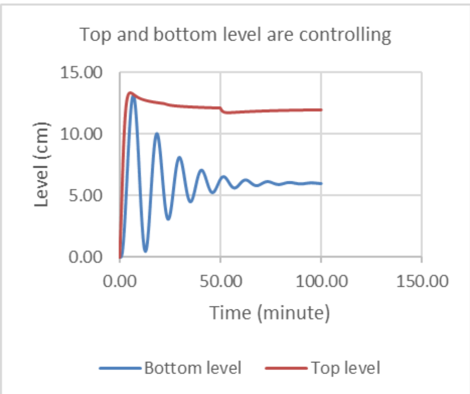

Figure 26: Comparison of the response of the level if setting by two loop controlling top and bottom level of PID of control structure 4.

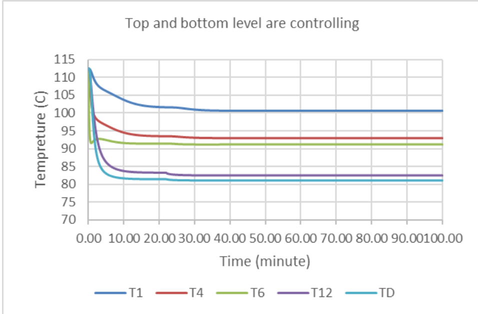

Figure 27: Comparison of the response of the temperature if setting by two loop controlling top and bottom level of PID of control structure 4.

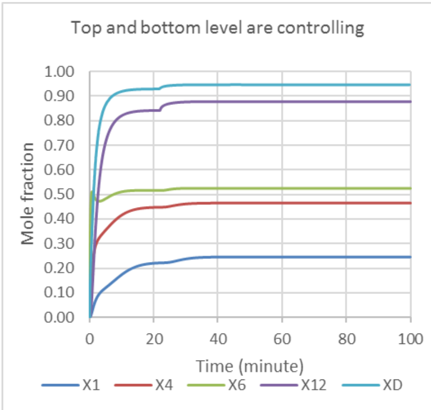

Figure 28: Comparison of the response of the mole fraction if setting by two loop controlling top and bottom level of PID of control structure 4.

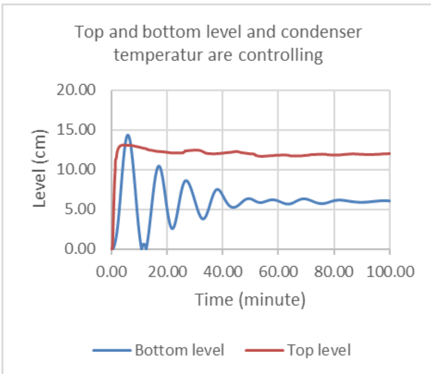

Figure 29: Comparison of the response of the level if setting by three loop controlling top and bottom level and condenser temperature of PID of control structure 4.

Citation: Ahmed DF, Nawaf MY (2018) Simulation Study in Control System Configuration of a Distillation Column. J Chem Eng Process Technol 9: 388. doi:10.4172/2157-7048.1000388

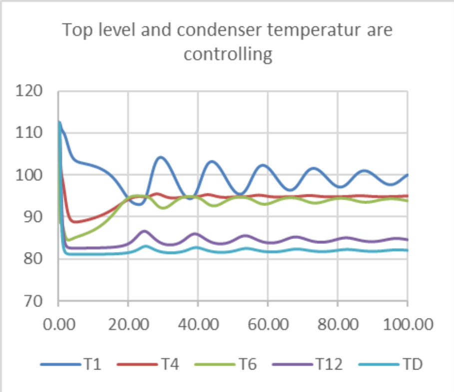

Figure 30: Comparison of the response of the temperature if setting by two loop controlling top level and condenser temperature of PID of control structure 4.

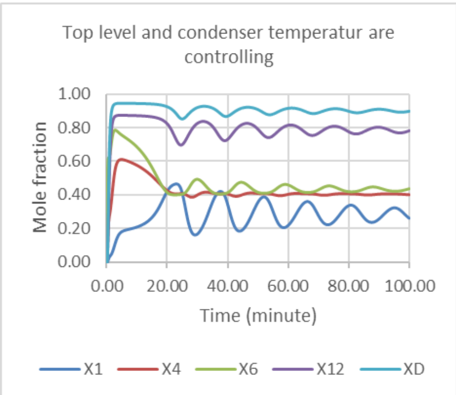

Figure 31: Comparison of the response of the mole fraction if setting by two loop controlling top level and condenser temperature of PID of control structure 4.

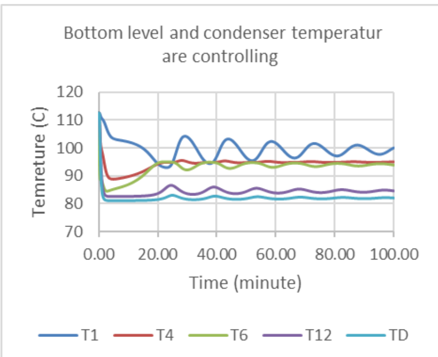

Figure 32: Comparison of the response of the temperature if setting by two loop controlling bottom level and condenser temperature of PID of control structure 4.

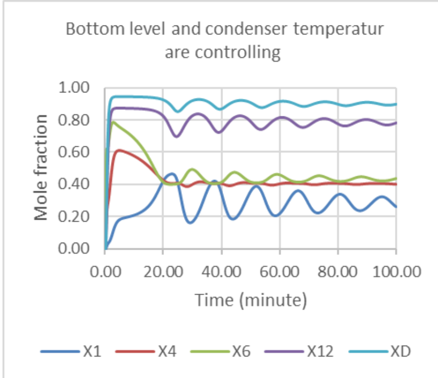

Figure 33: Comparison of the response of the mole fraction if setting by two loop controlling bottom level and condenser temperature of PID of control structure 4.

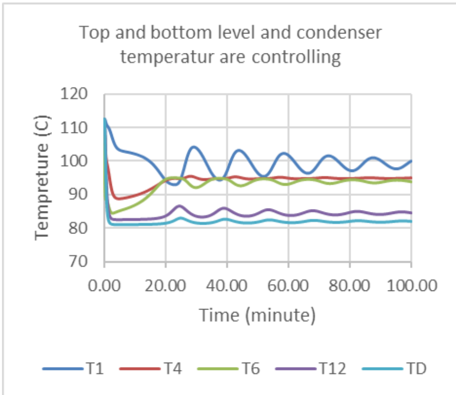

Figure 34: Comparison of the response of the temperature if setting by three loop controlling top and bottom level and condenser temperature of PID of control structure 4.

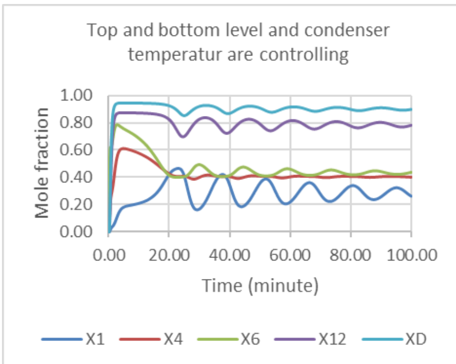

Figure 35: Comparison of the response of the mole fraction if setting by three loop controlling top and bottom level and condenser temperature of PID of control structure 4.

Conclusion

The primary purpose of this research is to choose the control structures of distillation columns. a non-linear dynamic distillation model is built which included level of both the condenser(top) and reboiler(bottom) and a dynamic model of distillation column.

The dynamic studies are made of open-loop responses to step change in feed flowrate, composition of feed and reflux ratio. Next, dynamic studies of the column equipped with the subject multi-loop control system were made to determine the closed-loop behavior of the system and the best structure of controller. Simulating systems in which dynamic model need to be studied over a wide range of operation, the model presented in this research would be very valuable. The inverse and highly non-linear responses shown in the open-loop response curves foreboded the possible problems that might be encountered when the control system studied here is implemented. In fact, the inverse behavior did surface when control runs are made. The ratio of the benzene-toluene ratio is changed, the effect on column temperature is greater than the effect of changing the feed rate itself. had the column been separating two key components with closer boiling points, this effect of composition on bubble point would have been much less, and the temperature control loop might have been less interacting with the composition control loop. The performance control loop of the distillation column is analyzed using IAE. Four control structure are compared. The results of control structure 4 controller are compared with other methods. It is shown that the structure 4 controller provide better performance compared to the other controllers because the structure controller has lower time to reach steady state value and the response much more stable.

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