



Modeling and Control of Binary Distillation Column

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ABSTRACT: Distillation is a continuous process in the petroleum and the chemical industries. A high amount of energy used in the chemical and petrochemical industries is consumed in the distillation process, since large amount of heat transfer takes place between the trays of distillation column. Thus, there is a need of energy efficient process to be designed and controlled, in order to make the final and the intermediate products of the distillation process of desired quality and the whole process being more economic. In the present study, graphical programming of LabVIEW software has been utilized to develop model for distillation column and also to control the product stream composition. Results from the simulation are helpful for initial steps of a petroleum project feasibility study and design.

KEYWORDS: Modeling, Binary Distillation Column, Controllers, Simulation.

I. INTRODUCTION

Distillation column plays a important role in petro -chemical and refinery industries. Distillation is a process in which a liquid or vapour mixture of two or more substances is separated into its component fractions of desired purity, by the application and removal of heat. Binary distillation column is used to separate two components (Component A & component B) . Example for Binary mixture is methanol and water. Fig 1.1 shows the schematic diagram of distillation column. Distillation column generally made up of tray of tower/column. Tray types have more efficient than other type. The distillation column consists of many trays, condenser, reboiler, and a vertical column for the separation purpose. In this paper, distillation is a continuous process. It means that the feed is given in continuous manner. Here feed is the input to the column. Distillation column is separated into two sections. They are stripping section and rectification section. The trays above the feed tray is called stripping section.

The binary mixture is fed into the feed tray with flow rate F_f and a molar fraction of component A, C_f . The overhead product is completely condensed using condenser and it is allowed to flow in the reflux drum. The liquid in the reflux drum partly pumped back to the top tray F_R and it is partially removed as distillate product with molar flow rate of F_D . let us consider M_{RD} is the liquid hold up in the reflux drum and X_D is the molar fraction of component A of liquid in the reflux drum. At the bottom of the column , a liquid stream is removed with flow rate of F_B and composition of X_B . A liquid stream with molar flow rate of v is also drawn from the bottom of the column and after that it has been heated in the reboiler, it returns back to the base of the column. The composition of the column base is X_B . Let M_B is the liquid holdup in the base of the column. The column contains N trays numbered from bottom of the column to top of the column.

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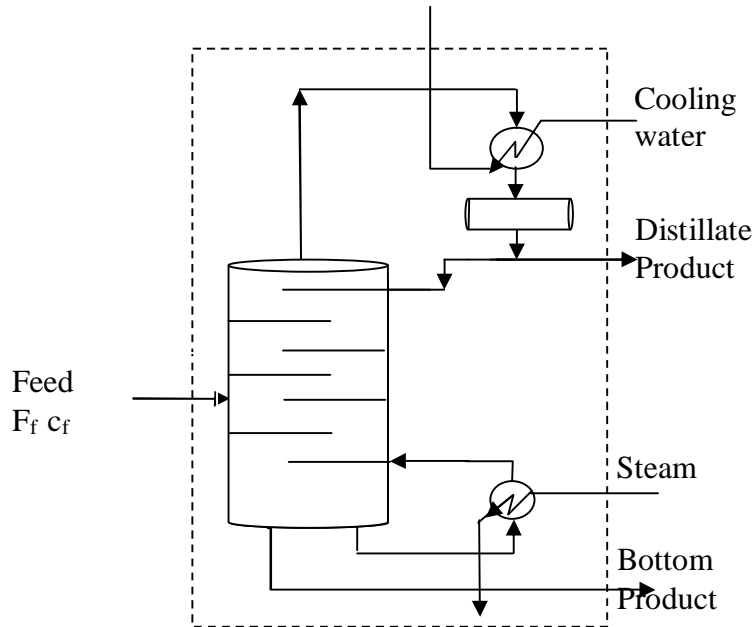


Fig. 1 Schematic diagram of binary distillation column

II. MATHEMATICAL MODELING

Distillation column, as described earlier is a long tower comprising of several components, which are responsible for the energy and mass transfer, such as

(i) the simple vapour liquid equilibrium is used,

$$Y_i = \alpha \frac{X_i}{1 + (\alpha - 1)X_i} \quad \text{---(2.1)}$$

Where,

X_i = Liquid composition on the nth tray (mole fraction more volatile component)

Y_i = Vapour composition on the nth tray (more fraction more volatile component)

α = Relative volatility.

(ii) $V_1 = V_2 = V_3 = \dots = V_N$ --- (2.2)

These vapour holdups are equal only at the steady state and are not necessarily constant with the time. The vapour boil up can be manipulated dynamically. The mathematical effect of assuming equimolal overflow is that, modeling doesn't need an energy equation for each tray.

(iii) The liquid rates is not same throughout the column, they will depend on the fluid dynamics of the tray. A simple Francis Weir formula relationship has been used to relate the liquid holdup on the tray (M_i) to the liquid flow rate leaving the tray (L_i).

$$L_i = f(M_i) \quad \text{--- (2.3)}$$



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Where,

M_i = Molar flow rate of I th column

L_i = Liquid hold up in the I th tray

2.2.1 DIFFERENTIAL EQUATIONS RELATED TO DIFFERENTIAL COLUMN

Taking all the above assumptions, now the equations which describe the system can be stated as,

(i) Condenser and Reflux Drum:

Component A:

$$\frac{d(M_{RD} X_D)}{dt} = V_N Y_N - (F_R + F_D) X_D \quad - (2.4)$$

(ii) Top Tray (i = N):

Component A:

$$\frac{d(M_N X_N)}{dt} = F_R X_D + V_{N-1} Y_{N-1} - L_N X_N - V_N Y_N \quad -$$

(2.5)

(iii) i th Tray:

Component A:

$$\frac{d(M_i X_i)}{dt} = L_{i+1} X_{i+1} + V_{i-1} Y_{i-1} - L_i X_i - V_i Y_i \quad -$$

(2.6)

(iv) Feed Tray (at i = f):

Component A:

$$\frac{d(M_f)}{dt} = F_f C_f + L_{f+1} X_{f+1} + V_{f-1} Y_{f-1} - L_f X_f - V_f Y_f \quad -(2.7)$$

(v) Reboiler & Column Base:

Component A:

$$\frac{d(M_B X_B)}{dt} = L_1 X_1 - V Y_B - F_B X_B \quad - (2.8)$$

Based on equations 2.1-2.8, the mathematical model of the distillation column has been designed in the LabVIEW. Top product (Distillate) composition and the bottom product composition are the output i.e. Controlled variables of the plant system, and the manipulate variables are the Reflux Rate (L) and the Vapour Boilup Rate (V).

MODEL PARAMETERS AND NOMINAL OPERATING POINT

The values of the model parameters used in this dissertation are as follows in table 2.1, [4]
Table 2.1. Distillation Column Model Parameters.

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Parameter	Value
No of tray	20
Feed tray	10
Relative volatility	1.5
Feed composition, mole fraction	0.5
Feed flow rate	1 kol/mol
Distillate product composition, mole fraction	0.99
Bottom product composition, mole fraction	0.01
Reflux rate, Kmol/min	2.7063
Boil up rate, Kmol/min	3.206
Liquid holdup in condenser	0.50
Liquid holdup in the reboiler	0.50

III. CONTROLLER

A controller generates the control signal based on the control strategy employed and the input variable. The input variable to a controller is the error i.e. difference between the reference input and output. In this paper, it deals with feedback controller such as, P, PI, PID controllers are used to obtain the desired quality of distillate and bottom products.

IV. SIMULATION DIAGRAM

Different parts/subsystems and their block diagram programming, are as following,

4.1 Ith TRAY

The equation of the n^{th} tray has two inputs i.e. x_{i+1} and x_{i-1} and one output i.e. x_i . In addition to that a vapour liquid equilibrium has also been used to calculate the value of y_i . This value of y_i is being used to solve equation further. Fig 4.1 shows the block diagram of i^{th} tray.

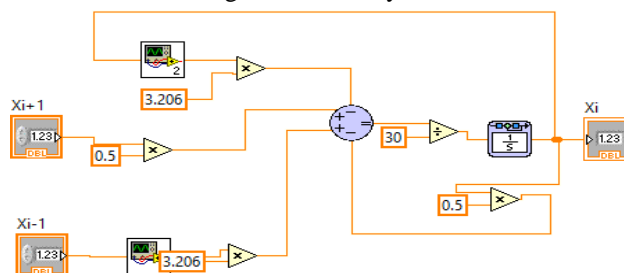


Fig 4.1 Block Diagram of i^{th} Tray

4.2 FEED TRAY

The equation of the Feed tray has four inputs i.e. x_{i+1} , x_{i-1} , x_f and z_f and one output. Fig 4.2 shows the block diagram of feed tray.

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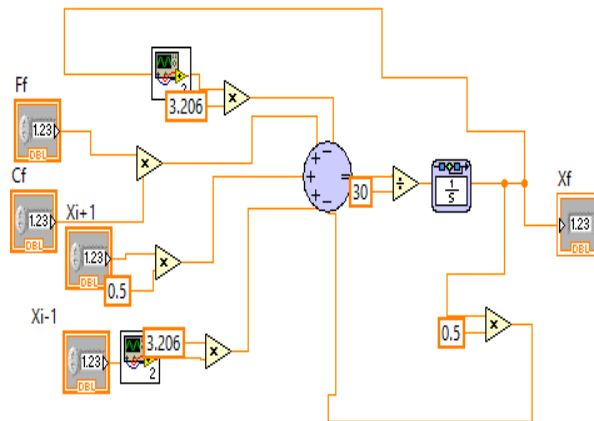


Fig 4.2 Block Diagram of Feed Tray

4.3 CONDENSER AND REFLUX DRUM

The equation for the condenser and reflux drum are not iterative, hence it is not made as subsystem and are used directly in the block diagram. Fig 4.3 shows the block diagram of reflux drum.

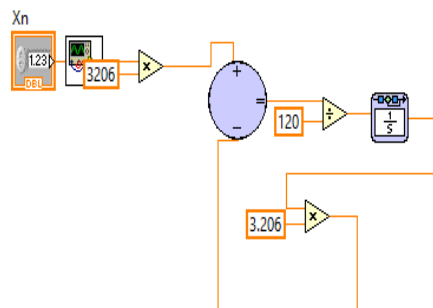


Fig. 4.3 Block Diagram of Reflux Drum

V. RESULT AND DISCUSSION

The model for Distillation Column is linearized around the point [99.25%, .10%] i.e. the distillate composition and the bottom product composition respectively. A step change of magnitude [.5 %] in the composition has been fed to the system and the variation in the composition of the top and the bottom product has been analyzed. Different combinations of the values for K_p , T_I , T_D has been used to observe the variation of the composition of the products and the variations are showed in fig 5.1 - fig 5.3 .

5.1 CONTROL STRATEGIES

Case 1: Simple Proportional Controller:
 $KP1 = 49.095, KP2 = 32.183$

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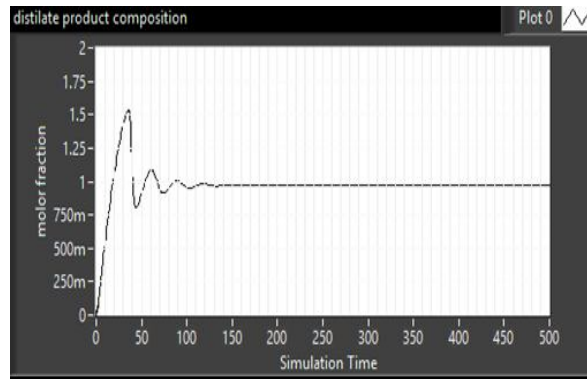


Fig. 5.1 Variation of Distillate Composition v/s time(P control)

Case 2: Proportional Integral Control:

$KP1 = 24.7486$ $Ti 1 = 9.24609$, $KP2 = 28.1027$ $Ti 2 = 7.3494$

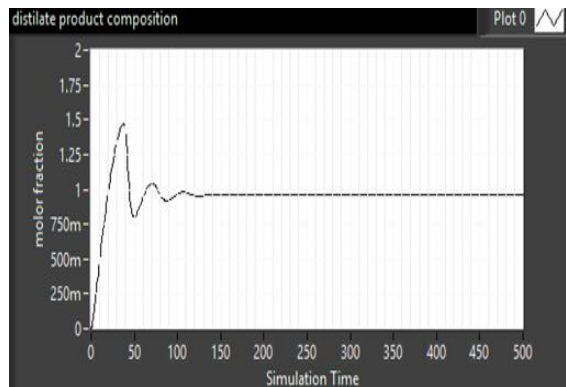


Fig. 5.2 Variation of Distillate v/s Time (PI Control)

Case 3: Proportional Integral Derivative Control:

$KP1 = 15.3594$ $Ti 1 = 6.54891$ $Td 1 = .1891$

$KP2 = 21.848$ $Ti 2 = 4.363$ $Td 1 = .1857$

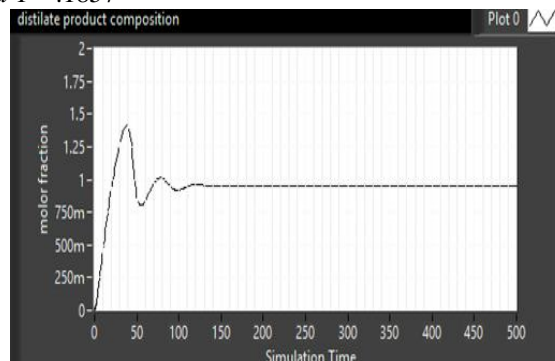


Fig. 5.3 Variation of Distillate v/s Time (PID Control)



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Case I: In this strategy only proportional controller is applied and the Steady state error found to be very large, thus the controller was not acceptable.

Case II: In this a PI controller is employed, although steady state error came down as integral controller has been added, but the Settling time and overshoot are still not acceptable, thus the PI controller with the said values of parameters is not acceptable.

Case III: In this case, PID controller with tight tuning of the parameters is used. Steady state error has been removed by application of the integral controller and the transient response has been improved with the application of the derivative controller and the peak overshoot will be reduced. The output response has less oscillation. It can be acceptable.

VI. CONCLUSION

The procedure has been introduced to build up a mathematical model and simulation for a binary distillation column based on the energy balance ($L-V$) equations. By comparing the simulation results of various controllers, PID controller can be acceptable, since the Steady state error has been removed by application of the integral controller and the transient response has been improved with the application of the derivative controller. Hence PID controller is better than other to obtain the desired distillate product.

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