International Journal of Research in Advanced Engineering and Technology
ISSN: 2455-0876; Impact Factor: RJIF 5.44
www.engineeringresearchjournal.com
Volume 2; Issue 3; May 2016; Page No. 12-17

Dynamics and control of binary distillation column

Ashwini K, Vimalnathan

Department of Chemical Engineering, Adhiyamaan College of Engineering, Anna University, Hosur, Tamil Nadu, India.

Abstract

The simulation of open loop response of binary distillation column is one of the major fields of study in present industrial revolution. Control of column is a challenging problem and has received considerable attention because of its importance in the process industry. This study provides the information about the simulation of open loop response in binary distillation column which consists of two component mixtures for different plates using Runge-Kutta method in excel. Next, we controlled the top tray(x2) and bottom tray(x10) of tower keeping L and V as the manipulated variable and by disturbing the feed composition we brought back to its original steady state level. The controllers used here is proportional integral derivative controller. We also set the delay of two hours and span of control for the distillation column.

Keywords: Binary distillation**,** Open loop response, Controlling**,** Proportional Integral Derivative controller

Introduction

A distillation is the separation or partial separation of a liquid feed mixture into components or fractions by selective boiling (or evaporation) and condensation. A distillation produces at least two output fractions. This fraction include at least one volatile distillate fraction, which has boiled and been separately captured as a vapor condensed to a liquid, and practically always a bottoms (or residuum) fraction, which is the least volatile residue that has not been separately captured as a condensed vapor. Distillation is a unit operation, or a physical separation process, and not a chemical reaction.

Methods

Batch distillation

Batch distillation refers to the use of distillation in batches, meaning that a mixture is distilled to separate it into its component fractions before the distillation still is again charged with more mixture and the process is repeated. This is in contrast with continuous distillation where the feedstock is added and the distillate drawn off without interruption. Batch distillation has always been an important part of the production of seasonal, or low capacity and high-purity chemicals. It is a very frequent separation process in pharmaceutical industry and in wastewater treatment units.

Continuous distillation

Continuous distillation is an ongoing distillation in which a liquid mixture is continuously (without interruption) fed into the process and separated fractions are removed continuously as output streams as time passes during the operation. Continuous distillation differs from batch distillation in the respect that concentrations should not change over time. Continuous distillation can be run at a steady state for an arbitrary amount of time. For any source material of specific composition, the main variables that affect the purity of products in continuous distillation are the reflux ratio and the number of theoretical equilibrium stages (practically, the number of trays or the height of packing). Reflux is a flow from the condenser back to the column, which generates a recycle that allows a better separation with a given number of trays. Equilibrium stages are ideal steps where compositions achieve vapor-liquid equilibrium, repeating the separation process and allowing better separation given a reflux ratio. A column with a high reflux ratio may have fewer stages, but it refluxes a large amount of liquid, giving a wide column with a large holdup. Conversely, a column with a low reflux ratio must have a large number of stages, thus requiring a taller column.

Binary Distillation

A binary distillation column has two component mixtures with different no of trays plus reboiler and condenser. Two variants of this system are:

An "uncontrolled" column with level controllers for condenser and reboiler, but with no temperature or composition control. The reflux *L* and the boil-up *V* remain as degrees of freedom.

A "controlled" column with an additional composition controller in the lower column part that manipulates the boil-up rate *V*. The uncontrolled system is used to explain the system equations and the reduction procedure. The major modeling assumptions are: Ideal trays, which means that liquid and vapor are in equilibrium at each tray; ideal mixture, which means that the vapor composition *y* can be expressed as a function of the liquid composition *x* assuming the constant relative volatility

$$y = k(x) = \frac{\alpha x}{[1+(\alpha-1)x]}$$

where, α is the relative volatility, x is the mole fraction; constant molar flows, which means that the energy balance is simplified; constant molar hold upon each tray and negligible mass in the vapor phase.

The column has one feed flow *F* at tray number nF. zF denotes the concentration of the first (light) component in the feed. A liquid flow *L* (or L + F for trays below the feed tray) and a vapor flow *V* enter and leave each tray. The condenser and reboiler levels are assumed to be controlled using the distillate D and bottom flow *B*, respectively. For simplicity, perfect level control is assumed, such that D = V − L and B = L + F – V. The concentrations in these flows determine the

purity of the distillation products and are therefore the most important output variables in the process. The feed flow rate *F* and the feed concentration zF can be seen as disturbance variables, and the flows *L* and *V* are manipulated variables.
The main process in binary distillation is that separates a feed mixture stream to two fractions: one distillate and one bottoms fractions (which is known as residue).

Runge-Kutta method
In numerical analysis, the Runge-Kutta methods are an important family of implicit and explicit iterative methods for the approximation of solutions of ordinary differential equations. These techniques were developed around 1900 by the German mathematicians C. Runge and M.W. Kutta.

Fourth order Runge-Kutta method
This method is simply a higher order approximation to the midpoint method. Instead of shooting to the midpoint, estimating the derivative, the shooting across the entire interval the Runge-Kutta method, in a sense takes, four steps, shooting across one quarter of the interval, estimating the derivative, and then shooting to the midpoint, and so on. The precise manner in which the method propagates across a time step is done in the optimal way for the four steps. We will not provide a formal derivation of the Runge-Kutta algorithm; instead we will present the method and implement it. The general system of ODEs can be written as,

$$\frac{dx}{dt} = f(t,x)$$

The Runge-Kutta method is defined as:

$$K_1 = \Delta t \ f(t_n, x_n)$$

$$K_2 = \Delta t \ f\left(t_n + \frac{\Delta t}{2}, x_n + \frac{K_1}{2}\right)$$

$$K_3 = \Delta t \ f\left(t_n + \frac{\Delta t}{2}, x_n + \frac{K_2}{2}\right)$$

$$K_4 = \Delta t \ f(t_n + \Delta t, x_n + K_3)$$

$$K = \frac{1}{6}(K_1 + 2K_2 + 2K_3 + K_4) \quad ; \quad x_{n+1} = x_n + K$$

Controllers
The mode of control is the manner in which a control system makes corrections relative to an error that exists between the desired value (set point) of a controlled variable and its actual value. The mode of control used for a specific application depends on the characteristics of the process being controlled. For example, some processes can be operated over a wide band, while others must be maintained very close to the set point. Also, some processes change relatively slowly, while others change almost immediately. The following four types of controllers are widely used in chemical process control applications.

Proportional Integral Derivative controller
The PID controller, which consists of proportional, integral and derivative elements, is widely used in feedback control of industrial processes. In applying PID controllers, engineers must design the control system: that is, they must first decide which action mode to choose and then adjust the parameters of the controller so that their control problems are solved appropriately. To that end, they need to know the characteristics of the process. As the basis for the design procedure, they must have certain criteria to evaluate the performance of the control system.
Algorithm of PID controller

$$x(t) = Bias + K\left(e(t) + \frac{1}{\tau_i}\int_0^t e(t)dt + \tau_D\frac{de(t)}{dt}\right)$$

When used in this way, the three elements of the PID controller produce outputs with the following nature:
- P element: proportional to the error at the instant *t*, which is the “present” error.
- I element: proportional to the integral of the error up to the instant *t*, which can be interpreted as the accumulation of the “past” error.
- D element: proportional to the derivative of the error at the instant *t*, which can be interpreted as the prediction of the “future” error.

Problem
The binary distillation column problem is given below:

$$\frac{dx_1}{dt} = \left(\frac{V}{H_1}\right)(y_2 - x_1)$$

$$\frac{dx_i}{dt} = \left(\frac{1}{H_i}\right)[-V\emptyset(y_i^* - y_{i+1}) + L(x_{i-1} - x_i)]$$

for i=2,3,…,f-1

$$\frac{dx_f}{dt} = \left(\frac{1}{H_f}\right)[-V\emptyset(y_f^* - y_{f+1}) + Lx_{f-1} - (L+F)x_f + Fz]$$

$$\frac{dx_j}{dt} = \left(\frac{1}{H_j}\right)[-V\emptyset(y_j^* - y_{j+1}) + (L+F)(x_{j-1} - x_j)]$$

For j=f+1,….,N-1

$$\frac{dx_N}{dt} = \left(\frac{1}{H_N}\right)[-V\emptyset y_N^* + (L+F)x_{N-1} - (L+F-V)x_N]$$

Where,

$$y_i^* = \frac{(\alpha x_i)}{[1+(\alpha-1)x_i]} \quad \text{For i=2,3,…,N}$$

$$y_i = y_{i+1} + \emptyset(y_i^* - y_{i+1}) \quad \text{For i=2,3,….,N-1}$$

$$y_N = \emptyset y_N^*$$

The stage are numbered from the top with the condenser as 1 and the reboiler as stage N. Here H, α, ∅, z and F are the hold –up, relative volatility, Murphree stage efficiency, feed composition and feed-flow rate respectively.

Operating conditions

Table 1: Operating conditions

H1	100mol	α	3
H2	25mol	∅	1
H3	25mol	z	0.5
H4	25mol	F	10 mol/h
H5	25mol	D	5 mol/h
H6	25mol	V	36.5 mol/h
H7	25mol	L	31.5 mol/h
H8	25mol		
H9	25mol		
H10	200mol		

Methodology

Simulation of open loop response

Finding of mole fraction values

In this problem the first step is to find the mole fraction values in steady state condition. To find these values we first assumed some rough values for the mole fraction from x1 to x10.

Then, we found out the mole fraction of vapour by using the below formulas. It is given by,

$$y_i^* = \frac{(\alpha x_i)}{[1+(\alpha-1)x_i]} \quad \text{For i=2,3,…,N}$$

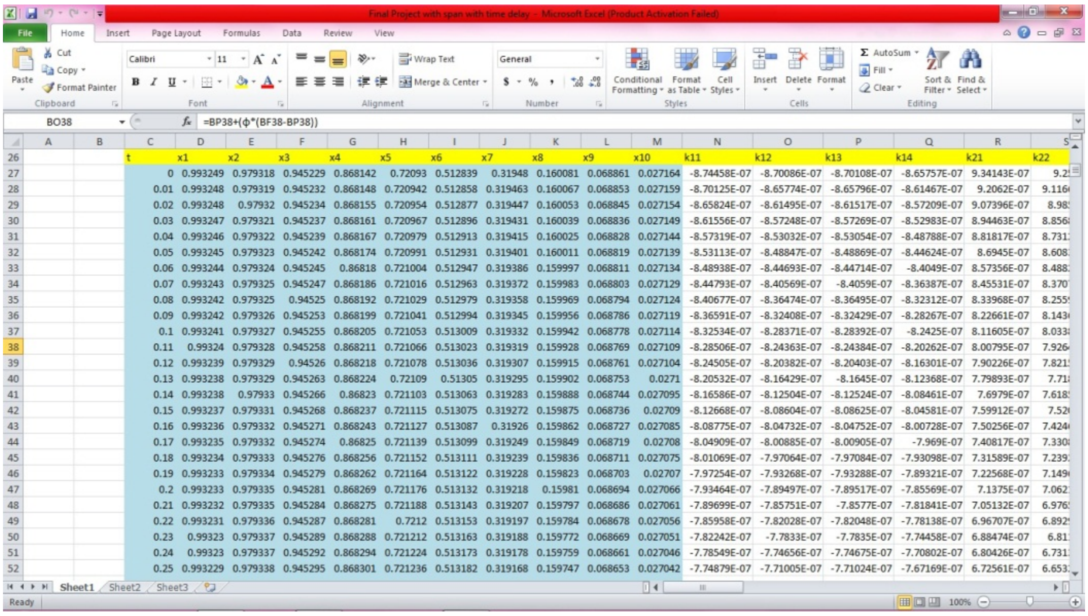
$$y_i = y_{i+1} + \emptyset(y_i^* - y_{i+1}) \quad \text{For i=2,3,….,N-1}$$

$$y_N = \emptyset y_N^*$$

By substituting these values in the differential equation which is given above in the problem we found out the mole fraction of liquid in steady state conditions by solving these numerical equations in MS excel sheet.

Table 2: steady state mole fraction values of liquid

x1	0.993249
x2	0.979318
x3	0.945229
x4	0.868142
x5	0.720931
x6	0.512839
x7	0.319487
x8	0.160081
x9	0.068861
x10	0.027164



BO38 =BP38+(ϕ*(BF38-BP38))

	t	x1	x2	x3	x4	x5	x6	x7	x8	x9	x10
27	0	0.993249	0.979318	0.945229	0.868142	0.72093	0.512839	0.31948	0.160081	0.068861	0.027164
28	0.01	0.993248	0.979319	0.945232	0.868148	0.720942	0.512858	0.319463	0.160067	0.068853	0.027159
29	0.02	0.993248	0.97932	0.945234	0.868155	0.720954	0.512877	0.319447	0.160053	0.068845	0.027154
30	0.03	0.993247	0.979321	0.945237	0.868161	0.720967	0.512896	0.319431	0.160039	0.068836	0.027149
31	0.04	0.993246	0.979322	0.945239	0.868167	0.720979	0.512913	0.319415	0.160025	0.068828	0.027144
32	0.05	0.993245	0.979323	0.945242	0.868174	0.720991	0.512931	0.319401	0.160011	0.068819	0.027139
33	0.06	0.993244	0.979324	0.945245	0.86818	0.721004	0.512947	0.319386	0.159997	0.068811	0.027134
34	0.07	0.993243	0.979325	0.945247	0.868186	0.721016	0.512963	0.319372	0.159983	0.068803	0.027129
35	0.08	0.993242	0.979325	0.94525	0.868192	0.721029	0.512979	0.319358	0.159969	0.068794	0.027124
36	0.09	0.993242	0.979326	0.945253	0.868199	0.721041	0.512994	0.319345	0.159956	0.068786	0.027119
37	0.1	0.993241	0.979327	0.945255	0.868205	0.721053	0.513009	0.319332	0.159942	0.068778	0.027114
38	0.11	0.99324	0.979328	0.945258	0.868211	0.721066	0.513023	0.319319	0.159928	0.068769	0.027109
39	0.12	0.993239	0.979329	0.94526	0.868218	0.721078	0.513036	0.319307	0.159915	0.068761	0.027104
40	0.13	0.993238	0.979329	0.945263	0.868224	0.72109	0.51305	0.319295	0.159902	0.068753	0.0271
41	0.14	0.993238	0.97933	0.945266	0.86823	0.721103	0.513063	0.319283	0.159888	0.068744	0.027095
42	0.15	0.993237	0.979331	0.945268	0.868237	0.721115	0.513075	0.319272	0.159875	0.068736	0.02709
43	0.16	0.993236	0.979332	0.945271	0.868243	0.721127	0.513087	0.31926	0.159862	0.068727	0.027085
44	0.17	0.993235	0.979332	0.945274	0.86825	0.721139	0.513099	0.319249	0.159849	0.068719	0.02708
45	0.18	0.993234	0.979333	0.945276	0.868256	0.721152	0.513111	0.319239	0.159836	0.068711	0.027075
46	0.19	0.993233	0.979334	0.945279	0.868262	0.721164	0.513122	0.319228	0.159823	0.068703	0.02707
47	0.2	0.993233	0.979335	0.945281	0.868269	0.721176	0.513132	0.319218	0.15981	0.068694	0.027066
48	0.21	0.993232	0.979335	0.945284	0.868275	0.721188	0.513143	0.319207	0.159797	0.068686	0.027061
49	0.22	0.993231	0.979336	0.945287	0.868281	0.7212	0.513153	0.319197	0.159784	0.068678	0.027056
50	0.23	0.99323	0.979337	0.945289	0.868288	0.721212	0.513163	0.319188	0.159772	0.068669	0.027051
51	0.24	0.99323	0.979337	0.945292	0.868294	0.721224	0.513173	0.319178	0.159759	0.068661	0.027046
52	0.25	0.993229	0.979338	0.945295	0.868301	0.721236	0.513182	0.319168	0.159747	0.068653	0.027042

	k11	k12	k13	k14	k21	k22
27	-8.74458E-07	-8.70086E-07	-8.70108E-07	-8.65757E-07	9.34143E-07	9.2
28	-8.70125E-07	-8.65774E-07	-8.65796E-07	-8.61467E-07	9.2062E-07	9.116
29	-8.65824E-07	-8.61495E-07	-8.61517E-07	-8.57209E-07	9.07396E-07	8.98
30	-8.61556E-07	-8.57248E-07	-8.57269E-07	-8.52983E-07	8.94463E-07	8.856
31	-8.57319E-07	-8.53032E-07	-8.53054E-07	-8.48788E-07	8.81817E-07	8.731
32	-8.53113E-07	-8.48847E-07	-8.48869E-07	-8.44624E-07	8.6945E-07	8.608
33	-8.48938E-07	-8.44693E-07	-8.44714E-07	-8.4049E-07	8.57356E-07	8.488
34	-8.44793E-07	-8.40569E-07	-8.4059E-07	-8.36387E-07	8.45531E-07	8.370
35	-8.40677E-07	-8.36474E-07	-8.36495E-07	-8.32312E-07	8.33968E-07	8.255
36	-8.36591E-07	-8.32408E-07	-8.32429E-07	-8.28267E-07	8.22661E-07	8.143
37	-8.32534E-07	-8.28371E-07	-8.28392E-07	-8.2425E-07	8.11605E-07	8.033
38	-8.28506E-07	-8.24363E-07	-8.24384E-07	-8.20262E-07	8.00795E-07	7.926
39	-8.24505E-07	-8.20382E-07	-8.20403E-07	-8.16301E-07	7.90226E-07	7.821
40	-8.20532E-07	-8.16429E-07	-8.1645E-07	-8.12368E-07	7.79893E-07	7.71
41	-8.16586E-07	-8.12504E-07	-8.12524E-07	-8.08461E-07	7.6979E-07	7.618
42	-8.12668E-07	-8.08604E-07	-8.08625E-07	-8.04581E-07	7.59912E-07	7.52
43	-8.08775E-07	-8.04732E-07	-8.04752E-07	-8.00728E-07	7.50256E-07	7.424
44	-8.04909E-07	-8.00885E-07	-8.00905E-07	-7.969E-07	7.40817E-07	7.330
45	-8.01069E-07	-7.97064E-07	-7.97084E-07	-7.93098E-07	7.31589E-07	7.239
46	-7.97254E-07	-7.93268E-07	-7.93288E-07	-7.89321E-07	7.22568E-07	7.149
47	-7.93464E-07	-7.89497E-07	-7.89517E-07	-7.85569E-07	7.1375E-07	7.062
48	-7.89699E-07	-7.85751E-07	-7.8577E-07	-7.81841E-07	7.05132E-07	6.976
49	-7.85958E-07	-7.82028E-07	-7.82048E-07	-7.78138E-07	6.96707E-07	6.892
50	-7.82242E-07	-7.7833E-07	-7.7835E-07	-7.74458E-07	6.88474E-07	6.81
51	-7.78549E-07	-7.74656E-07	-7.74675E-07	-7.70802E-07	6.80426E-07	6.731
52	-7.74879E-07	-7.71005E-07	-7.71024E-07	-7.67169E-07	6.72561E-07	6.653

Fig 1: MS excel spread sheet for mole fraction values

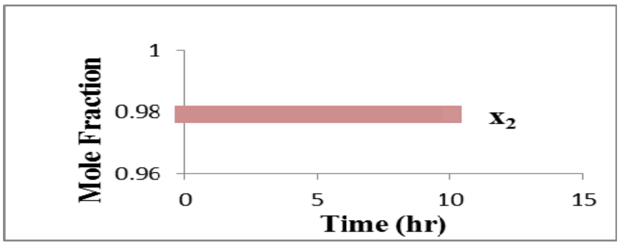


Fig 2: Steady state top tray graph

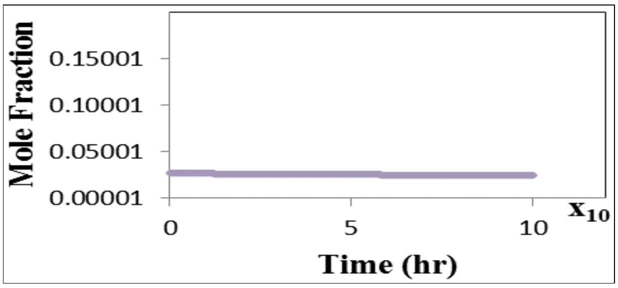


Fig: 3 Steady state bottom tray graph

Finding out 'k' values for different plates of tower

The k values is found out numerically for ten plates of the distillation column using fourth order Runge-Kutta method in MS excel.

The fourth order Runge-Kutta formula is given by

$$k_1 = hf\,(x_n, y_n)$$

$$k_2 = hf\left(x_n + \frac{h}{2}, y_n + \frac{k_1}{2}\right)$$

$$k_3 = hf\left(x_n + \frac{h}{2}, y_n + \frac{k_2}{2}\right)$$

$$k_4 = hf\,(x_n + h, y_n + k_3)$$

$$k = \frac{1}{6}(k_1 + 2k_2 + 2k_3 + k_4)$$

The ten differential equations which is used to find out k values are,

$$\frac{dx_1}{dt} = \left(\frac{V}{H1}\right)[y_2 - x_1]$$

$$\frac{dx_2}{dt} = \left(\frac{1}{H_2}\right)[-V\emptyset(y_2^* - y_3) + L(x_1 - x_2)]$$

$$\frac{dx_3}{dt} = \left(\frac{1}{H_3}\right)[-V\emptyset(y_3^* - y_4) + L(x_2 - x_3)]$$

$$\frac{dx_4}{dt} = \left(\frac{1}{H_4}\right)[-V\emptyset(y_4^* - y_5) + L(x_3 - x_4)]$$

$$\frac{dx_5}{dt} = \left(\frac{1}{H_5}\right)[-V\emptyset(y_5^* - y_6) + L(x_4 - x_5)]$$

$$\frac{dx_6}{dt} = \left(\frac{1}{H_6}\right)[-V\emptyset(y_6^* - y_7) + Lx_5 - (L+F)x_6 + Fz]$$

$$\frac{dx_7}{dt} = \left(\frac{1}{H_7}\right)[-V\emptyset(y_7^* - y_8) + (L+F)(x_6 - x_7)]$$

$$\frac{dx_8}{dt} = \left(\frac{1}{H_8}\right)[-V\emptyset(y_8^* - y_9) + (L+F)(x_7 - x_8)]$$

$$\frac{dx_9}{dt} = \left(\frac{1}{H_9}\right)[-V\emptyset(y_9^* - y_{10}) + (L+F)(x_8 - x_9)]$$

$$\frac{dx_{10}}{dt} = \left(\frac{1}{H_{10}}\right)[-V\emptyset y_{10}^* + (L+F)x_9 - (L+F-V)x_{10}]$$

By solving these differential equations we found out the k values for different plates of the tower.

Thus, we simulated the open loop response of binary distillation column in excel.

3.2 Control of top and bottom tray

Next, is the control of top and bottom tray of the tower by disturbing the feed composition from z = 0.7 to 0.3 using proportional integral derivative controller in excel.

Here, we use L and V as the manipulated variable for top and bottom tray of the distillation column. The controller formula is given by,

$L = L_{max}$ co

$$Co = 0.5 + k_c\left(E + 1/\tau_i \int_0^t E\,dt + \tau_d dE/dt\right)$$

We also set the delay of two hours and span of control to both the trays of binary distillation column. The distillation column controller block diagram for top and bottom tray is given below.

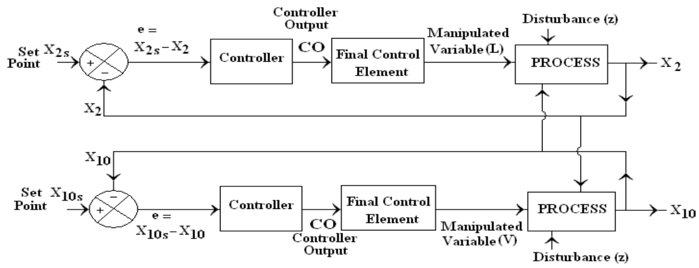
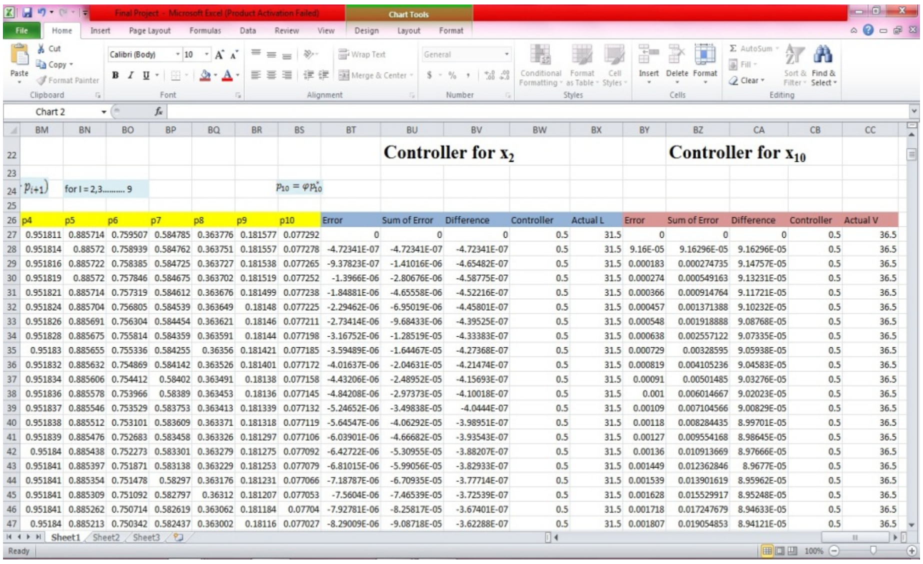


Fig 4: Block diagram of distillation column control



	BM	BN	BO	BP	BQ	BR	BS	BT	BU	BV	BW	BX	BY	BZ	CA	CB	CC
22									Controller for x_2					Controller for x_{10}			
23																	
24	$p_{i+1})$	for I = 2,3.......... 9					$p_{10} = \varphi p_{10}^*$										
25																	
26	p4	p5	p6	p7	p8	p9	p10	Error	Sum of Error	Difference	Controller	Actual L	Error	Sum of Error	Difference	Controller	Actual V
27	0.951811	0.885714	0.759507	0.584785	0.363776	0.181577	0.077292	0	0	0	0.5	31.5	0	0	0	0.5	36.5
28	0.951814	0.88572	0.758939	0.584762	0.363751	0.181557	0.077278	-4.72341E-07	-4.72341E-07	-4.72341E-07	0.5	31.5	9.16E-05	9.16296E-05	9.16296E-05	0.5	36.5
29	0.951816	0.885722	0.758385	0.584725	0.363727	0.181538	0.077265	-9.37823E-07	-1.41016E-06	-4.65482E-07	0.5	31.5	0.000183	0.000274735	9.14757E-05	0.5	36.5
30	0.951819	0.88572	0.757846	0.584675	0.363702	0.181519	0.077252	-1.3966E-06	-2.80676E-06	-4.58775E-07	0.5	31.5	0.000274	0.000549163	9.13231E-05	0.5	36.5
31	0.951821	0.885714	0.757319	0.584612	0.363676	0.181499	0.077238	-1.84881E-06	-4.65558E-06	-4.52216E-07	0.5	31.5	0.000366	0.000914764	9.11721E-05	0.5	36.5
32	0.951824	0.885704	0.756805	0.584539	0.363649	0.18148	0.077225	-2.29462E-06	-6.95019E-06	-4.45801E-07	0.5	31.5	0.000457	0.001371388	9.10232E-05	0.5	36.5
33	0.951826	0.885691	0.756304	0.584454	0.363621	0.18146	0.077211	-2.73414E-06	-9.68433E-06	-4.39525E-07	0.5	31.5	0.000548	0.001918888	9.08768E-05	0.5	36.5
34	0.951828	0.885675	0.755814	0.584359	0.363591	0.18144	0.077198	-3.16752E-06	-1.28519E-05	-4.33383E-07	0.5	31.5	0.000638	0.002557122	9.07335E-05	0.5	36.5
35	0.95183	0.885655	0.755336	0.584255	0.36356	0.181421	0.077185	-3.59489E-06	-1.64467E-05	-4.27368E-07	0.5	31.5	0.000729	0.00328595	9.05938E-05	0.5	36.5
36	0.951832	0.885632	0.754869	0.584142	0.363526	0.181401	0.077172	-4.01637E-06	-2.04631E-05	-4.21474E-07	0.5	31.5	0.000819	0.004105236	9.04583E-05	0.5	36.5
37	0.951834	0.885606	0.754412	0.58402	0.363491	0.18138	0.077158	-4.43206E-06	-2.48952E-05	-4.15693E-07	0.5	31.5	0.00091	0.00501485	9.03276E-05	0.5	36.5
38	0.951836	0.885578	0.753966	0.58389	0.363453	0.18136	0.077145	-4.84208E-06	-2.97373E-05	-4.10018E-07	0.5	31.5	0.001	0.006014667	9.02023E-05	0.5	36.5
39	0.951837	0.885546	0.753529	0.583753	0.363413	0.181339	0.077132	-5.24652E-06	-3.49838E-05	-4.0444E-07	0.5	31.5	0.00109	0.007104566	9.00829E-05	0.5	36.5
40	0.951838	0.885512	0.753101	0.583609	0.363371	0.181318	0.077119	-5.64547E-06	-4.06292E-05	-3.98951E-07	0.5	31.5	0.00118	0.008284435	8.99701E-05	0.5	36.5
41	0.951839	0.885476	0.752683	0.583458	0.363326	0.181297	0.077106	-6.03901E-06	-4.66682E-05	-3.93543E-07	0.5	31.5	0.00127	0.009554168	8.98645E-05	0.5	36.5
42	0.95184	0.885438	0.752273	0.583301	0.363279	0.181275	0.077092	-6.42722E-06	-5.30955E-05	-3.88207E-07	0.5	31.5	0.00136	0.010913669	8.97666E-05	0.5	36.5
43	0.951841	0.885397	0.751871	0.583138	0.363229	0.181253	0.077079	-6.81015E-06	-5.99056E-05	-3.82933E-07	0.5	31.5	0.001449	0.012362846	8.9677E-05	0.5	36.5
44	0.951841	0.885354	0.751478	0.58297	0.363176	0.181231	0.077066	-7.18787E-06	-6.70935E-05	-3.77714E-07	0.5	31.5	0.001539	0.013901619	8.95962E-05	0.5	36.5
45	0.951841	0.885309	0.751092	0.582797	0.36312	0.181207	0.077053	-7.5604E-06	-7.46539E-05	-3.72539E-07	0.5	31.5	0.001628	0.015529917	8.95248E-05	0.5	36.5
46	0.951841	0.885262	0.750714	0.582619	0.363062	0.181184	0.07704	-7.92781E-06	-8.25817E-05	-3.67401E-07	0.5	31.5	0.001718	0.017247679	8.94633E-05	0.5	36.5
47	0.95184	0.885213	0.750342	0.582437	0.363002	0.18116	0.077027	-8.29009E-06	-9.08718E-05	-3.62288E-07	0.5	31.5	0.001807	0.019054853	8.94121E-05	0.5	36.5

Fig 5: MS excel spread sheet for control of top and bottom trays

Disturbances and control of top and bottom tray

When the feed composition is disturbed to z=0.7 the graph of top and bottom try is given by,

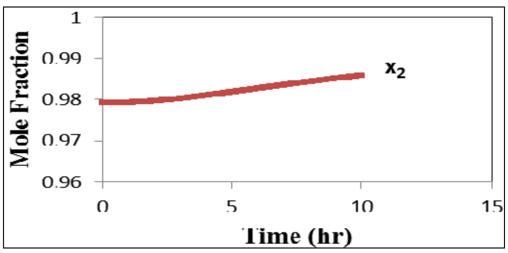


Fig 6: Feed disturbance graph for top tray when z =0.7

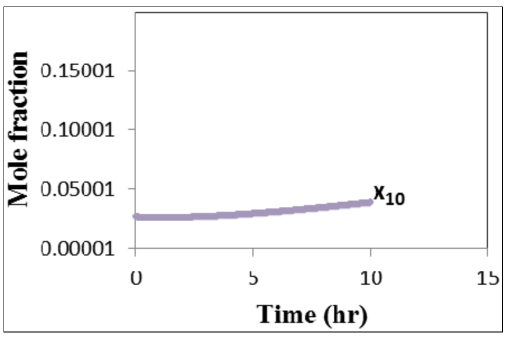


Fig 7: Feed disturbance graph for bottom tray when z

Table 3: Disturbances controlled by using PID controller for top and bottom tray

pc1	0.5	pc2	0.5
kc1	5.5	kc2	0.0001
τi1	0.001	τi2	0.01
τd1	0.01	τd2	0.01

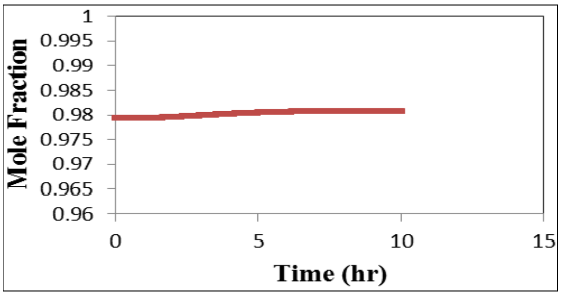


Fig 8: Control graph for top tray when z = 0.7

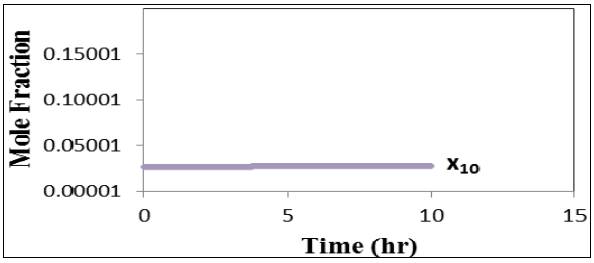


Fig 9: Control graph for bottom tray when z = 0.7

When the feed composition is disturbed to z =0.3 the graphs are given by,

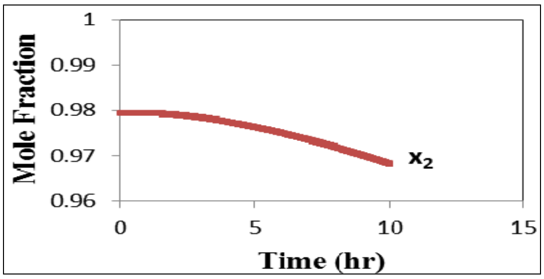


Fig 10: Feed disturbance graph for top tray when z = 0.3

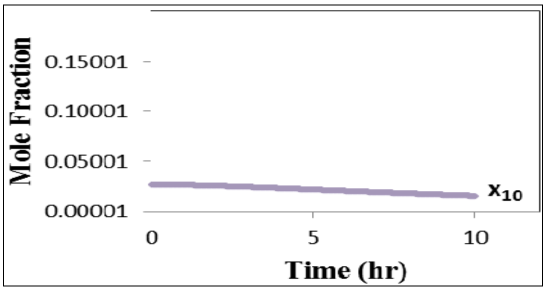


Fig 11: Feed disturbance graph for bottom tray when z = 0.3

Table 4: control numerical values for both trays

pc1	0.5	pc2	0.5
kc1	20	kc2	0.00001
τi1	0.0001	τi2	0.0001
τd1	0.001	τd2	0.01

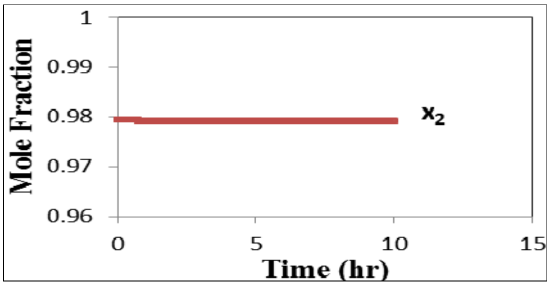


Fig 12: Control graph for top tray when z = 0.3

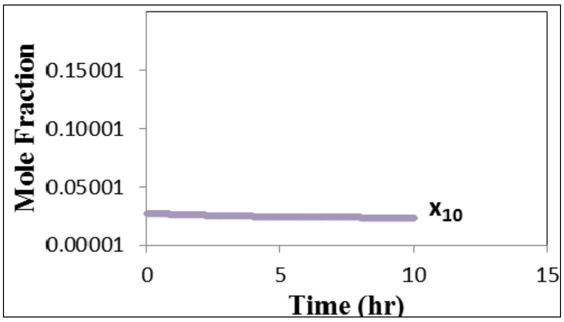


Fig 13: Control graph for bottom tray when z = 0.3

Control of trays with span and time delay

As we said before we have set the delay of two hours and span of control to the column, the numerical values and control graphs with span and time delay for both the trays is shown below,

Table 5: The numerical values when z =0.7

pc1	0.5	pc2	0.5
kc1	5	kc2	0.0001
τi1	0.001	τi2	0.01
τd2	0.01	τd2	0.01

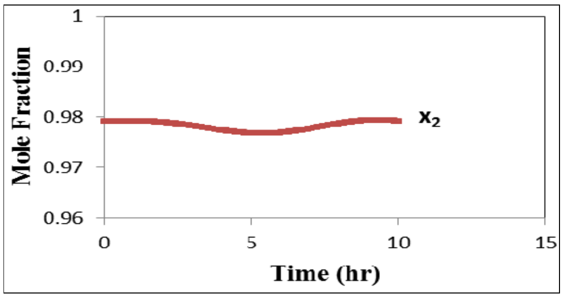


Fig 14: Control graph for top tray hen z = 0.7 with span and time delay

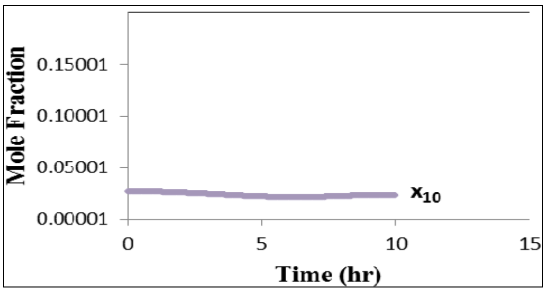


Fig 15: Control graph for bottom tray hen z = 0.7 with span and time delay

Table 6: The numerical values when z =0.3

Pc1	0.5	pc2	0.5
Kc1	4.8	kc2	0.0001
Ti1	0.001	τi2	0.001
τd1	0.01	τd2	0.01

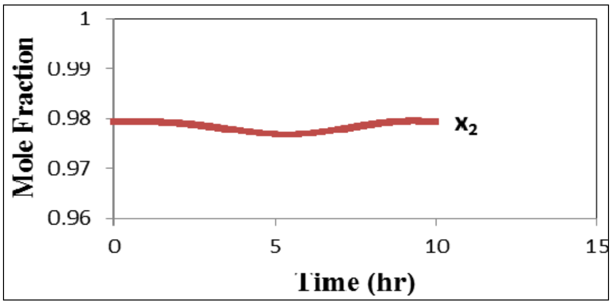


Fig 16: Control graph for top tray hen z = 0.3 with span and time delay

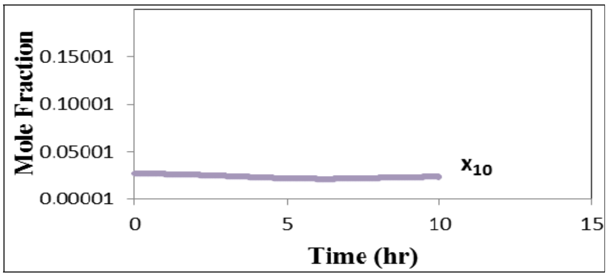


Fig 17: Control graph for bottom tray when z = 0.3 with span and time delay

Conclusion

So, the dynamics of binary distillation column was studied with open loop response and we controlled the top and bottom tray of the distillation column using proportional integral derivative controller in excel.

Acknowledgement

First and foremost, our thanks are to the almighty, the great architect of the universe, who blessed us to successes fully pursue our Bachelor of Engineering and to successfully accomplish our project work. We would like to express our heart full thanks to Dr. G. RANGANATH, Principal, Adhiyamaan college of Engineering, Hosur for his endorsement of this project work.

We express our sincere gratitude to Dr. N.S.BADARINARAYANAN, Dean for his encouragement. We are grateful to Mr.K. Selva kumar**,** Head of the Department, Department of Chemical Engineering, for providing the facilities to carry out this project work successfully.

We express our profound sense of gratitude and grateful to our guide Dr. C.AHMED BASHA, Department of Chemical Engineering for his valuable and continuous support in completing the project work. We would also extend our thanks to, Prof. P.MANOKARAN**,** Asso. Professor, Department of Chemical Engineering and other staffs of Department of Chemical Engineering for extending their co-operation to carry out this project work as fruitful one.

We feel elated expressing deep and sincere thanks to the guide Finally, we thank our parents to support us to do this project work with their blessing.

References

1. Computational performance of aggregated distillation models by Andreas Linhart & Sigurd Skogestad.
2. Dynamic compartmental models for separation processes AIChe Journal by Benallou, A., Seborg, D. E., & Mellichamp, D. A., 1986.
3. Nonlinear model predictive control'. *Progress in* systems theory by Allgower, F., & Zheng, A., 2000.
4. Compartmental modeling of high purity air separation columns by Bian, S., Khowinij, S., Henson, M. A., Belanger, P., & Megan, L., 2005.
5. Dynamic compartmental modeling of nitrogen purification columns. Separation and Purification Technology by Khowinij, S., Henson, M. A., Belanger, P., & Megan, L., 2005.
6. Quality control of binary distillation columns via nonlinear aggregated models'. *Automatica*, *27*, 463 by Levine, J., & Rouchon, P., 1991.