DESIGN OF DECOUPLER AND PERFORMANCE ANALYSIS OF A DISTILLATION COLUMN

A Thesis submitted for partial fulfilment for the degree of

Master of Technology

in

Electronics and Instrumentation Engineering

By

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CERTIFICATE

This is to certify that the project report titled **"DESIGN OF DECOUPLER AND PERFORMANCE ANALYSIS OF A DISTILLATION COLUMN"** submitted by Manash Kumar Sethi (210EC3319) in the partial fulfilment of the requirements for the award of Master of Technology Degree in Electronics and Instrumentation Engineering during session 2010-2012 at National Institute of Technology, Rourkela (Deemed University) and is an authentic work carried out by them under my supervision and guidance.

To the best of my knowledge, the matter embodied in the thesis has not been submitted to any other university/institute for the award of any Degree or Diploma.

Date:

Prof. T.DAN
Department of E.C.E
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Manash Kumar Sethi

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List of acronyms

MIMO	Multiple Input and Multiple Output
SISO	Single Input and Single Output
TITO	Two Inputs and Two Outputs
P	Proportional
PI	Proportional Integral
PID	Proportional Integral and Derivative
RGA	Relative Gain Array
RNGA	Relative Normalized Gain Array
ETF	Estimated Transfer Function
DRF	Decentralized Relay Feedback
GA	Genetic Algorithm

Abstract

The main aim of this thesis is to control the basic parameters of distillation column .The distillation column is basically a MIMO process that means all the inputs and outputs are coupled to each other. It is very difficult to control such type of process so, we have to reduce or eliminate the interaction between the inputs and outputs. Therefore the process can be converted to a single input and single output system (SISO).In order to convert the MIMO system to SISO system it is necessary to design a decoupler which will eliminate the interaction among all the inputs and outputs.

My focus is here to design that decoupler, which is very difficult to design when the process variables are more. It is easy when we consider TITO system. When we consider a 'three input and three input' and 'four input and four output' system it is necessary to follow some methods like RGA and RNGA methods. After getting those parameters we have to follow the ETF method so that the decoupler will be designed.

After designing the decoupler, our aim to be controlled the distillation process. Here I am using PID controller to control the process. In order to design the PID controller we basically emphasized on the tuning parameters of the PID. To get the tuning parameters we must follow certain methods such as Ziegler Nichols, Cohen coon and decentralized relay feedback methods, BLT methods etc. Here we are using decentralized relay feedback method in to do the tuning of the controller. After tuned the PID .We will get the desired output what we actually want to get.

The result was obtained and shown that implemented work is successfully done. Finally the interactions are rejected and the process was controlled.

Here all the simulations are done by the MATLAB tool and Microsoft window 7operating system

Chapter 1

Introduction

1.1 Goal and Motivation

1.1.1 Goal

The distillation column is highly interactive multivariable process, so it's very difficult to control such a complex process without eliminating that interactions, therefore in order to completely eliminate these interactions a decoupler is necessarily important. Hence in this thesis my goal is to design a decoupler for the distillation column process, which will make the multiple input multiple output system (MIMO) to a single input and single output system (SISO) and making the process simple by which we can easily perform the controlling strategies that make the system in control.

1.1.2 Motivation

Use of distillation column is very vast in process control industries and oil refineries. It plays a major role in separating the constituent of mixture to make the process automatic. So we must analyse, its controlling parameters by which this distillation process can be controlled. This process seems to be a very complex one. The controlled variables and manipulated variables must be paired so that the necessary controlling action can be taken place. After getting this valuable information about distillation column I found that it's an interesting field to work on. Here my main motive is to reduce the interaction among its inputs and outputs of the distillation column.

1.2 Objective

The objective of this thesis is to separate the liquid or vapour mixture of two or substances are separated into its component fractions of the desired purity by application and removal of heat. Distillation is based on the fact that the vapour of boiling mixture will be richer in the components that have lower boiling points therefore, when the vapour is cooled and condensed the condensate will contain more volatile components at the same time, the original mixture will contain more of the less volatile material. Distillation column is basically a MIMO process means multiple inputs and multiple outputs so when I design and control this process I must have to decide which variables are controlled through which manipulating variables then designing an appropriate pairing between the control variable and manipulated variable and construct decoupler for this distillation process to eliminate the interaction between the inputs and outputs and make the MIMO process that is distillation column process to be a SISO means single input single output system, so that I can easily tuned the process parameter using any controller such a PID controller like that. After tuning those process parameters I have to get the appropriate set points what I have already set before.

Chapter 2

OVERVIEW OF DISTILLATION COLUMN

2.1 Use of Distillation Column.

Distillation column is process of separation so all the physical separation of mixture is done through this distillation column therefore; it has many applications in the fields like 1.Industrial distillation 2.Azeotropic distillation 3.Distillation in food processing etc.

2.2 No of control variable and its importance.

Mainly distillation column has 4N+10 variables and control variables are X_D,X_B,M_D, M_B, and Pressure.

X_D:Distillate Composition, mol fraction

X_B: Bottoms Composition, mol fraction

M_D: Liquid holdup in reflux drums, mols

M_B: Liquid holdup in column base/reboiler, mols

2.3 Different approaches for controlling these control variables

There are many approaches for controlling the control variables of the distillation column such as feedback control, feed forward control and MPC control etc. Here I am doing the control through feed forward controlling scheme. Through this feed forward control scheme I am using PID control to do the all controls.

2.4 PID control outputs and its advantages and disadvantages

The PID algorithm is the most popular feedback controller used in the process industries. It has been successfully used over many years because of its robustness and easily understood algorithm it can perform excellent control despite the varied dynamic characteristics of plant.

PID algorithm consists of three basic modes such as proportional mode, integral mode and derivative mode at the time of doing this algorithm it is highly necessary to use which mode (P, I and D).and specify the parameters for each mode. Generally three basic algorithms are used P, PI or PID.

2.4.1 Proportional algorithm

The mathematical representation is,

$$\frac{mv(s)}{e(s)} = k_c (Laplace\, domain)\, or\, mv(t) = mv_{ss} + k_c e(t)\, (time\, domain)$$

The proportional mode adjusts the output signal in direct proportion to the controller input that is error signal "e". The adjustable parameter to be specified control gain, k_c.The larger the value of k_c more the control output is changed for a given error. The proportional gain can also be called as Proportional Band (PB) instead of k_c.

A proportional controller reduces the error but does not eliminate it so there is an offset between actual and desired value will normal exist.

2.4.2 A Proportional Integral algorithm

$$\frac{mv(s)}{e(s)} = k_c[1 + \frac{1}{T_i s}]\ or\ mv(t) = mv_{ss} + k_c[e(t) + \frac{1}{T_i}\int e(t)dt]$$

The integral mode often referred as reset correct for any offset (error) that may occurred between the desired value and the process output automatically over the time. The adjustable parameter is here (T_i) of the controller. Here the integral term accelerates the movement of the process towards the set point and eliminates the residual steady state error that occurs in the pure proportional controller .However, since the integral term respond to the accumulated errors from the past, it can cause the present value to overshoot the set point value.

2.4.3 A Proportional Integral and Derivative algorithm

$$\frac{mv(s)}{e(s)} = k_c[1 + \frac{1}{T_i s} + T_D s]\ or\ mv(t) = mv_{ss} + k_c[e(t) + \frac{1}{T_i}\int e(t)dt + T_D\frac{de(t)}{dt}]$$

Derivative action (also called rate or pre act) anticipate where the process is heading by looking at the time rate of change of control variable (its derivative).T_D is the rate time and this characteristics the derivative action .Here derivative action always improve the dynamic response and it does in many loops. Derivative action depends on the slope of the error unlike P and I. If the error is constant derivative action has no effect. The integral term eliminate the steady state error. So as a whole PID control contains the whole advantages over P and PI controller.

2.4.4 PID Tuning algorithm

The controller tuning involves the best values of k_c, T_i and T_D. There are various methods are used to tune the PID controller such as 1. Ziegler Nichols closed loop method 2.Cohen Coon method 3.Decentralized relay feedback method 4.Colonial competitive algorithm 5.Genetic algorithm etc.

2.4.4.1 Ziegler Nichols closed loop method

This method was proposed by *Nathaniel B. Nichols* .It is performed by setting the integral and derivative gain to zero. The proportional gain i.e. k_c is increased from zero to ultimate gain K_u ,at which the controller output oscillates with a constant oscillation with constant amplitude .K_u and oscillation period T_u are used to set P,I and D gain depending upon the type of controller used.

For PID controller these are the values given by *Ziegler Nichols,*

$$K_i = 2K_p / T_u \, , K_c = 0.60K_u \, , K_d = K_p T_u / 8$$

2.4.4.2 Cohen Coon method

Cohen Coon method depends upon the identification of a suitable process model. Cohen Coon recommended the following setting to give responses having ¼ decay ratios, minimum offset and other favourable property.

This technique generally used in small delay increasingly larger control gains will be predicted ,therefore this method is used where the delay is almost equal to zero or no delay.

Those are the values given by *Cohen Coon*,

$$K_c = \frac{1}{K_p} \frac{\tau}{\theta} (\frac{4}{3} + \frac{\theta}{4\tau}) \, , T_i = \theta \frac{32 + 6(\theta / \tau)}{13 + 8(\theta / \tau)}$$

2.4.4.3 Decentralized relay feedback method

The relay feedback method has been introduced in control application since many years

It is mainly introduced to design for tuning of PID control [1].

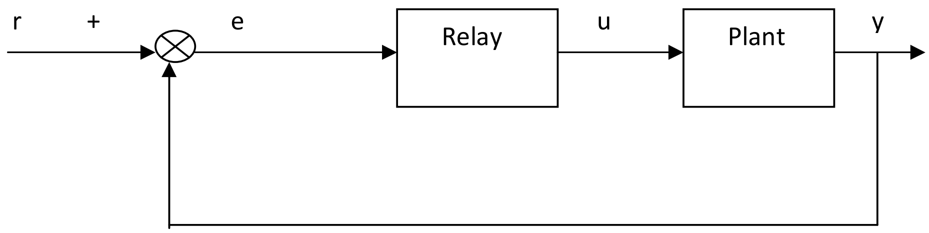


Figure2.1 Relay Feedback system

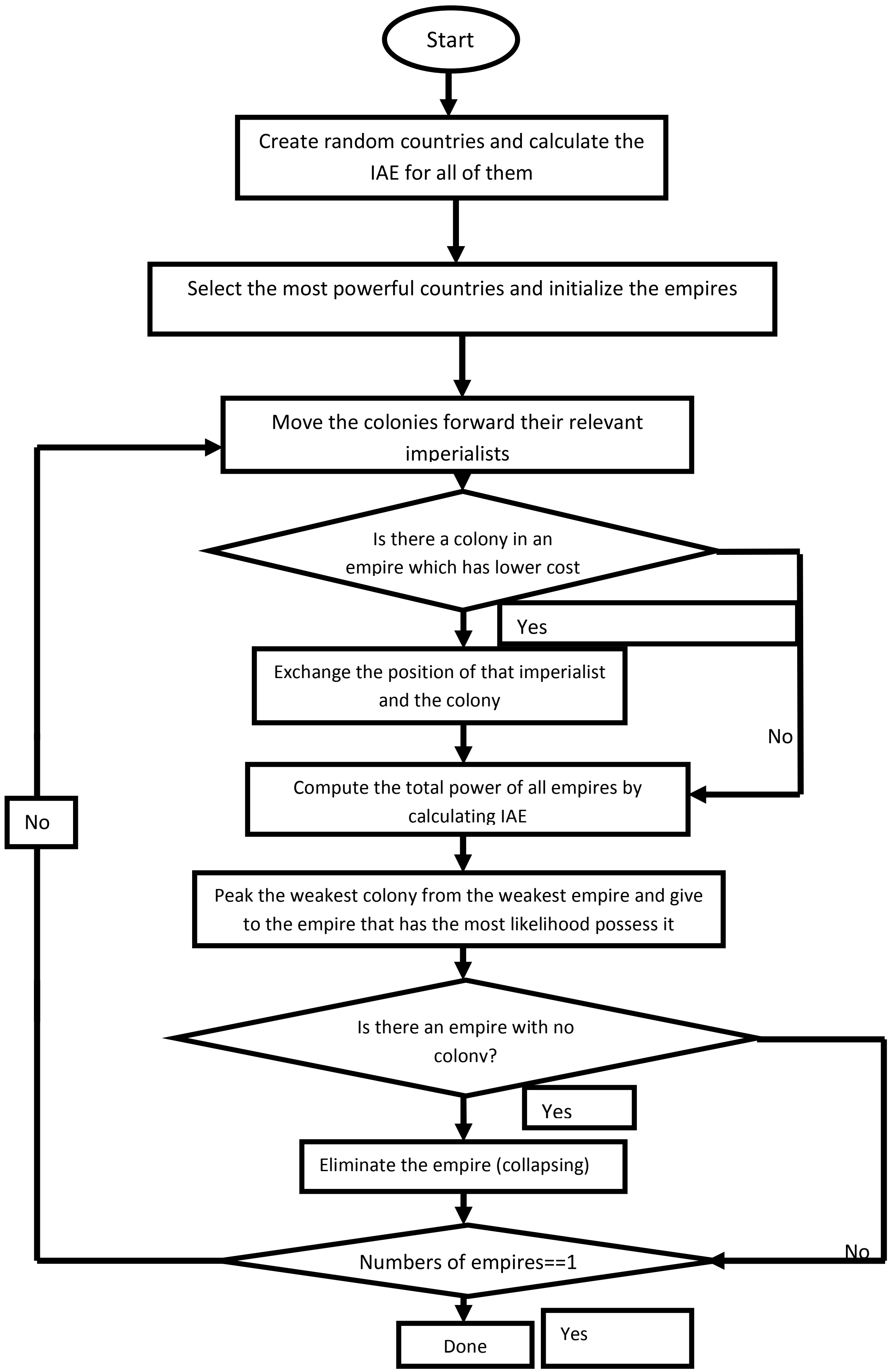
Continuous cycles of control variable are generated by this relay feedback method and the model equation can be easily extracted by it. It's a one shot experiment method.

2.4.4.4 Colonial competitive algorithm

CCA is search strategy that uses the socio political competition among empires as a source of inspiration, CCA begins with an initial empire .Any individual of an empire is a country. There are two types of countries colony and imperial state that in a group form empire [1].

Imperialistic competition among those countries forms the basis of CCA. During this competition weak empires collapse and powerful ones take possession of their colonies. Imperialistic competitions reporting to a state in which there exist only one empire and its colonies are in the same position and have the same cost as the imperialists [1].

CCA is used to design PID control tuning and for characterization of material property from sharp notch test.


Start
Create random countries and calculate the IAE for all of them
Select the most powerful countries and initialize the empires
Move the colonies forward their relevant imperialists
Is there a colony in an empire which has lower cost
Yes
No
Exchange the position of that imperialist and the colony
Compute the total power of all empires by calculating IAE
Peak the weakest colony from the weakest empire and give to the empire that has the most likelihood possess it
Is there an empire with no colony?
Yes
No
Eliminate the empire (collapsing)
Numbers of empires==1
No
Yes
Done

2.4.4.5 Genetic algorithm

Genetic algorithm is an academic global search method that mimics the process of natural evolution. Using the genetic algorithm to perform the tuning of the controller will result in the optimum controller being evaluated for the system every time.

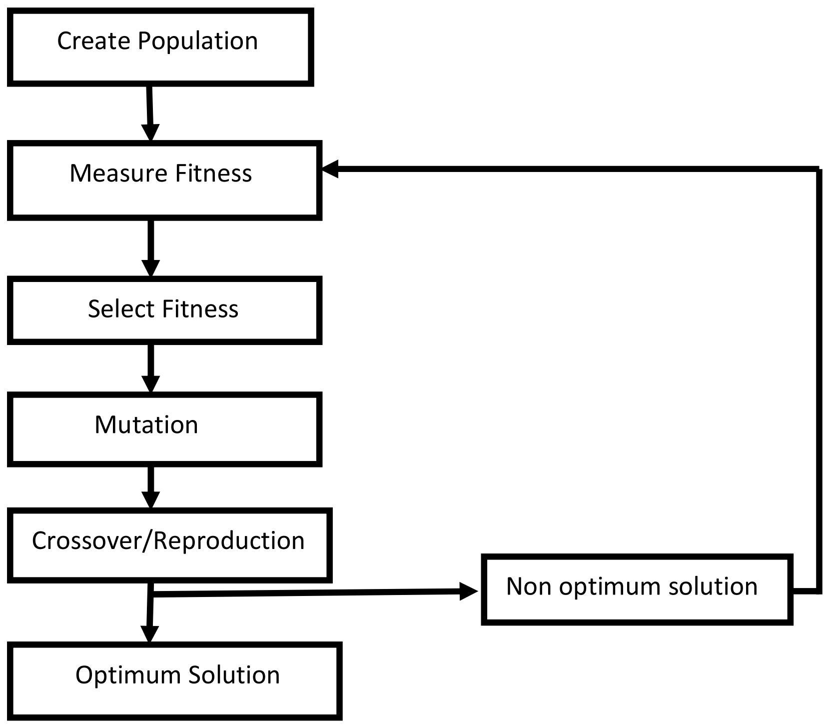


Figure2.2Genetic algorithms

Chapter 3

Theory of Distillation Column

3.1 Schematic Diagram

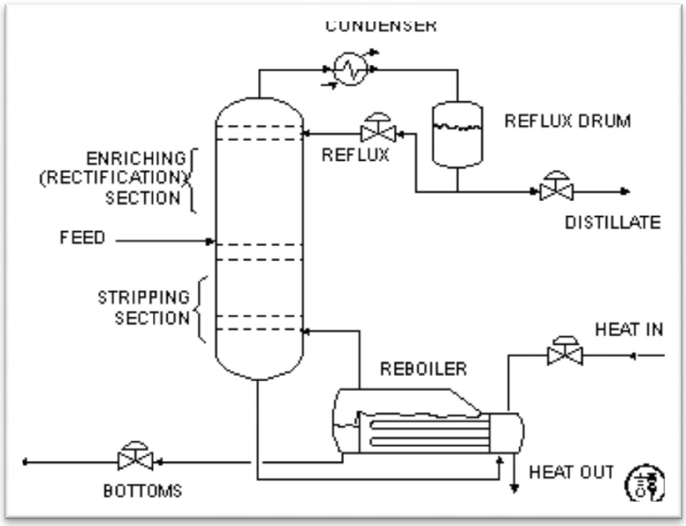


Figure3.1Distillation Column

The above diagram is a Binary Distillation column. Distillation is a process in which a liquid or a vapour mixture of two or more substances is separated into its component fractions of desired purity by the application and removal of heat. It is well known that pure liquid exhibits different volatilities (i.e., vapour pressure) at a given temperature, and thus if heat is applied to a liquid mixture of these substances, the vapour so generated will be richer in the more volatile substances, those having higher vapour pressures. If this vapour condensed, it should be clear that a certain amount of purification will be achieved. This is a basic principle underlying a distillation process [13].

A distillation process may be classified in one of two ways: Binary distillation refers to the separation of two substances and multicomponent distillation involves more than two substances [13].

Distillation may be carried out either as a batch operation or as a continuous operation. It may be affected in continuous contact equipment (i.e. packed towers) or in stage wise contact towers [13].

The above schematic diagram consists of typical stage type distillation column (Fig. 1).The equipment consists of vertical shell with a number of equally spaced trays mounted inside of it. Each tray contains two conduits, one on each side, called down comers by gravity from each tray to the one bellow. A weir on one side of the tray maintains the liquid level at a suitable height on that tray. A variety of tray types is available commercially. The simplest is a sieve tray, a sheet of metal containing a number of perforations that are provided for vapour flow. The flow of vapour must be sufficiently high to prevent weeping of liquid through the holes. A bubble cap tray has a riser fitted over each hole and a cap that covers the space between the riser and cap for the passage of vapour. Vapour rises through the chimney and directed downwards by the cap, which finally discharges from the slots in the cap and bubbles through the liquids on the tray [13]. For years bubble cap trays were best known vapour liquid contracting devices in the chemical and petroleum industry. Vapour flow directs the flow of vapour horizontally into the liquid, which provides better mixing than, is possible in sieve trays where the vapour passes straight upward through the liquid. Because of their efficiency, wide operating range and cost factors, sieve and valve trays have replaced bubble cap trays in many applications.

The vertical shell is connected by suitable piping to heating devices called a re-boiler, which provides the necessary vaporization for the distillation operation, and to a condenser, which condense the over head vapours.

Feed enters the central portions of the column an tray called the feed tray. It flows down by gravity from tray to tray and in the process comes into contact on each tray with the vapour rising from the tray below. The liquid from tray 1 flows into the base of the column and then into re-boiler, where it is partially vaporized .The unvapourized liquid is one of the product, and is removed from the re-boiler in the arrangement shown in fig.1. The bottom product has the highest concentration of least volatile substances and thus its temperature, which is also the temperature of the vapour generated in the re-boiler, the highest of any location in the column. Since the volatility of the two substances involved in the distillation process is different, the vapour generated in the re-boiler is richer in the more volatile component. This vapour rises and comes into contact with descending stream of liquid on each tray, beginning with that on tray 1. The mixing of warmer vapour with the liquid results in the transfer of heat and mass, and the net result is some vaporization of the more volatile component and condensation of a thermally equivalent amount of less volatile component. Thus the feed is stripped of its more volatile component as it flows downward and it becomes the trays below it constitute what is called the stripping section .The vapour rising from the feed tray comes into contact with a liquid that is more volatile substance vaporizes at the expense of some of the less volatile substance vaporizes at the expense of some of the less volatile substance which condences. Thus the vapour becomes enriched in the more volatile substances as it flows up the column. The vapour from the top tray contains a higher concentration of the more volatile substance than anywhere else in the column. It is condensed in a total condenser in the arrangement shown in fig.1, a part of it is removed as the distillate product, and the terms of the more volatile component, contact the ascending vapour rising from the feed tray and heat and mass transfer processes occur as described earlier. The reflux stream combines with the feed and serves as the liquid phase in the stripping section.

Binary distillation process may be solved by either graphical or analytical methods. McCabe – Thiele method and The Ponchon – Savarit method. The analytical method selected is the one developed by Smoker (1938).

3.2 Graphical methods for Binary distillation column

3.2.1 McCabe-Thiele Method

This method can be used when following conditions are satisfied

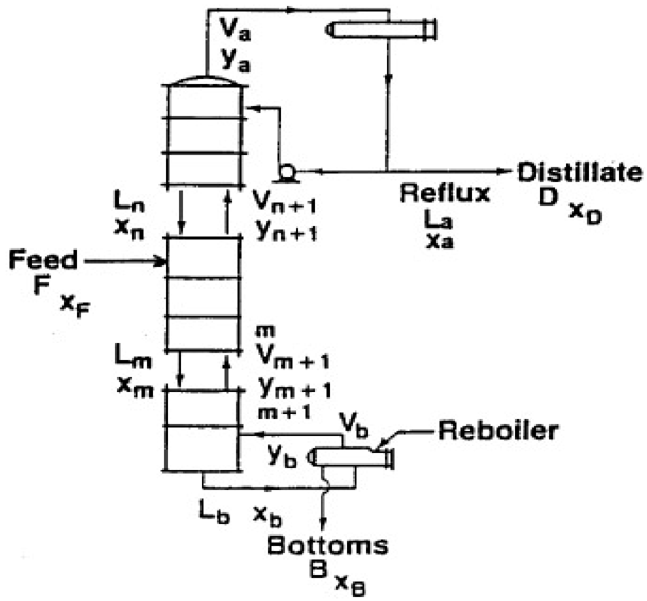
1. Molal heat of vaporization of two substances are roughly the same.
2. Heat effects (heats of solution, heat losses to and from column) are negligible.

Theses so called constant molal overflow assumptions imply that for every mol of vapour condensed, 1 mol of liquid is vaporized. Thus the liquid and vapour rates within each section of the tower remain constant [13].

[13] The McCabe –Thiele method utilizes material balance and equilibrium relationships. These relationships are written for the enriching section and the stripping section and then combined to solve the binary distillation problem. We first consider the determination of the number of ideal trays required for a specified separation and then learn to account for tray efficiency. Component material balance around a general tray n in the enriching section is written as,

$$y_n = \frac{L}{V} x_{n+1} + \frac{D}{V} x_D \qquad \text{.................... (1)}$$

The equation for n^{th} tray in stripping section is written as,



……………………..eq2

Figure4 Material Balance Envelopes for Operating Line

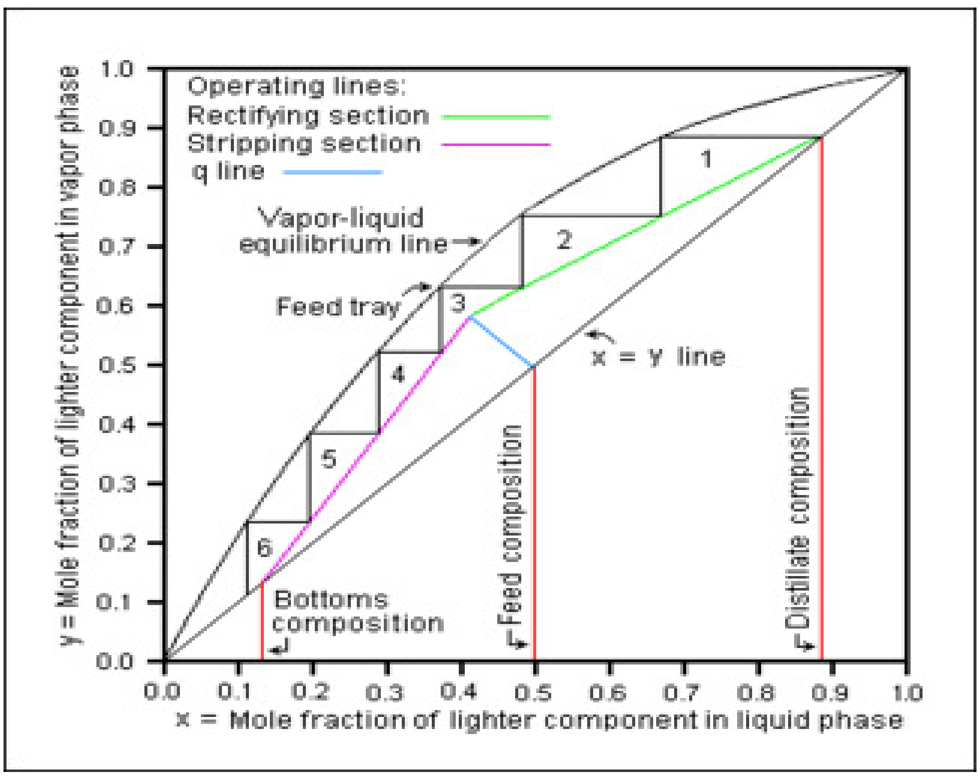


Figure 3.3 McCabe Thiele Plot

3.2.2 Calculation of Trays

Here first we have tried to find the total number of trays that are required to separate a binary mixture (two substances) into its components fractions. For this considered an example from "Distillation Dynamics and Control" author Pradeep B. Deshpande Example 3.4 (page no 94).

Example: 5.1 Distillate, Bottoms and feed mole fractions are equal to 92% mol, 8% mol and 60% mol respectively. Reflux Ratio (R) =1.51435. feed is two-phase mixture with feed quality equal to 0.85.

Total material balance equation (see chapter 3) is $F = D + B$

$$100 = D + B$$

And component balance equation is

$$Fx_F = Dx_D + Bx_B$$

$$0.6(100) = 0.92D + 0.08B$$

Solving simultaneously: Distillate flow (D), = 50.0 mols

Bottoms flow (B), = 50.0 mols

For calculating the number of trays we use McCabe-Thiele graphical method. McCabe-Thiele method and the construction of stages are explained in the Chapter 2.2. For calculating the number of stages Matlab domain is used. The computation of number of stages for $x_{D=}$0.92, x_B=0.08, F=0.6 with Reflux ratio of 1.51435, q=0.85 and constant relative volatility (α) =2.45 is shown in the Figure 5.1.

As the numbering in the Figure 5.1 minimum 9 stages are required for the extant separation of binary mixture.

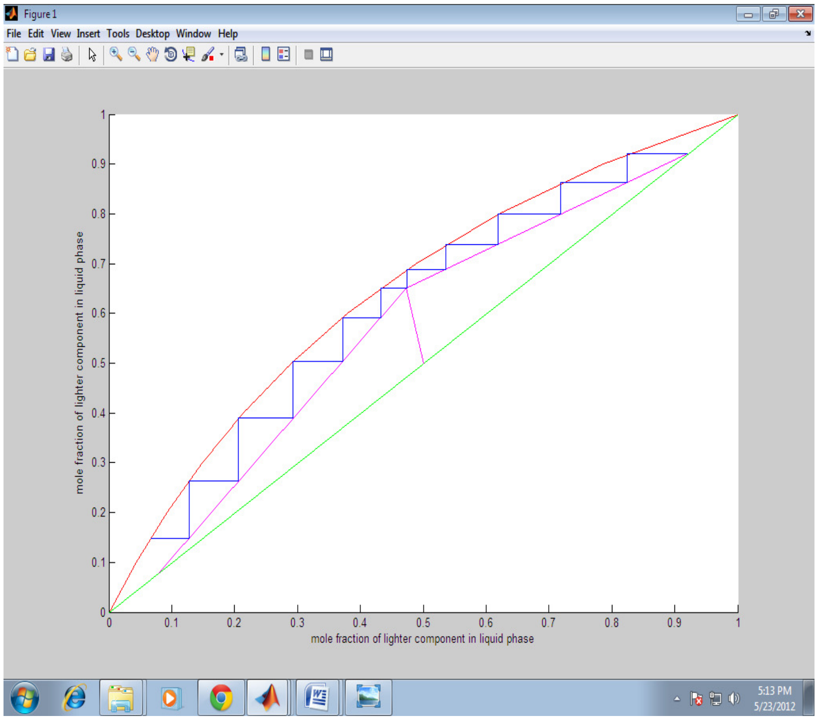


Figure3. 5 McCabe-Thiele Plot for no of trays

3.4 Controlled Variables in Distillation

Having studied steady state distillation concepts, we now shift our focus to control concepts. How many variables should be controlled to achieve the production objectives? How should the manipulated variables be selected? How should the manipulated and controlled variables be paired? Where should the sensor be located?

Intuition suggests that the variables should be controlled for the following reasons

- Product composition
- Column pressure
- Base level
- Reflux accumulator level

First, production objectives normally require the delivery of products of acceptable quality. The column pressure should be maintained constant.

Table3.1 No of equations in model.

	Number of equations
Reflux drum: overall balance	1
Reflux drum: component balance	1
Tray balance: overall balance	N
Tray balance: component balance	N+1
Equilibrium relationships(trays+reboiler)	N
Hydraulic relationships	1
Reboiler: overall balance	1
Total	4N+5

Table3.2 No of variables in the equations.

	Number of unknowns
Tray compositions(x_n,y_n)	2N
Tray liquid holdups(M_n)	N
Liquid flows(L_n)	N
Distillate composition(x_D)	1
Reflux drum flows(L_{N+1},D)	2
Reflux drum holdup(M_D)	1
Reboiler compositions(x_B, y_B)	2
Reboiler flows(V,B)	2
Column base holdup	1
Column pressure	1
Total	4N+10

1. [13] We specified that the five variables will be maintained constant. This is the closed loop situation. For reasons cited earlier, the variables selected are x_D, x_D, M_D, M_B, P. With these variables so specified, we can calculate the values of all other variables in the system. In particular, we can calculate the values of five manipulated variables that can hold the five controlled variables at set point. These manipulated variables are the flow rates that the controllers may adjust. There are

D,B,M_s,M_c,L_{N+1}.We have not discussed which of these five manipulated variables will be used to control the selected controlled variables.

2. We specify the values of five manipulated variables. This is the open loop situation. Again the given F and x_F the specification of the five manipulated variables will enable us to calculate the values of all the remaining variables in the system. In particular, the values of five variables resulting from the specification of the five manipulated variables may occur [13].

3.5 Basis for distillation column strategies

[13] For simplicity we will consider here the separation of a single binary feed F into the two products *D* and *B*.

The overall material balance equation for this column,

$$F = D + B \qquad \text{……………} (3)$$

The component material balance for the more volatile substance,

$$Fx_F = Dx_D + Bx_B \qquad \text{…………….} (4)$$

The control objectives in distillation operations are to maintain x_D and/or x_B at set point in the presence of disturbances. These disturbances may be characterized as (1) Process load, (2) Changes in cooling and heating medium supply conditions, and (3) equipment fouling.

There are two major controlling strategies,

1. Single control composition control problem
2. Dual composition control problem

3.5.1.1 Control of x_D or x_B for upsets in F

It applies to those cases in which tight control of one of the two product compositions is deemed sufficient. We can easily show that x_D can be maintained at set point by manipulating *D* while holding *V* constant [13].

$$\frac{D}{F} = \frac{x_{F_s} - x_B}{x_{D_s} - x_B} \qquad \text{…} (5)$$

$$S = \frac{x_{D_s}(1 - x_B)}{x_B(1 - x_{D_s})} \qquad \ldots (6)$$

3.5.1.2 Control of x_D or x_B for upsets in x_F

The control system developed for control of x_D (or x_B) for upsets in x_F.The governing equation for control of x_D are

$$\frac{D}{F_s} = \frac{x_F - x_B}{x_{D_s} - x_B} \qquad \ldots \ldots (7)$$

$$S = \frac{x_{D_s}(1 - x_B)}{x_B(1 - x_{D_s})} \qquad \ldots\ldots\ldots\ldots (8)$$

3.5.2.1 Control of x_D and x_B for upsets in F

This is a dual composition control problem. For this case governing equations are

$$\frac{D}{F} = \frac{x_{F_s} - x_{B_s}}{x_{D_s} - x_{B_s}} \qquad \ldots. (9)$$

$$S_0 = \frac{x_{D_s}(1 - x_{B_s})}{x_{B_s}(1 - x_{D_s})} \qquad \ldots (10)$$

3.5.2.2 Control of x_D and x_B for upsets in x_F

The relevant equation in this case,

$$\frac{D}{F_s} = \frac{x_F - x_{B_s}}{x_{D_s} - x_{B_s}} \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (11)$$

And

$$S_0 = \frac{x_{D_s}(1 - x_{B_s})}{x_{B_s}(1 - x_{D_s})} \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (12)$$

3.6 Pairing and interaction in distillation

A systematic procedure is given [13] for determining the correct pairing of control and manipulative variables in single composition control and dual composition control. In dual composition control, interaction among control loops can be significant problem even when the variables are paired correctly. Here it is proposed a method for eliminating interaction in dual composition control.

The distillation column upon which our material is based is assumed to separate a binary or multi-component feed into two liquid products in a tray type distillation column that is equipped with a total condenser and a re-boiler. On the basis of the material ,we note that with the column on pressure control, there are four control variables viz x_D ,x_B ,h_D,h_B and four manipulated variables D,B, L_{N_T+1} ,and V .These are 4! That means 24 possible ways of pairing these variables. In single composition method three of these controlled variables are connected to three manipulated variables and the fourth variable is not adjusted. In dual composition control all four control variables are connected to four manipulated variables [13].

Chapter 4

Proposed work

4.1 Control strategy of distillation column

4.1.1 Control strategy of a 2×2 distillation column with a PID controller.

Here in this thesis we consider 2×2 distillation column process .The matrix transfer function of the distillation column process is

$$\begin{pmatrix} x_D(s) \\ x_B(s) \end{pmatrix} = \begin{pmatrix} \dfrac{12.8e^{-s}}{1+16.7s} & \dfrac{-18.9e^{-3s}}{1+21s} \\ \dfrac{6.6e^{-7s}}{1+10.9s} & \dfrac{-19.4e^{-3s}}{1+14.4s} \end{pmatrix} \begin{pmatrix} R(s) \\ S(s) \end{pmatrix}$$

Where x_D and x_B are percentage of methanol in the distillate and percentage of methanol in the bottom products respectively. *R(s)* and *S(s)* are reflux flow rate and stream flow rate in the boiler respectively [1].

A 2×2 distillation column is a MIMO process with strong interaction between two input and output pairs. The four transfer function in the multivariable process of first order dynamics and significant time delays. Here the objective of control the outputs y_1 and y_2 by control inputs u_1 and u_2.

A multivariable PID control for the distillation column system is designed using the decentralized relay feedback (DRF) method. The diagonal and off diagonal elements of the controller are designed in PI and PID forms respectively. This controller is as,

$$C(s) = \begin{pmatrix} 0.18 + 0.047\dfrac{1}{s} & 0.01 - 0.023\dfrac{1}{s} + 0.008s \\ 0.067 + 0.016\dfrac{1}{s} - 0.054s & 0.07 - 0.016\dfrac{1}{s} \end{pmatrix}$$

The integral absolute error

$$\begin{aligned} IAE &= \int_0^\infty |e_{11}(t)|\,dt + \int_0^\infty |e_{12}(t)|\,dt + \int_0^\infty |e_{21}(t)|\,dt + \int_0^\infty |e_{22}(t)|\,dt \\ &= IAE_{11} + IAE_{12} + IAE_{21} + IAE_{22} \end{aligned}$$

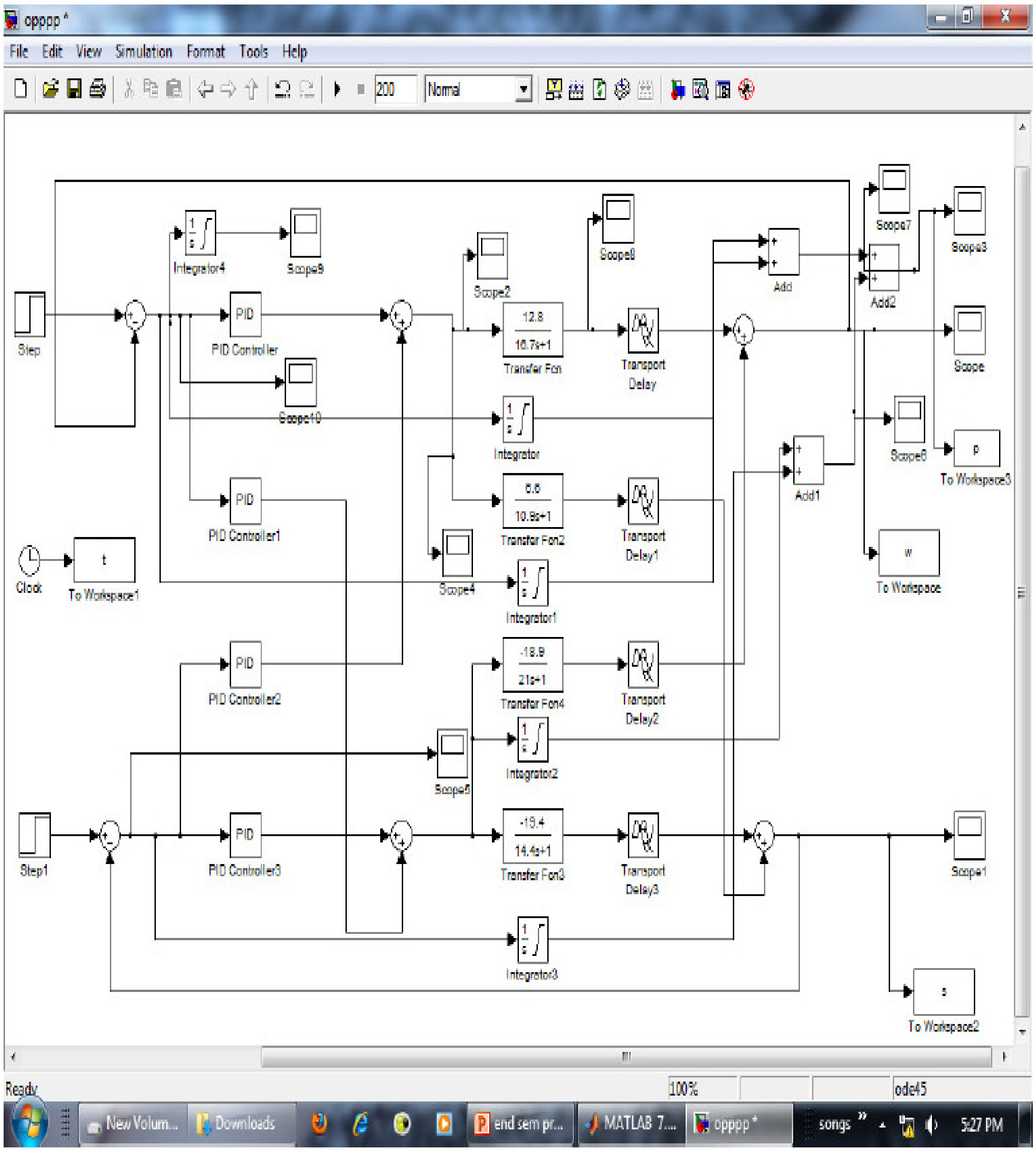


Figure 4.1 Simulated diagram 2×2 **system**

4.1.1.1 Step response of 2×2 distillation column process

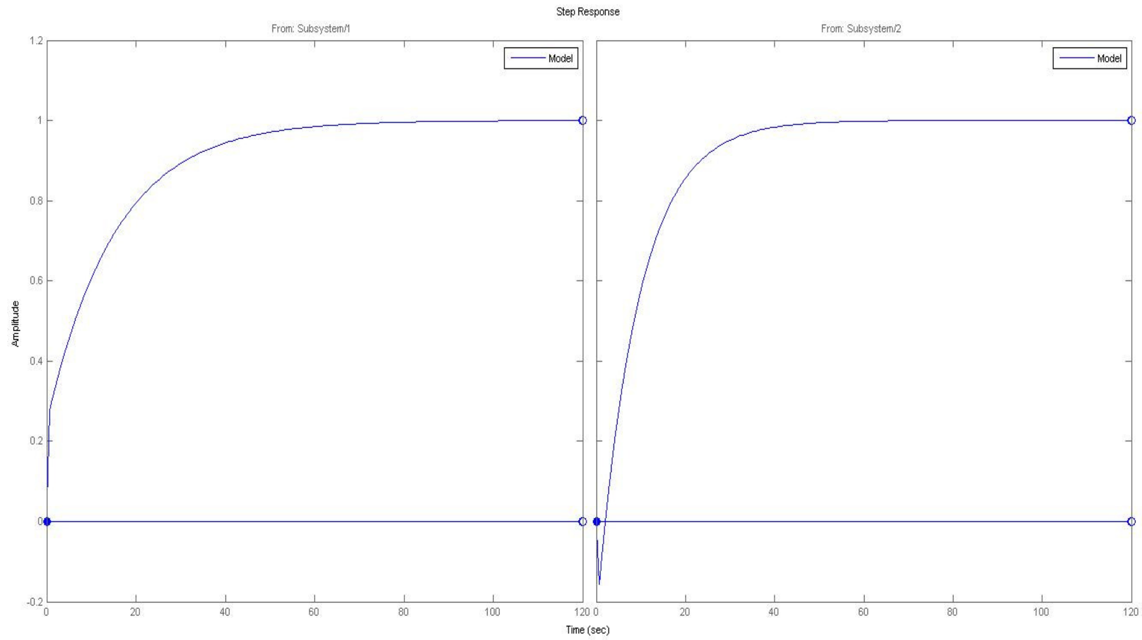


Figure 6.2 Step Responses

4.1.2 Control strategy of a 3×3 distillation column with a PID controller.

Here we are considering the process model of Ogunnaike-Ray [5] column model express the following equation,

$$G(s) = \begin{pmatrix} \dfrac{0.66e^{-2.6s}}{6.7s+1} & \dfrac{-0.61e^{-3.5s}}{8.64s+1} & \dfrac{-0.0049e^{-s}}{9.06s+1} \\ \dfrac{1.11e^{-6.5s}}{3.25s+1} & \dfrac{-2.36e^{-3s}}{5s+1} & \dfrac{-0.01e^{-1.2s}}{7.09s+1} \\ \dfrac{-34.68e^{-9.2s}}{8.15s+1} & \dfrac{46.2e^{-9.4s}}{10.9s+1} & \dfrac{0.87(11.61s+1)e^{-s}}{(3.89s+1)(18.8s+1)} \end{pmatrix}$$

For the controlling this process transfer function, PID controller is used here. Tuning the PID controller I am used BLT method. So the controller transfer function is,

$$C(s) = \begin{pmatrix} 0.96 + \dfrac{0.96}{8s} + 1.0464s & 0 & 0 \\ 0 & -0.2s - \dfrac{0.2}{6.5s} & 0 \\ 0 & 0 & 2.37 + \dfrac{2.37}{11s} + 7.6314s \end{pmatrix}$$

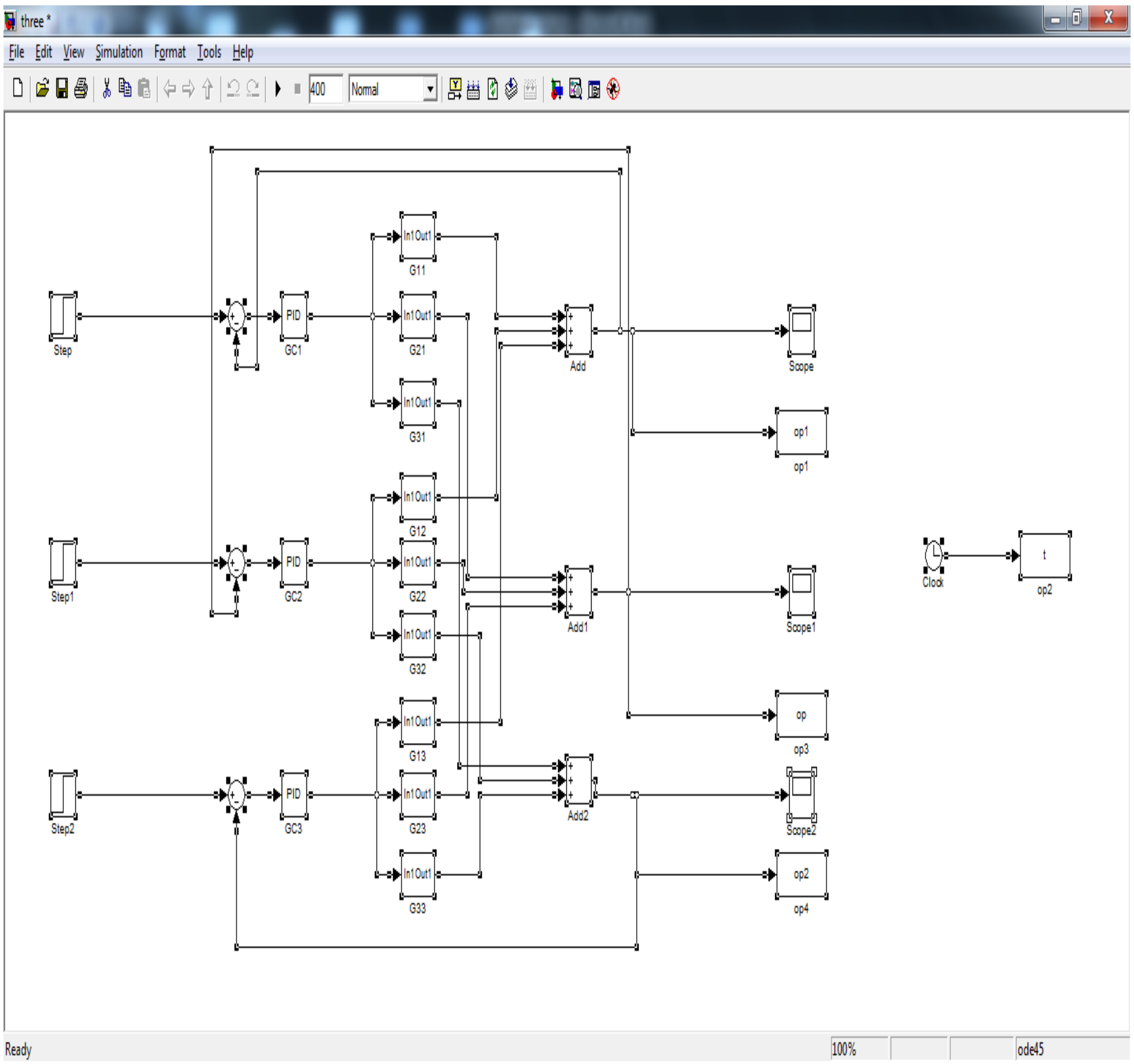


Figure 4.3 Simulated model of a 3×3 system

4.1.2.1 Step response of a 3×3 distillation column

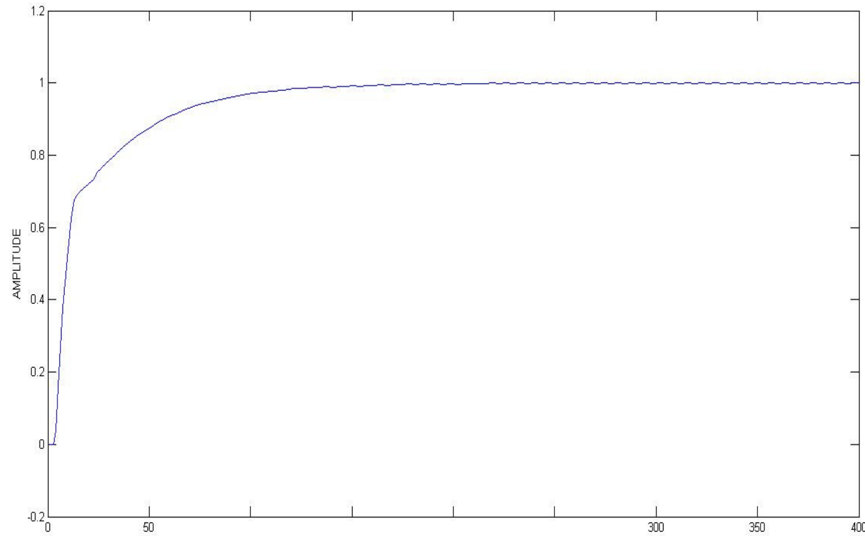


Figure4.7. A

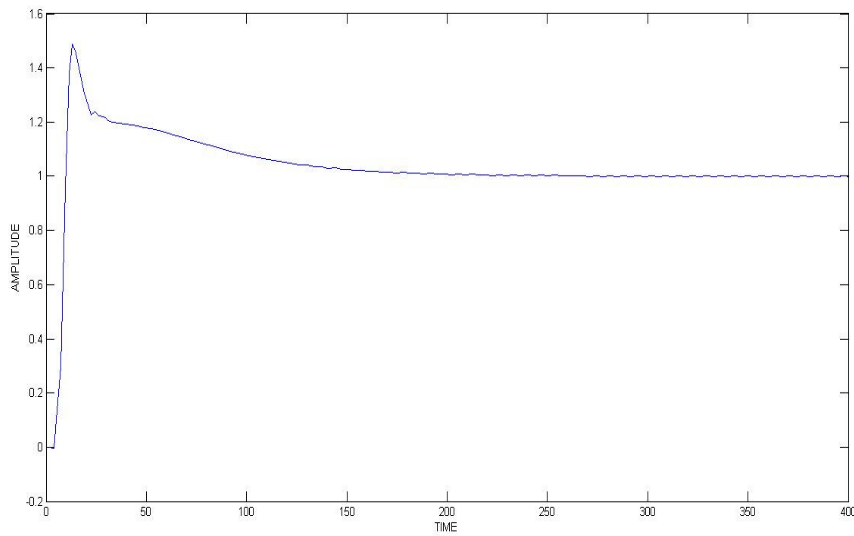


Figure4.5.B

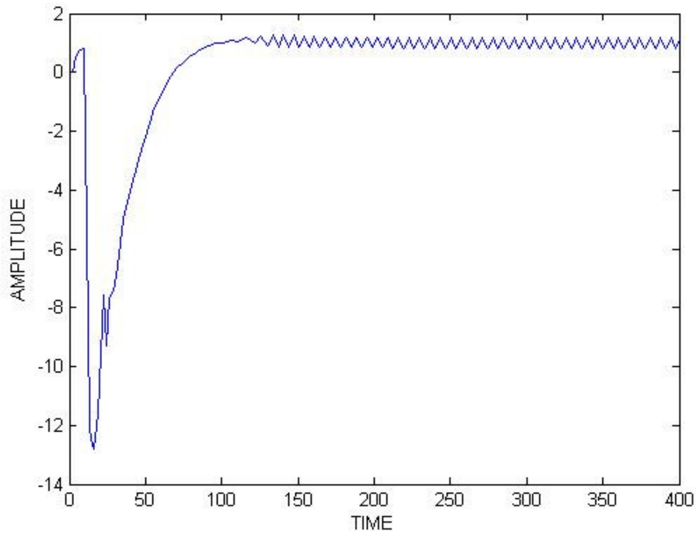


Figure4. 5. C

4.1.3 Control strategy of a 4×4 distillation column with PID controller.

Here it is considered 4×4 process by A1 column. They studied the dynamic of a distillation column. Here our objective is to maintain four composition specifications on the product stream [14]. The model is given by,

$$G(s) = \begin{pmatrix} \dfrac{2.22e^{-2.5s}}{(36s+1)(25s+1)} & \dfrac{-2.94(7.9s+1)e^{-0.05s}}{(23.7+1)^2} & \dfrac{0.017e^{-0.2s}}{(31.6s+1)(7s+1)} & \dfrac{-0.64e^{-20s}}{(29s+1)^2} \\ \dfrac{-2.33e^{-5s}}{(35s+1)^2} & \dfrac{3.46e^{-1.01s}}{32s+1} & \dfrac{-0.51e^{-7.5s}}{(32s+1)^2} & \dfrac{1.68e^{-2s}}{(28s+1)^2} \\ \dfrac{-1.06e^{-22s}}{(17s+1)^2} & \dfrac{3.511e^{-13s}}{(12s+1)^2} & \dfrac{4.41e^{-1.01s}}{16.2s+1} & \dfrac{-5.38e^{-0.5s}}{17s+1} \\ \dfrac{-5.73e^{-2.5s}}{(8s+1)(50s+1)} & \dfrac{4.32(25s+1)e^{-0.01s}}{(50s+1)(5s+1)} & \dfrac{-1.25e^{-2.8s}}{(43.6s+1)(9s+1)} & \dfrac{4.78e^{-1.15s}}{(48s+1)(5s+1)} \end{pmatrix}$$

The controlled and manipulated variables are y1(toluene impurity in bottom),y2(toluene impurity in the distillate),y3 (benzene impurity in the side stream),and y4 (xyene impurity in the side stream);u1(side stream flow),u2(reflux ration),u3(reboil duty) and u4(side draw location).

Here PID controller is used to control the above process. The tuning is done by the method,

$$C(s)=\begin{pmatrix} 2.1822+\dfrac{2.1822}{29.725s}+54.55s & 0 & 0 & 0 \\ 0 & 4.4807+\dfrac{4.4807}{8.08s} & 0 & 0 \\ 0 & 0 & 1.6656+\dfrac{1.6656}{8.08s} & 0 \\ 0 & 0 & 0 & 4.366+\dfrac{4.366}{9.2s} \end{pmatrix}$$

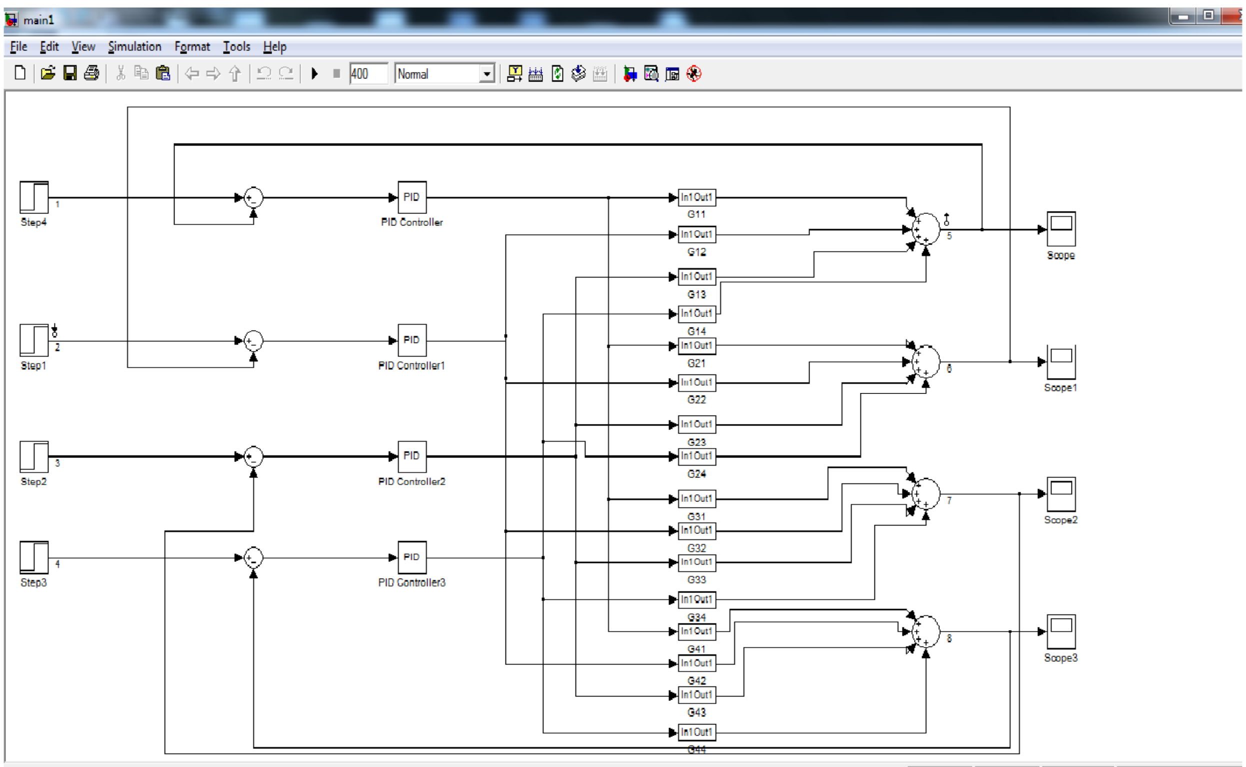


Figure4.6 Simulink design for a distillation column

4.1.3.1 Step response of a 4×4 distillation column

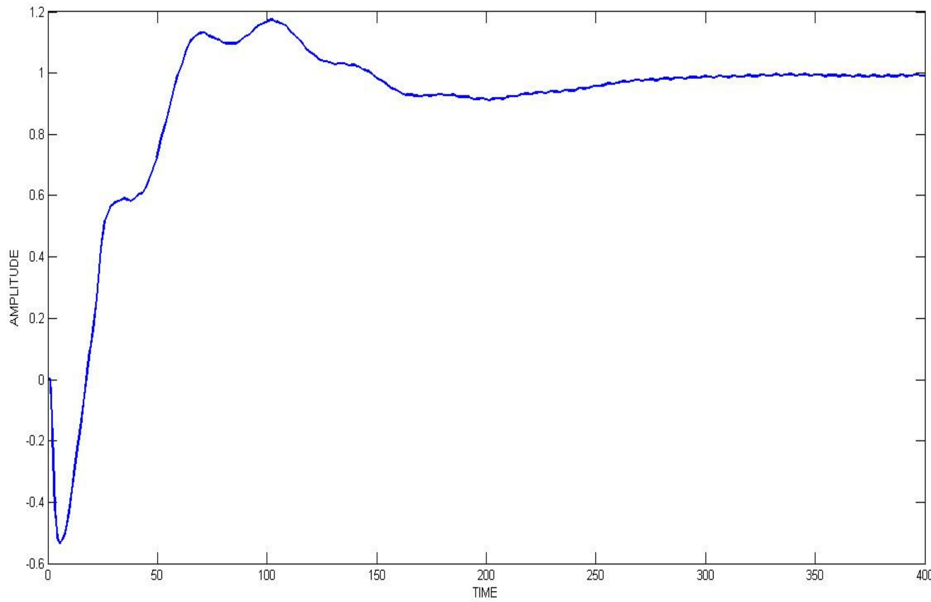


Figure4.8.A

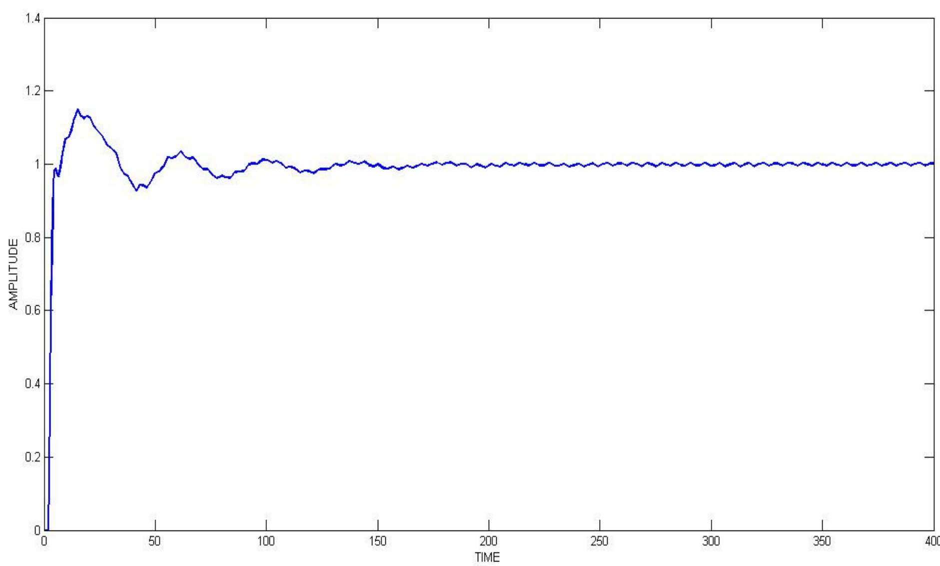


Figure 4.10. B

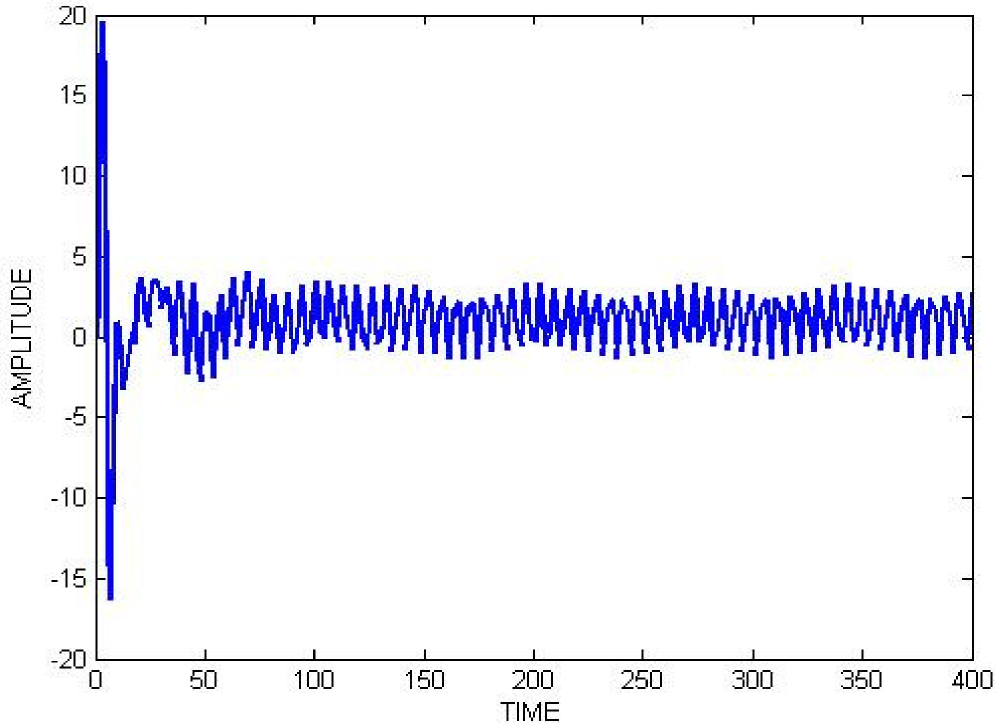


Figure4.10. C

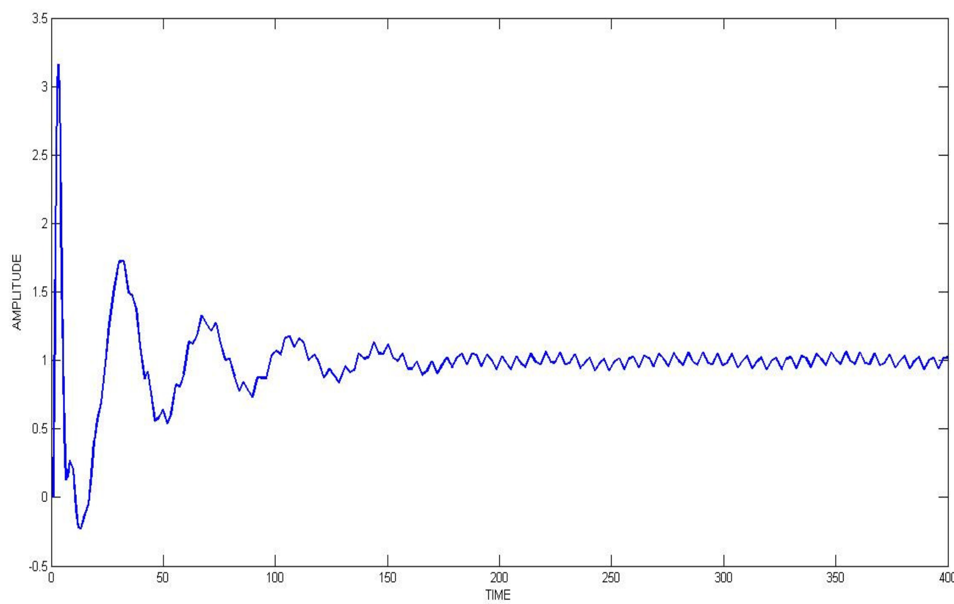
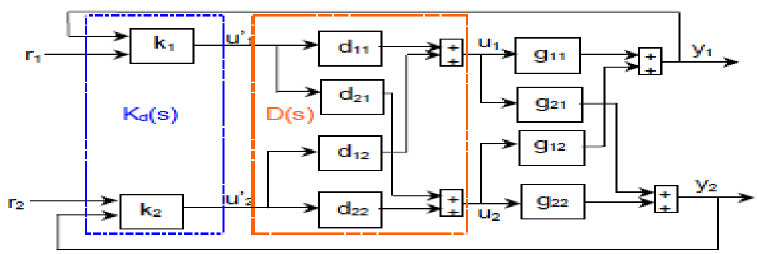


Figure4.10.D

4.2 Design of a decoupler for 2×2 distillation column process

The process transfer function is [4],

$$\begin{pmatrix} x_D(s) \\ x_B(s) \end{pmatrix} = \begin{pmatrix} \dfrac{12.8e^{-s}}{1+16.7s} & \dfrac{-18.9e^{-3s}}{1+21s} \\ \dfrac{6.6e^{-7s}}{1+10.9s} & \dfrac{-19.4e^{-3s}}{1+14.4s} \end{pmatrix} \begin{pmatrix} R(s) \\ S(s) \end{pmatrix}$$



Block diagram for 2*2 system with decoupler

Figure4.9

The decoupling matrix is given by,

$$D(s) = \begin{pmatrix} 1 & \dfrac{24.66s+1.48}{21s+1} \\ \dfrac{4.89s+0.34}{10.9s+1} & 1 \end{pmatrix}$$

The controller matrix is given by,

$$C(s) = \begin{pmatrix} 0.34+\dfrac{1}{0.06924s}+0.1156s & 0 \\ 0 & -0.046-\dfrac{1}{0.0048s}-0.0349s \end{pmatrix}$$

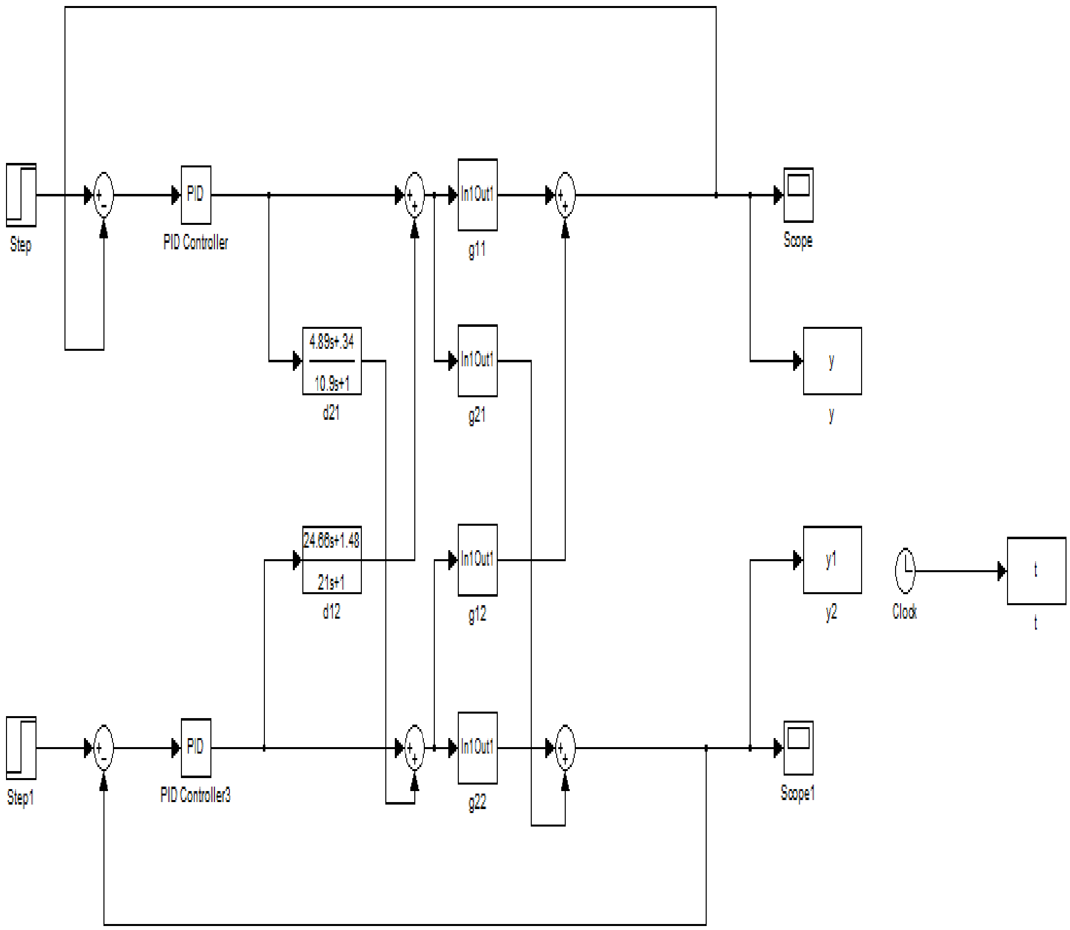


Figure4.10 Simulink diagram of a system with decoupler

4.2.1 Step response of a 2×2 distillation column with decoupler

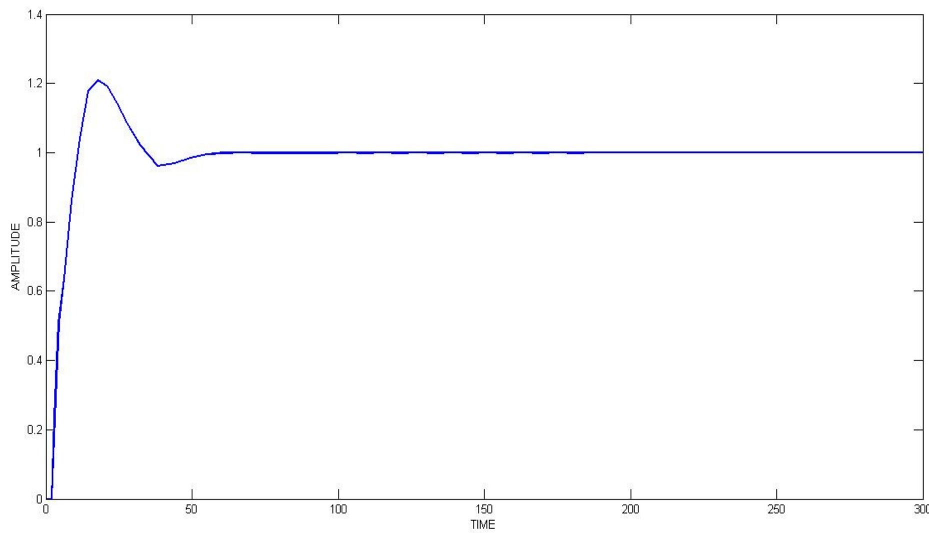


Figure4.11. A

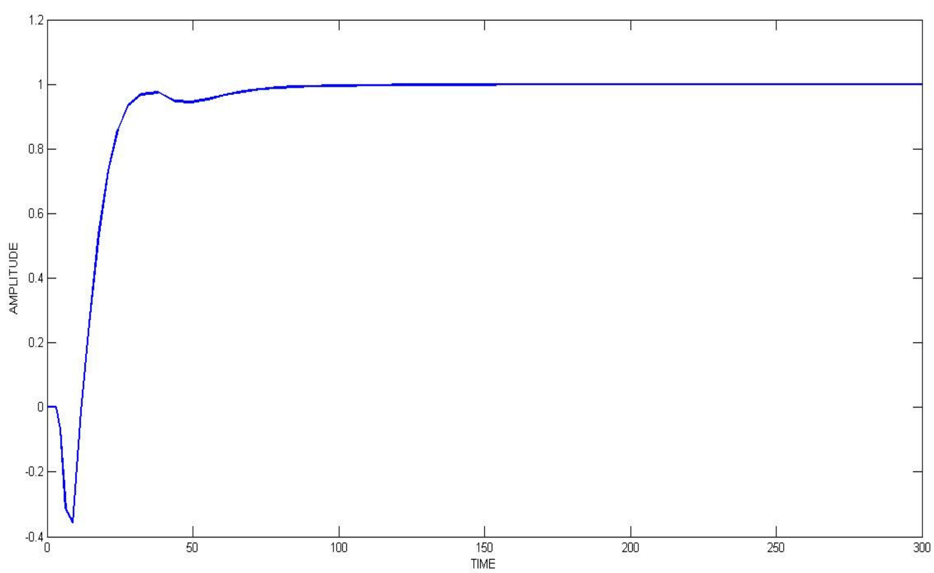


Figure4.16. B

4.2 Designing of a decoupler for a 4×4 distillation column

The decoupler matrix is used to minimize the interaction effect. The process is presented by Doukas and Luyban(1978) [14].They studied dynamic of a distillation column producing a liquid sidestream product.The objective is to maintain four composition specification on the product stream.The transfer function matrix is of (4×4) model is given below. The control and manipulated variables are y1 (toluene impurity in bottom), y2 (toluene impurity at distillate), y3 (benzene impurity in side stream), y4 (xyene impurity at side stream),u1 (side stream flow rate), u2(reflux ratio), u3 (reboil duty) and u4 (side draw location).

$$G(s) = \begin{pmatrix} \dfrac{2.22e^{-2.5s}}{(36s+1)(25s+1)} & \dfrac{-2.94(7.9s+1)e^{-0.05s}}{(23.7s+1)^2} & \dfrac{0.017e^{0.2s}}{(31.6s+1)(7s+1)} & \dfrac{-0.64e^{-20s}}{(29s+1)^2} \\ \dfrac{-2.33e^{-5s}}{(35s+1)^2} & \dfrac{3.46e^{-1.01s}}{32s+1} & \dfrac{-0.51e^{-7.5s}}{(32s+1)^2} & \dfrac{1.68e^{-2s}}{(28s+1)^2} \\ \dfrac{-1.06e^{-22s}}{(17s+1)^2} & \dfrac{3.511e^{-13s}}{(12s+1)^2} & \dfrac{4.41e^{-1.01s}}{16.2s+1} & \dfrac{-5.38e^{-0.5s}}{17s+1} \\ \dfrac{-5.73e^{-2.5s}}{(8s+1)(50s+1)} & \dfrac{4.32(25s+1)e^{-0.01s}}{(50s+1)(5s+1)} & \dfrac{-1.25e^{-2.8s}}{(43.6s+1)(9s+1)} & \dfrac{4.78e^{-1.15s}}{(48s+1)(5s+1)} \end{pmatrix}$$

The design of an ideal-diagonal decoupler problem was transformed to determine the decoupler

$$G_R(s) = G(s) G_I(s)$$

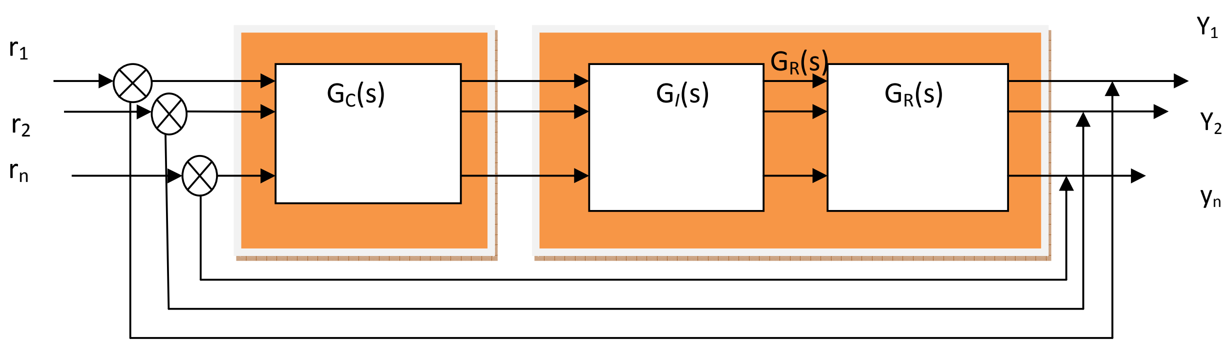


Figure4.12 Block diagram of a decoupling control system

$$K_N = K \odot T_{ar}$$

$$K_N = \begin{pmatrix} k_{N,11} & \cdots & k_{N,1n} \\ \vdots & \ddots & \vdots \\ k_{N,n1} & \cdots & k_{N,nn} \end{pmatrix}$$

$$\Phi = K_N \otimes K_N^{-T}$$

$$\Phi = \begin{pmatrix} \Phi_{11} & \cdots & \Phi_{1n} \\ \vdots & \ddots & \vdots \\ \Phi_{n1} & \cdots & \Phi_{nn} \end{pmatrix}$$

$$\Gamma = \begin{pmatrix} \gamma_{11} & \cdots & \gamma_{1n} \\ \vdots & \ddots & \vdots \\ \gamma_{n1} & \cdots & \gamma_{nn} \end{pmatrix}$$

$$G_I(s) = \hat{G}^T(s) G_R(s)$$

$$\widehat{G}^T(s) = G^{-1}(s)$$

$$G_I(s) = \begin{pmatrix} g_{I,11}(s) & g_{I,12}(s) \ldots g_{I,1n}(s) \\ g_{I,21}(s) & g_{I,22}(s) \ldots g_{I,2n}(s) \\ . & \\ . & \\ g_{I,n1}(s) & g_{I,n2}(s) \ldots g_{I,nn}(s) \end{pmatrix}$$

$$\Rightarrow \begin{pmatrix} 1/\hat{g}_{11}(s) & 1/\hat{g}_{21}(s) & \ldots 1/\hat{g}_{n1}(s) \\ 1/\hat{g}_{12}(s) & 1/\hat{g}_{22}(s) & \ldots 1/\hat{g}_{n2}(s) \\ . & & \\ . & & \\ 1/\hat{g}_{1n}(s) & 1/\hat{g}_{2n}(s) & \ldots 1/\hat{g}_{nn}(s) \end{pmatrix} \times \begin{pmatrix} g_{R,11}(s) & 0 & 0 \\ 0 & g_{R,22}(s & 0 \\ . & & \\ . & & \\ 0 & 0 & g_{R,nn}(s) \end{pmatrix}$$

To observe how the problem's definition and the design method of a normalized decoupling control system was different from the existing methods, each element of the process transfer function matrx. ETF [15] was represented, and the desired forward transfer function elements were of the form

$$g_{R,ii}(s) = \frac{e^{-\theta_{R,ii}s}}{\tau_{R,ii}s + 1} \quad i, \ j = 1, 2, \ldots .n,$$

Where $\tau_{R,ii}$ and $\theta_{R,ii}$ are the adjustable time constant and the dead time of $g_{R,ii}(s)$ respectively, the element in the ideal decoupler matrix has the form

$$g_{I,ij}(s) = \frac{e^{-(\theta_{R,ii} - \hat{\theta}_{ji})s}}{\hat{k}_{ji}} \times \frac{\tau_{ji}s + 1}{\tau_{R,ii}s + 1} \quad i, \ j = 1, 2, \ldots n$$

After determination $G_I(s)$ the parameters in the controller, $G_C(s)$, could be tuned individually for the corresponding elements of $G_C(s)$.Thus, the present SISO PID tuning methods could be directly applied to assurance the stability and performance of each loop. In this work, the GPM (gain and phase margin) method was implemented for putting into practice because of its simplicity and robustness [15]. Since each element in $G_R(s)$ was represented by a FOPDT model, the standard PI controller of the forms A diagonal matrix, $G_R(s)$ was specified such that holds. Consequently, the design of the normalized decoupler started from the obtained $\hat{G}^T(s)$ determined the diagonal forward transfer function matrix, $G_R(s)$, such that the decoupler, $G_I(s)$, from Eq. satisfied realizable conditions.

$$g_{c,ii}(s) = k_{p,ii} + \frac{k_{i,ii}}{s}$$

The closed-loop forward transfer function was defined as

$$g_{c,ii}(s) g_{R,ii}(s) = \frac{k_{i,ii}}{s} e^{-\theta_{RiiS}}$$

The individual elements could be derived as

$$K = \begin{pmatrix} 2.2200 & -2.9400 & 0.0170 & -0.6400 \\ -2.3300 & 3.4600 & -0.5100 & 1.6800 \\ -1.0600 & 3.5110 & 4.4100 & -5.3800 \\ -5.7300 & 4.3200 & -1.2500 & 4.7000 \end{pmatrix}$$

$$\Lambda = \begin{pmatrix} 3.4568 & -0.9689 & 0.0518 & -1.5397 \\ -3.3193 & 2.6946 & -0.7307 & 2.3554 \\ -0.1169 & 0.1175 & 2.3053 & -1.3059 \\ 0.9794 & -0.8432 & -0.6264 & 1.4902 \end{pmatrix}$$

$$\Phi = \begin{pmatrix} 1.4979 & 0.0220 & 0.0371 & -0.5570 \\ -1.0785 & 1.3103 & -0.3012 & 1.0695 \\ -0.0066 & -0.0021 & 1.6285 & -0.6199 \\ 0.5872 & -0.3303 & -0.3644 & 1.1075 \end{pmatrix}$$

$$\Gamma = \begin{pmatrix} 0.4246 & -0.0205 & 0.7395 & 0.3710 \\ 0.3207 & 0.4733 & 0.4286 & 0.4656 \\ 0.0546 & -0.0159 & 0.7195 & 0.4867 \\ 0.6148 & 0.4016 & 0.5965 & 0.7493 \end{pmatrix}$$

According to the RGA-NI-RNGA criterion, the input-output, Pairs are selected as 1-1/2-2/3-3/4-4,

The ETF parameters were,

$$\hat{K} = \begin{pmatrix} 0.6292 & 2.7308 & 0.3387 & 0.4262 \\ 0.6929 & 1.2498 & 0.7257 & 0.7314 \\ 8.8079 & 26.9076 & 1.9483 & 4.2242 \\ -5.9989 & -5.2532 & 2.0461 & 3.2339 \end{pmatrix}$$

$$\hat{T} = \begin{pmatrix} 25.8986 & -0.9701 & 28.5451 & 21.5158 \\ 22.4504 & 15.1454 & 27.4331 & 26.0717 \\ 1.8566 & -0.3810 & 11.6556 & 8.2742 \\ 35.6571 & 22.0884 & 31.3770 & 39.7103 \end{pmatrix}$$

$$\hat{L} = \begin{pmatrix} 1.0614 & -0.0010 & 0.1479 & 7.4192 \\ 1.6036 & 0.4780 & 3.2148 & 0.9311 \\ 1.2013 & -0.2064 & 0.7267 & 0.2434 \\ 1.5369 & 0.0040 & 1.6703 & 0.8616 \end{pmatrix}$$

Which gives,

$$\hat{G}^{T}(s) = \begin{pmatrix} \frac{25.8986s+1}{0.64}e^{1.0614s} & \frac{22.4505s+1}{0.6929}e^{1.6036s} & \frac{1.8566s+1}{8.8079}e^{1.2013s} & \frac{35.6571s+1}{-5.9989}e^{1.5369s} \\ \frac{-0.9701s+1}{2.7308}e^{-0.0010s} & \frac{15.1454s+1}{1.2498}e^{0.4780s} & \frac{-0.381s+1}{26.9076}e^{-0.2064s} & \frac{22.0884ss+1}{-5.2532}e^{0.004s} \\ \frac{28.5451s+1}{0.7257}e^{0.1479s} & \frac{27.4331s+1}{0.7257}e^{3.2148s} & \frac{11.6535s+1}{1.9483}e^{0.7267s} & \frac{31.3770s+1}{2.0461}e^{1.6703.s} \\ \frac{21.5158s+1}{0.4262}e^{7.4192s} & \frac{26.6717s+1}{0.7314}e^{0.9331s} & \frac{8.2742s+1}{4.2242}e^{0.2434s} & \frac{39.7163s+1}{3.2339}e^{0.8616s} \end{pmatrix}$$

Using the normalized decoupling control system design rules, the decoupled forward transfer function was selected as

$$G_R(s) = \begin{pmatrix} \frac{1}{28.5451s+1}e^{-7.4192s} & & & \\ & \frac{1}{27.4331s+1}e^{-3.2148s} & & \\ & & \frac{1}{11.6556s+1}e^{-1.2013s} & \\ & & & \frac{1}{39.7163s+1}e^{-1.6703s} \end{pmatrix}$$

Which gives a stable, causal and proper decoupler?

$$G_I(s) = \hat{G}^{T}(s)G_R(s)$$

$$G_I(s)=\begin{pmatrix} \frac{41.16s+1.5893}{28.5451s+1}e^{-6.3578s} & -\frac{32.40063s+1.4432}{27.4331s+1}e^{-2.0135s} & \frac{0.2107s+0.1135}{11.6556s+1} & -\frac{5.9439s+0.1667}{31.716s+1}e^{-0.1394s} \\ -\frac{0.3552s-0.3661}{28.5451s+1}e^{-7.4202s} & \frac{12.11825s+0.8001}{27.4331s+1}e^{-2.7368s} & -\frac{0.0141s-0.03716}{63.84s+1}e^{-1.4077s} & -\frac{4.2046s+0.1903}{58.49s+1}e^{-1.6663s} \\ \frac{3.9334s+1.3779}{28.5451s+1}e^{-7.2713s} & -\frac{27.4331s+1.3779}{27.4331s+1} & \frac{5.9362s+0.5132}{11.6556s+1}e^{-0.4746s} & \frac{15.335s+0.4887}{39.716s+1} \\ \frac{50.489s+2.3463}{28.5451s+1} & -\frac{30.9976s+1.3672}{27.4331s+1}e^{-2.2837s} & \frac{1.9587s-0.2367}{11.6556s+1}e^{-0.9579s} & \frac{12.2816s-0.3092}{39.716s+1}e^{-0.8087s} \end{pmatrix}$$

The modified ETFs are determined as ,

$$g_{13} = \frac{0.017e^{0.2s}}{(31.6s+1)(7s+1)}$$ \

$$g_{24} = \frac{1.68e^{-2s}}{(28s+1)^2}$$

$$g_{32} = \frac{4.41e^{-1.01s}}{16.2s+1}$$

$$g_{41} = \frac{-5.73e^{-2.5s}}{(8s+1)(50s+1)}$$

To apply the SIMC method consequently, the PID controllers are obtained as

$$g_{c,13}(s) = 0.02185 + \frac{2.54853}{s}$$

$$g_{c,24}(s) = 0.727145 + \frac{2.33354}{s} - 7.10071s$$

$$g_{c,32}(s) = 2.2204 + \frac{8.8422}{s}$$

$$g_{c,41}(s) = 2.2204e^{16} + \frac{9.3969e^{15}}{s}$$

4.2 .1 Step response of a 4×4 distillation column with decoupler

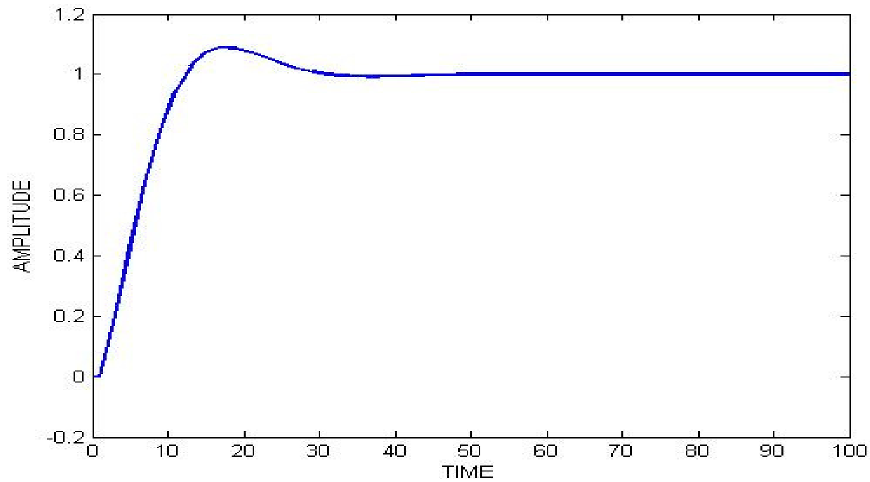


Figure4.13. A

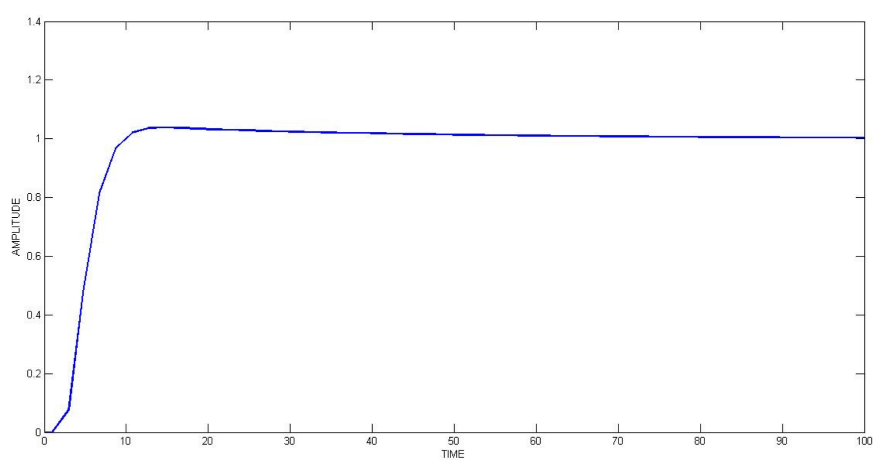


Figure4.19. B

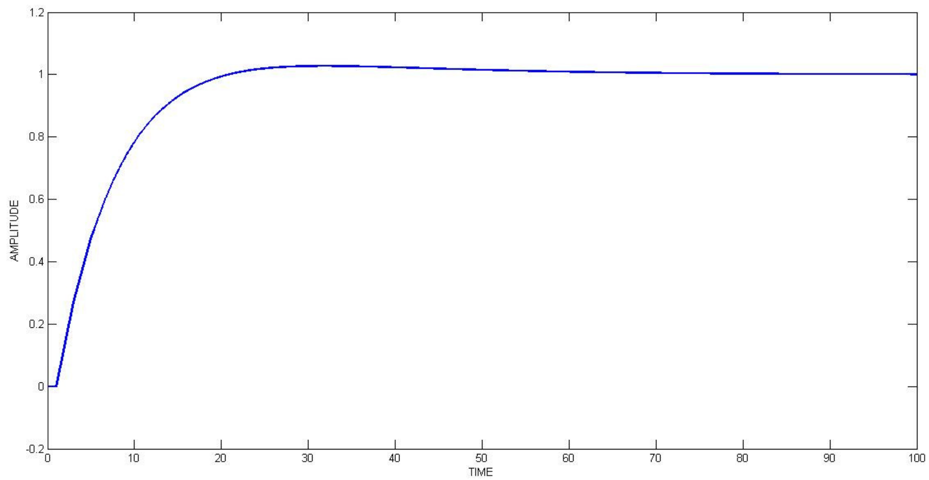


Figure 4.19.C

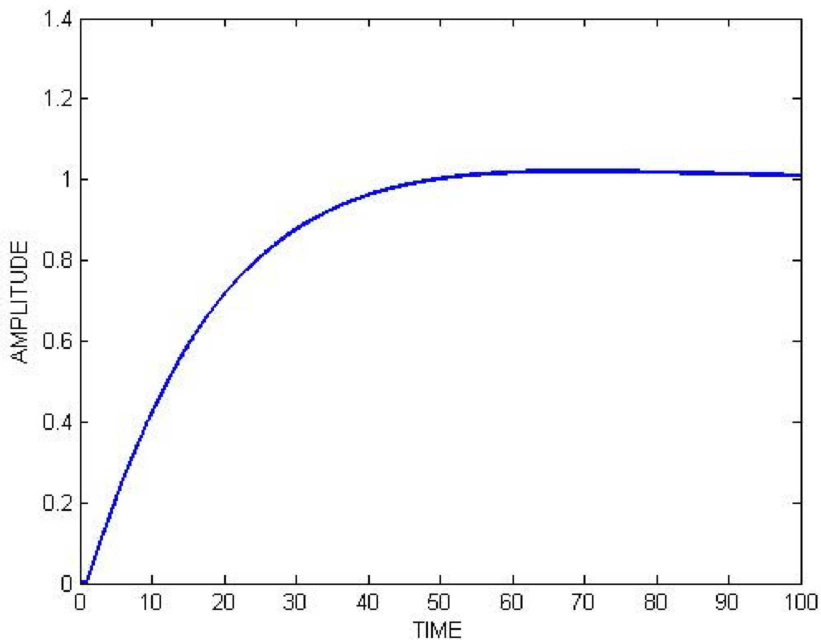


Figure4.19.D

Chapter 5

CONCLUSION AND FUTURE WORK

5.1 Conclusion

In this thesis all the distillation column processes such as ‘two input two output’, ’three input three output’ and ‘four input and four output’ systems have been studied, analyzed and simulated .The pairing between control variable and manipulated variables are done. Finally the main problem of the distillation column that is interaction between all the inputs and outputs have been eliminated and made the MIMO system into a simple SISO system with the help of appropriate decouplers, so that the PID controllers are designed independently to control the process. All the simulations are done by MATLAB tool in windows 7 operating system.

5.2 Future work

Distillation column is a multivariable process it has many non linearity. Linearization of distillation column is a very vast field to work. If we talking about Optimization of distillation column, it would be an interesting field to work on. In this thesis we worked on up to 4×4 distillation column process, 5×5 distillation column process may be good idea to work. Cascading of two or more distillation control and controlling its different parameters is an interesting field. New approaches to design decoupler and cost reduction are the few future works of distillation column process.

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