SIMULATION ON DYNAMIC AND CONTROL OF DISTILLATION COLUMN

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by

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CERTIFICATION

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ABSTRACT

The distillation column is one of the essential separation processes in modern chemical and petrochemical industries so simulation of dynamic and control of distillation column was conducted in this work.

For the control system, the important controlled variables were chosen to be the distillate composition (X_D), and the side stream composition (X_S). The manipulated variables are the reflux flow rate (R), and the side stream flow rate (L_S).

Tuning of Control Parameters was found by three methods; Zigler-Nicolas, Tyreus-Luyben and Process Reaction Curve methods to find the best values of Proportional gain (K_c), Integral time (τ_I) and Derivative time (τ_D).

P, PI and PID controllers were used as a control strategy for the system.

The degree of interaction was determined based on Relative Gain Array (RGA) and should be avoided by implementing a decoupling system.

The decoupling method was designed to eliminate the interaction effects between the control loops.

Fuzzy logic control system was used as another strategy to compare with conventional control system. For all cases, the Fuzzy logic controller is found to be preferable.

The frequency response method (bode diagram) was used in this work to study the stability of the system and it was found that a stable system is recommended.

MATLAB program was used as a tool of solution for all cases used in this work.

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NOMENCLATURE

Symbol	Definition	Unit
$\angle G(jw)$	Phase angle of the open-loop system	Degree
D	Transfer function of decoupler	-
de	Change in error	
dt	Change in time	s
$G(jw)$	Open-loop Transfer function	-
$G(s)$	system transfer function	-
G_C	Transfer function of controller	-
Gm	Gain margin	dp
K	Steady-state gain	
K	Constant of equation 3.4	
K_C	Controller gain	
Ku	Ultimate gain	
L_S	Side stream flow rate	lb.Mole/s
Pm	Phase margin	degree
Pu	Ultimate period	s
Pu	Ultimate period	s
QB	reboiler heat duty	BTU/s
QBS	stripper reboiler heat duty	BTU/s
R	Reflux flow rate	lb. moles/s
s	Laplace form	-
t	Time	s
td	Time delay	s
$u_{n(t)}$	step function	
X_D	Distillate composition	-
X_S	Side stream composition	-

Greek litters

Symbol	Definition	Unit
β_{ij}	ijth element in the relative gain array	-
λ_{ij}	Elements of relative gain array	-
μ	Membership function	-
τ	Time constant	s

τ_D	Derivative time	s
τ_I	Integral time	s

Abbreviations

Symbol	**Definition**
AE's	Algebraic equations
AR	Amplitude Ratio
BLT	Biggest Log Modulus Tuning
dB	Decibels
FLC	Fuzzy Logic Controller
GUI	Graphical User Interface
MATLAB	Matrix Laboratory
MIMO	Multi-Input/Multi-Output
NB	Negative Big
NCB	Negative change of Error Big
NEB	Negative Error Big
NES	Negative Error Small
NS	Negative Small
NUB	Negative control action Big
NUS	Negative control action Small
ODE's	Ordinary Differential Equations
P	Proportional Controller
PB	Positive Big
PCM	Process Control Modules
PES	Positive Error Small
PEB	Positive Error Big
PI	Proportional-Integral controller
PID	Proportional-Integral-Derivative controller
POR	Peak overshoot ratio
RGA	Relative Gain Array
SISO	Single-Input/Single-Output
SSS	Side Stream Column-Stripper

ZC	Zero change of error
ZE	Zero Error
ZN	Zeigler-Nicholas
ZU	Zero control action

Chapter One

Introduction

Distillation column is one of the most common forms of separation processes in modern chemical and petroleum plants, so problem of predicting the behavior of multistage distillation columns has been the basis of many studies [1].

The separation of liquid mixtures into their various components is one of the major operations in the process industries, and distillation is the most widely used method of achieving this separation and it is the key operation in any oil refinery. The vertical cylindrical column provides, in a compact form and with the minimum of ground requirements, a large number of separate stages of vaporization and condensation [1].

Distillation columns exhibit long time lags in internal and external flow of liquid and vapor. A change in reflux ratio is transmitted from tray to tray inside the column at the rate which depends upon the hold-up on each tray and the capacity of the downcomer. Similarly any change in feed rate and composition cannot be transmitted instantaneously through the column. The controllability of the column is largely affected by the long time lags, particularly for rapid input disturbances.

Distillation column is a multi-input/multi-output (*MIMO*) process which is more difficult to analyzed and operates than single-input/single-output (*SISO*) process.

Distillation columns were the subject of many dynamic and control studies because of their unique and challenging control problems. These systems have many troublesome features from a control standpoint: slow dynamic response, high order behavior, significant dead times, nonlinearity, and multivariable interaction. The stable and reliable performance of one or more distillation columns is imperative for safe and economic operation of many plants.

The conventional method of process control is to use feed-back control loops employing one, two or three terms controllers. Feed-back system give satisfactory control for a wide range of processes. The feed-back gives poor performance in processes which involve considerable time lags between the measurements of the controlled variable and the effect of the control action. If the lag is large and the controlled variable fluctuates quite rapidly, by the time the control action takes effect, it may well augment the disturbance rather than reduce it. Even if the disturbance is reduced eventually the controlled variable can be outside specification limits for a considerable time. [2].

Other control technique involves decoupling control it which is applied to multivariable processes, where there is interaction between control loops. This technique eliminates the effect of this interaction by designing suitable decouplers for the loops. It requires a wide knowledge of the dynamic behavior of the controlled variables for change in disturbance and in the manipulated variables [3].

Mathematical modeling and simulation are important and useful areas of computer application. In the mid 1970, engineers in industry were skeptical of simulation as a valid way to solve manufacturing problem, then an accurate predictions from mathematical models was believed. However the prevailing view in industry is that it is much less expensive and more reproducible to run simulation experiment than it is to perform repeated experiments involving actual equipment [4].

The computers play an important role in the design of modern control systems. The computer and software can remove the hard work from the task. At computer's desktop, to performance analysis, design, and simulation with one program, with the ability to simulate a design rapidly, it can easily make changes and immediately test a new design. A computer model of the system behavior may be utilitized to investigate various designs of a planned system without actually building the system itself [4].

Assuming that the model and simulation are reliably accurate, computer simulation has the following advantages:

1- Reduced cost.

2- Flexibility in responses to design changes.

3- Noise immunity.

4- System performance can be observed under all conceivable conditions.

5- Trial of system under test can be accomplished in a much-reduced period.

6- Simulation result can be obtained at lower cost than real experimentation.

7- Computer modeling and simulation is often the only feasible safe technique to analyze and evaluate a system.

The advantage of using computer is that many loops can be controlled or compensated by the same computer through time-sharing. Further, more, any adjustments of compensator parameters required to yield a desired response can be made by change in software rather than hardware.

The confidence level in what can be done with simulation has risen considerably, and this is having a deep influence on the practice of process engineering. The process control modules (PCM) area set programs written in the MATLAB SIMULINK environment (MATLAB7 and SIMULINK) (MATLAB7) which is a software packages that enables many of calculation associated with control systems including matrix computation. There are many specialized built-in function for such things as eigen value computation and bode analysis; the model based on fundamental process model of industrial unit operation (distillation column) [5].

❖ Scope of the Present Work

1. Study the dynamic of open and close loop systems of distillation column and the process control for different strategies.

2. Study the stability of the system using frequency response analysis (Bode diagram)

3. Relative gain array (RGA) will be used as an interaction measurement to decide the pairing of the control loops.

4. Decoupling control will be applied to the two point composition control scheme.

5. A tuning procedure will be carried out to select the best set of system parameters.

6. The control strategies, feedback, and fuzzy logic controllers are applied to the schemes to control distillate product, and liquid flow rate of the side stream product

Chapter Two

Literature Survey

A typical simple distillation column separates a mixture of chemical component into two product streams the lighter component at the top and the heavier component at the bottom. The separation achieved in a particular column depends upon the number of trays in the column, the reflux ratio, the relative volatilities and the way the feed is split between overhead and bottom product.

Complex industrial distillation column is difficult to control automatically. This difficulty is due to their non-linear time varying behavior and the poor quality measurements available. In such cases automatic control is applied to those subsidiary variables such as temperature, pressure and flow rates which can be measured and controlled. The overall process control objective, such as the quality and quantity of produced product, has in the past been left in the hands of the human operator [1].

For an existing operating column where the number of trays is fixed and the pressure is held constant so that the relative volatilities are fixed, the product composition can be controlled by only two variables, (a) the product split (i.e. overall material balance) and (b) reflux ratio (i.e. the energy input).

However, controllers based on microcomputers instead of human operators were developed for different industrial plants. In many cases a computer gave faster response and more accurate control than a skilled human operator [3].

2.1 Simulation on Dynamics of Distillation Column

The dynamic and steady state simulation models of distillation columns consist of a system of equations based on mass and energy balances around each plate of the column.

Typically for the dynamic problems, these balances lead to system of ordinary differential equations (*ODE's*) or to mixed system of equations (*ODE's* and Algebraic Equations *(AE's)*) [1].

There are many studies on computer applications to distillation calculations. This is superficially due to the repetitive nature of these calculations, which render them suitable for solution by computer. Most of the earlier studies; Lapidus and Amundson [6], Rose and Jahnson [7], Armstrong and Wood [8], dealt with only one section of the column, and the problem was more difficult when both the stripping and enriching sections were considered.

Morris and Sevreck [9] developed a simulator for multicomponent distillation which had a highly modular structure and an explicit integration scheme.

Thomas [10] reviewed the digital dynamic solution of distillation processes, and presented a new approach which allows for consideration of the effect of varying vapor holdups. He was found that calculation of the instantaneous calculation of the instantaneous component boil-off rates was reduced to the problem of solving a set of linear simultaneous equations of the same orders as the number of components present.

Berber [11] developed a simulation program to predict the dynamic behavior of a theoretical distillation column fractionating a three component feed mixture. Berber and Ates [12] used a dynamic mathematical model based on the previous study [11], to predict the transient response of a continuous stage wise distillation column.

Chimowitz et al. [13], presented an algorithm using local thermodynamic and physical property models in dynamic simulation of multicomponent distillation column. The dynamic models used were relatively simple, but provided a good description of the dynamics of many distillation processes. The local model concept, however, could be interfaced with any dynamic model of distillation, regardless of its complexity.

Gani et al. [14] presented a generalized model for the dynamic simulation of distillation columns. The successful application of the model to solve different type of test problems demonstrated its wide applicability and flexibility. The good matching of the industrial data showed that the model was reliable and could be used for the study of industrial processes. Even where the industrial data were not available, the results obtained when analyzed qualitatively, seemed to be varying reasonability.

Ranzi et al. [15] analyzed and discussed the role of energy balances in simulating the transient behavior of multicomponent distillation columns. A few examples were considered to show how big discrepancies could be observed as a result of neglecting the time derivative of the energy balance hold-ups. Comparisons were presented to show the possible important benefits related to a simultaneous solution of the whole system of algebraic and differential equations. The results showed that the enthalpy balance equations should be taken into account.

2.2 Control Strategies of Distillation Column

Distillation column is a Multi-Input/Multi-Output (*MIMO*) process which is more difficult to analyze and operates than Single-Input/Single-Output (*SISO*) process.

2.2.1 Feedback Controller

Feedback control is achievement and maintenance of a desired condition by using an actual value of this condition and comparing it with a reference value, then using the difference between these two values to eliminate any difference between them.

Hu and Fredramirez [16] applied control theory for distillation column control. Both linear and non-linear distillation models were developed and tested. They achieved a good result by using an optimal multivariable Proportional integral (*PI*) controller for systems with unmeasurable disturbances. When the disturbances were measurable, an optimal multivariable proportional controller with error coordination was recommended. Their results showed that due to the non-linear behavior of the system the multivariable proportional controller algorithm forced the top and bottom compositions near the original steady-state with some offset.

A simple practical approaches to the problem of finding reasonable *PI* controller settings of the *N* single-input single-output controllers in an *N*th order typical industrial multivariable process was presented by Luyben [17]. The procedure was straight-forward extension of the familiar Nyquist method and required only nominal computing power. The method was tested on ten multivariable distillation column examples taken from the literature. The resulting settings gave reasonable and stable responses.

Al-Elg and Palazoglu [18] developed a rigorous dynamic model of a high purity double effect distillation column to study loop interaction as well as the impact of modeling errors on the effectiveness *PI* controller. Their results indicated the severity of interactions and control performance degradation associated with high purity specifications.

Anderson [19] used a frequency domain approach to compare the nominal performance and robustness of dual-composition distillation column control tuned according to Ziegler-Nichols (*ZN*) and Biggest Log Modulus Tuning (*BLT*) described by Luyben [17] for three binary distillation columns. The scope of their work was to examine whether *ZN* and *BLT* designs yield satisfactory control of distillation columns. Further, *PI* controllers were tuned according to a proposed multivariable frequency domain method.

2.2.2 Decoupling Control

Being the major energy consumers in a chemical plant, distillation columns offer most challenging design and control problems [20]. In order to save energy, dual composition control has been proposed and its merits have been extensively studied by Luyben [21] and Shinsky [22]. However, control of both top and bottom compositions usually results in undesirable interaction among the control loops. To be able to cope with such interactions, a lot of research effort has been devoted to decoupling control (Luyben [23], Wood and Berry [24], Ryskamp [25], Waller [26], Schwanke [27], Shinsky [21], Jafary and McAvoy [28], McAvoy [29], Weischedel and McAvoy [30] Fagervik [31]). Because of the simplicity and transparency of the design procedure decoupling is the most popular control strategy in distillation however the need for accurate models to active decoupling limits the success of the method. [15]

Luyben [23] presented a quantitative study of two types of decoupling elements to achieve a non-interacting feedback control of overhead and bottom compositions in binary distillation, ideal and simplified decoupling. It was concluded that in ideal decoupling, leading to unstable feedback loops, while simplified decoupling was effective and stable and appeared to be easily implemented with commercial control instrumentation.

Weischedel and McAvoy [30] studied the two variables controls using the reflux and boil up as manipulated variables to control both top and bottom compositions. They concluded that complete decoupling was not feasible for many (high purity) columns due to sensitivity to model error.

Frey et al. [32] examined the control variable pairing by using the relative gain array. Their results indicated that decoupling of the column was possible by proper selection and pairing of the manipulated and control variables. It was shown that the column decoupled at steady-state compositions independently to design parameters.

Arkun and Morgan [33] gave numerical robustness analysis for low purity and high purity columns with no decoupling, simplified decoupling and ideal decoupling control schemes.

2.2.3 Fuzzy Logic Controller

The plant models whether they are based on physical and chemical relationships or parameter estimation methods are approximations to the real process and may require a large amount of computer time. Some successful applications have been reported, but difficulties have been experienced where process operator works over a wide range of conditions and suffers from stochastic disturbances.

King and Mamdani [34] stated that an alternative approach to the control of the complex processes is to investigate the control strategies employed by the human operator. In many cases the process operator can control a complex process more effectively than automatic control when he experiences difficulty this can often be attributed to the rate of or manner of information display or the depth to which he may evaluate decisions.

Mamdani and Assilian [35] realized that the fuzzy logic controller could be used not only in treatment of complex systems, but it could be applied to hard systems such as industrial plants controllers. A small boiler steam engine controller specifying heuristic fuzzy control rules for two feedback loops as implemented. The results showed that the quality of control with the fuzzy controller was found to be better than the best control obtained by the fixed controller.

Tong [36] gave a good review of the work done on fuzzy logic controllers and their application. A brief description of fuzzy set theory and its use and advantages was presented.

Mamdani [37, 38] applied the fuzzy logic approach to control a steam engine. The controller was actually composed of two separate algorithms for the two control loops. Each algorithm could base the decision on all four output variables and thus coped with the interactive nature of the plant. Results showed that this approach could give similar, if not better, results compared with classical controllers, and a control low similar to the two term PI controller. It was concluded that this method is chiefly applicable in the control of plants that are difficult to model, such as those in the cement, chemical or iron and steel industries.

King and Mamdani [35] applied the fuzzy logic controller to boiler and temperature control of stirred tank and they concluded that processes could be controlled effectively using heuristic rules based on fuzzy statements. The designer requires some knowledge of the process in formulating the rules for instance: process delays, speed and magnitude of responses, but only approximate values are required and can be obtained by operating the process. They concluded that fuzzy control system was much less sensitive to process parameter changes and gave good control at all operating points.

Mamdani [39] surveyed the research work done on fuzzy controllers and briefly discussed its application to cement kilns. The heuristic continue to maintain the PI nature of the classical control only making it non-linear and the controller was robust to plant parameter changes. Concerning stability problem, fuzzy controller can be analyzed qualitatively to gain assurance that a runaway instability will not occur. Confidence in the quality of control can always be obtained by running it on an open loop with the human operator present to make changes.

Umber and king [39] concluded that a fuzzy controller is essential to this application. Larsen [40] claimed that reduction in fuel consumption was obtained be fuzzy control. Mamdani et al. [41] stated that rule based control methods are being used commercially for the control of cement kilns.

Kickent and Lemke [42] applied three types of fuzzy controllers to control the temperature of warm water plant, and compared the results with normal PI controller. Continuous type membership function was used to describe the fuzzy sets. The process had difficult control properties, arising from nonlinearities, a symmetric behavior for heating and cooling, noise and dead times. Also the ambient temperature influenced the process behavior. The three controllers were similar to, PI, I and P respectively.

Tong et al. [43] applied fuzzy controller to a sludge water treatment process. The process is usually controlled manually with many problems. They considered the controller to reflect actual operational practice and they concluded that the algorithm did rather well and it could be useful for such processes.

Ray and majumber [44] designed a set of fuzzy controllers for a non-linear multivariable steam generating unit (200 MW) which was decoupled using the output feedback theory. They concluded that good response was obtained even in case of improper decouplers or system parameter variation because of the robustness of the process, but poor responses were obtained if no decoupler was used.

Sicking et al. [45] designed a fuzzy supervisor with a PID controller. A great improvement was noticed when applied to different systems.

Ali [46] studied the different features of fuzzy controllers for different simulated systems, and then the controller was applied to a lab temperature control process. The results showed that fuzzy controller can give as good, if not better results as PID controller in spite of the limit cycle which could be reduced or eliminated proper controller tuning.

Lin and Lu [47] studied an active vibrating compensation via a two plate platform. They concluded that the designed fuzzy logic controller possesses the following features: robustness, ease of design, high speed of response, adaptability of adjustment of rules, and it is readily implementable by microelectronic devices.

Naoum [48] designed a rule-based fuzzy logic controller for a binary distillation column separating methanol-water mixture. The results obtained when applying fuzzy logic controller to the control of top and bottom temperature were as good, if not better, than those obtained using PI controllers. Fuzzy decouplers were proposed as an alternative to other types of decouplers namely simple steady state decouplers and the results were very promising compared with PI controllers and fuzzy controller without decoupler.

Al-jibory [49] designed and evaluated a rule-based controller that incorporates fuzzy logic controller to control a continuous stirred tank reactor to carry out the reaction of methylacetate with water. The effect of interaction between control loops was eliminated using the decoupling technique. A comparison was made between fuzzy and PID controllers to test the effectiveness of these controllers on behavior of the system. His results indicated the priority of the fuzzy controller which gave better results compared with a PID controller.

Faroq [50] used fuzzy logic control system as another strategy to compare with conventional control system. For all cases, the Fuzzy logic controller was preferable because it did not require an accurate mathematical model to be built for the process. On the other hand, all other control strategies used needed a wide knowledge of the process dynamics and an accurate mathematical model to be built and solved. In addition Fuzzy logic control gave lower value of ISE when compared with optimized PID control.

2.3 Frequency Response Methods

Frequency response methods were developed by the Nyquist [51] and Bode [52]. This method is the most conventional methods available to control engineer for analysis and design of control [53]. Pollard [54] described the frequency response method which can be applied experimentally to existing systems by a use of a suitable sine wave generator to inject a sinusoidal signal into either open or closed-loop, and in this way the frequency response characteristics of the complete system or of the elements of a system can be obtained when the transfer functions are not known. The advantages that they generally possess which make them very practical are the following [4, 55]:

1- The frequency response methods can be worked on limited amount of experimental data.

2- The frequency response methods cope easily with on-line tuning requirements.

3- The frequency response method is ready available of sinusoid test signals for various ranges of frequencies and amplitudes. Thus the experimental determination of the frequency response of a system is easily completed and is the most reliable and uncomplicated method for the experimental analysis of a system. Often as we shall find, the unknown transfer function of system can be deduced from the experimentally determined frequency response of a system.

4- The frequency response method is that the transfer functions describing the sinusoidal steady–state behavior of a system can be obtained by replacing s with jw in the system transfer function $G(s)$. The transfer function representing the sinusoidal steady–state behavior of a system is then a function of a complex variable jw and is itself a complex function $G(jw)$ that possesses a magnitude and phase angle. The magnitude and phase angle of *G(jw)* are readily represented by graphical plots that provide significant insight into the analysis and design of control systems.

The basic disadvantages of the frequency response method for the analysis and design are the indirect link between the frequency and the time domain, except for case second order system. [53]

2.4 Control of a Complex Sidestream Column-Stripper Distillation Configuration [56]

The dynamic and control of a complex, multivariable, interaction side stream column-stripper distillation configuration (SSS) were explored via digital simulation. It was found to be controllable by using two conventional SISO controllers.

Therefore, the dynamic operability of these more interacting and more multivariable processes has yet to be firmly established. Single sidestream columns have been studied by Tyreus and Luyben [57] and ogunnaike et al. [58]. Prefractionator schemes have been studied by Doukas and Luyben [59], Lenhoff and Morari [60], and Elaahi and Luyben [61].

The dynamic and control of distillation system is used for separating ternary mixture that contain small amount (less than 20%) of the intermediate component in the feed.

The SSS system presents a challenging 2X2 multivariable control problem. Significant questions must be addressed concerning tuning, stability and control systems structure.

2.4.1 Steady State Design

The ternary system benzene / toluene / o-xylene was chosen as a typical industrially important separation. Benzene and xylene product purities of 95 mole% and toluene product purity 90 mole% were used. The steady state value of reflux flow rate is (445.14lb.mole/hr) and the side steam flow rate is (106.25lb.mole/hr).The steady state design of the complex SSS configuration is given in **fig. 2.1**

The SSS configuration was designed by using an evolutionary procedure. Detailed design is given by Alatiqi and Luyben [62] for the system.

A major aspect in the steady-state design of the SSS system is the amount of liquid sideraw rate (LS) fed to the stripper. The higher this rate, the lower the total energy consumption (main reboiler heat duty QB plus stripper reboiler heat duty QBS). However, there is a limiting value of LS beyond which the purity of the toluene product from the stripper base can no longer be attained. This is due to the increase in the heaviest component (xylene) around the sideraw tray as LS is increased. Any xylene that inters the stripper leaves in the toluene product. Therefore, LS cannot be increased beyond the limiting rate and still attain toluene product purity (90 mole %)

In order to provide some room for the SSS system to handle changes in feed concentrations, the design value of LS was set at 90% of a maximum value. This resulted in a toluene product with 0.46 mole% xylene and 9.54 mole% benzene impurities.

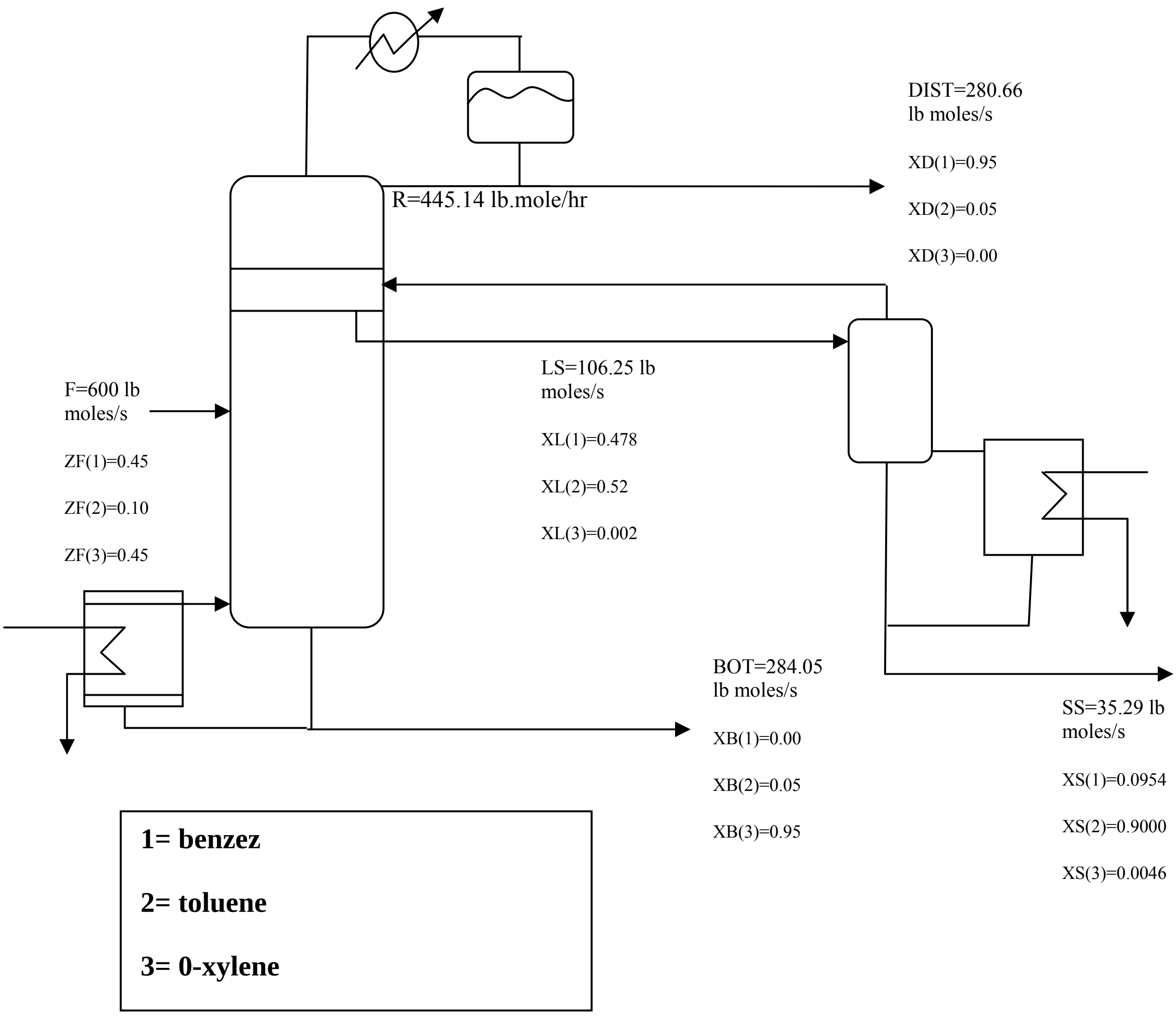


Figure 2.1 Basic design for SSS system [56]

2.4.2 Dimensions of the Control System

The SSS configuration has two compositions to be controlled and two manipulated variables as shown below:

$$\frac{X_D}{R} = \frac{4.09e^{-1.3s}}{(33s+1)(8.3s+1)} \quad \dots (2.1)$$

$$\frac{X_D}{L_s} = \frac{-4.17e^{-5s}}{45s+1} \quad \dots (2.2)$$

$$\frac{X_S}{R} = \frac{-0.49e^{-6s}}{(22s+1)^2} \quad \dots (2.3)$$

$$\frac{X_S}{L_S} = \frac{1.53e^{-3.8s}}{48s+1} \quad \dots (2.4)$$

A major question in the SSS control problem was the manipulation of the side raw rate LS. In theory, the LS rate could be held constant, and the other manipulated variable could be used to control the product purities. However, when the intermediate feed concentration was changed, it was found that LS manipulation was necessary to maintain toluene product purity and to minimize energy consumption. Parametric steady state studies showed that maintaining a constant temperature difference (ΔT) between trays above and below side raw tray by manipulating LS kept energy consumption near its minimum.

Chapter Three

Theory

3.1Process Simulation

Process simulation technology has evolved dramatically with the increasing application of object-oriented programming. Many packages are available which allows intuitive visualization of process data coupled with user-friendly graphical interface, which allows rapid synthesis of process flow sheets using click and drag operations. These packages are common in distillate process for both modeling and real-time process control.

This process described the MATLAB's dynamic simulation engine, simulink, in process dynamics and control using the graphical user interface *(GUI)* design tool, dubbed *GUIDE* by MATLAB [63].

3.1.1 Process Control Modules

Process control modules are set of programs written in the MATLAB environment. The modules based on fundamental process models of distillation column, and incorporate a realistic graphical user interface to emulate an industrial control environment they were developed to allow educator to strike a proper balance between theory and practice using a computer-based control laboratory.

Such laboratory can address a variety of issue in process control, modeling, identification, simulation, analysis and design. The software tools that have been chosen for the instructional laboratory are MATLAB *(matrix laboratory*) and SIMULINK (*dynamic system simulation software*). They represent an interactive program for scientific and engineering calculations.

The primary distinction of the PCM software package is the ability to incorporate complex dynamical process models with convenient graphical user interfaces (developed in the MATLAB graphics language). In away to achieve a realistic emulation of an industrial distributed control system operator's console. Furthermore, the entire packages based on commercial software plate form (MATLAB) which allows individual instructors customize or add to existing modules [64].

3.1.2 Simulink Response

Simulink Response Optimization can also improve the tuning of look up tables and aid in gain scheduling. Simulink has block libraries from which subsystems, sources (transfer function), and sink (scope). Sub system blocks are variable for responding linear, nonlinear, and discrece systems. Therefore, the responses of distillation column are obtained using Simulink. Simulink Response Optimization is a tool that helps to tune design parameters in Simulink models by optimizing time-based signals to meet user-defined constraints.

It optimizes scalar, vector, and matrix-type variables and constrains multiple signals at any level in the model. Simulink Response Optimization supports continuous, discrete, and multirate models and enables you to account for model uncertainty by conducting Monte Carlo simulations [65].

3.2 Obtaining Transfer Function

The analytical determination for the system's transfer function is difficult. The individual component values been know, or the internal configurations of the system may not accessible. In such cases, the frequency response of the system, from input to output is obtained experimentally and using the frequency response plot experimentally, by using a sinusoidal force or signal generator at the input of the system and measure the output steady state sinusoidal amplitude and phase angle. Repeating this process at a number of frequencies yield data for a frequency response plot, the transfer function can be presented as a block diagram, where the input on the left and the output on the right and the system transfer function inside the block [4, 66].

3.3 The Transfer Function Models

There are three processes before doing the simulator, first, the transfer functions between the input and output of the open loop process must be determined, and then decoupler transfer functions can be computed to minimize loop interaction. The next non-trivial step is to determine dynamic of the process, the decoupler and the cross-decoupling term, and the third is based on the use of an external computer simulator [67].

The transfer function model for the column is due to Alatiqi and Luyben [56] given below is: -

$$\begin{matrix} x_D \\ x_S \end{matrix} = \begin{bmatrix} \frac{4.09e^{-1.3s}}{(33s+1)(8.3s+1)} & \frac{-4.17e^{-5s}}{45s+1} \\ \frac{-0.49e^{-6s}}{(22s+1)^2} & \frac{1.53e^{-3.8s}}{48s+1} \end{bmatrix} \begin{matrix} R \\ L_S \end{matrix} \qquad \ldots (3.1)$$

Where:-

X_D: the distillate composition

X_S: the side stream composition

R: the reflux flow rate

L_S: the side stream flow rate

3.4 Disturbances

Disturbances can be classified and defined in several ways: [67]

a. **Step**: Step disturbances are functions that change instantaneously from one level to another and are thereafter constant. If the size of the step is equal to unity, the disturbance is called the unit ***step function*** $u_{n(t)}$ defined as:

$u_{n(t)} = 1$ for $t > 0$ … (3.2)

$u_{n(t)} = 0$ for $t \leq 0$

The response of a system to a step disturbance is called the step ***response*** or the ***transient response.***

b. **Pulse**: A pulse is a function of arbitrary shape (but usually rectangular or triangular) that begins and ends at the same level. A rectangular pulse is simply the sum of one positive step function made at time zero and one negative step function made D minutes later. D is the length of the pulse.

Rectangular pulse of height 1 and width $D = u_{n(t)} - u_{n(t-d)}$ …(3.3)

c. **Impulse**: The impulse is defined as the Dirac delta function, an infinitely high pulse whose width is zero and whose area is unity. This kind of disturbance is, of course, a pure mathematical fiction.

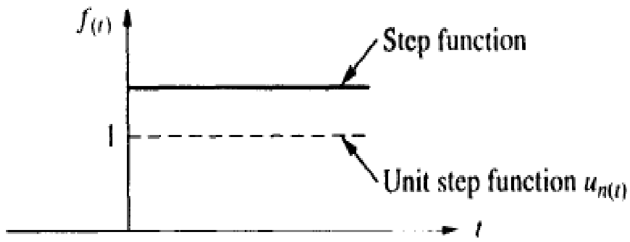
d. **Ramp**: Ramp inputs are functions that change linearly with time.

Ramp function = Ct … (3.4)

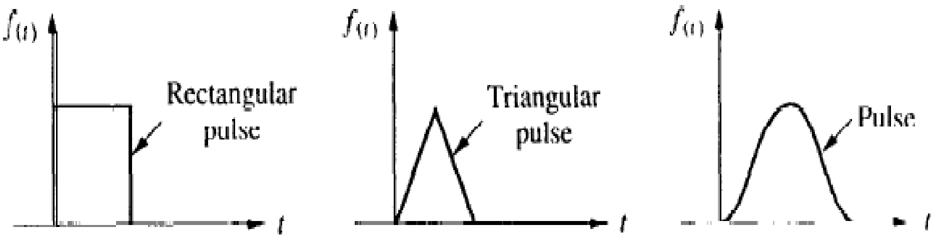
Where *C* is a constant. Chemical engineering examples include batch reactor temperature or pressure set point changes with time.

e. **Sinusoid**: Pure periodic sine and cosine inputs seldom occur in real chemical engineering systems. However, the response of systems to this kind of forcing function (called the frequency ***response*** of the system) is of great practical importance.

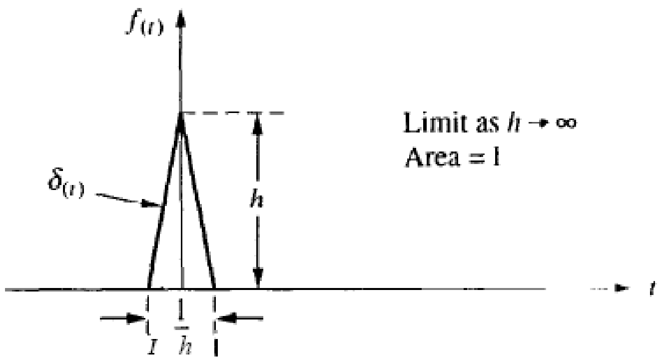
1. step



2. Pulses



3. Impulse (Dirac delta function $\delta_{(t)}$)



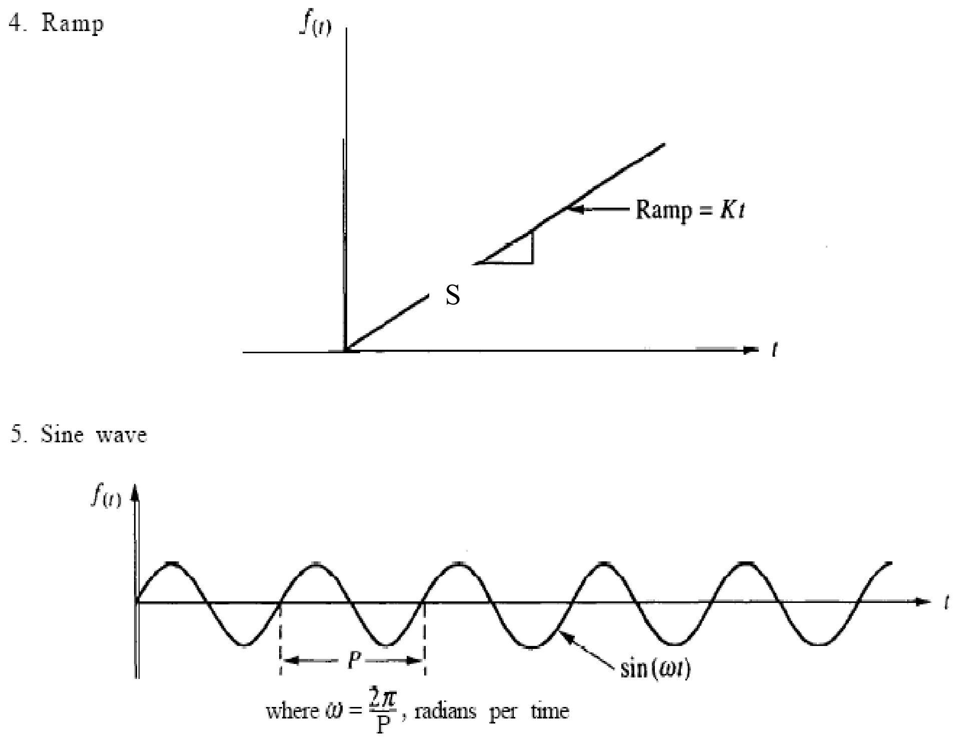


Figure 3.1 Disturbances shape [67]

3.4.1 Step Disturbance

The most direct way of obtaining an empirical linear dynamic model of a process is to find the parameters (dead time, time constant, and damping coefficient) that fit the experimentally obtained step response data. The process being identified is usually open loop, but experimental testing of closed loop systems is also possible.

u(t) is putted in a step disturbance and the output variable y(t) is recorded as a function of time, as illustrated in **Fig.3.1**. [67].

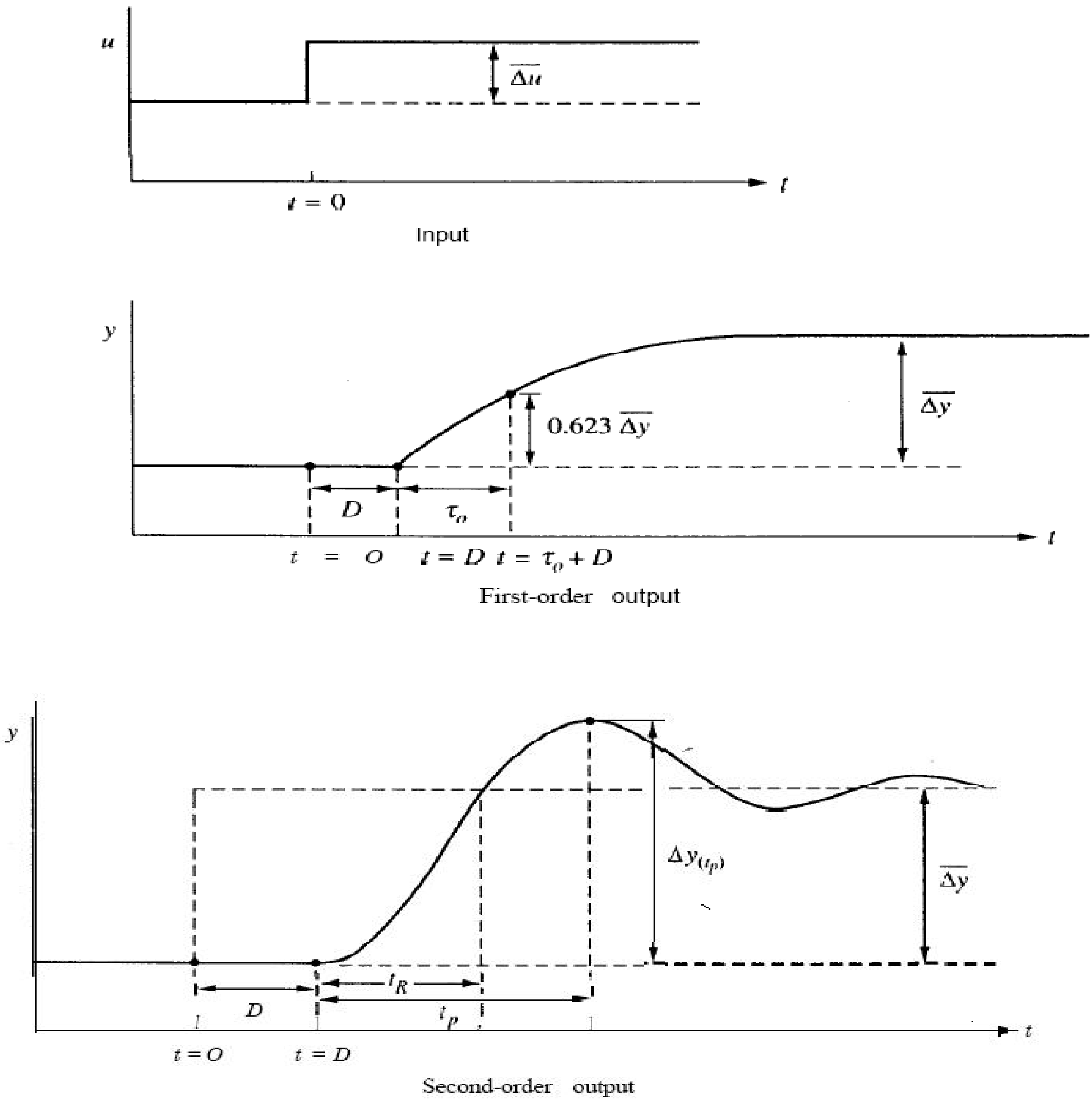


Figure 3.2 Step response [67]

$$G_{(s)} = K_p \frac{e^{-Ds}}{\tau_o s+1} \qquad \dots (3.5)$$

The steady-state gain K_p is easily obtained from the ratio of the final steady-state change in the output Δy over the size of the step input Δu. The dead time can be easily read from the y(t) curve. The time constant can be estimated from the time it takes the output y(t) to reach 62.3 percent of the final steady-state change. Closed loop processes are usually tuned to be somewhat under damped, so a second-order under damped model must be used.

$$G_{(s)} = K_p \frac{e^{-Ds}}{\tau^2 s^2 + 2\tau\zeta s + 1} \quad \ldots (3.6)$$

As shown in **Fig. (3.2)**, the steady-state gain and dead time are obtained in the same way as with a first-order model. The damping coefficient can be calculated from the "peak overshoot ratio," POR, using **Eq. (3.7**).

$$POR = CAe^{\frac{-\pi\delta}{\sqrt{1-\delta^2}}} \quad \ldots (3.7)$$

Where $POR = \frac{\Delta y(tp) - \Delta y}{\Delta y}$ ... (3.8)

$\Delta y(t_p)$ = change in y(t) at the peak overshoot.

t_p time to reach the peak overshoot (excluding the dead time) .

Then the time constant τ can be calculated from Eq. (3.9).

$$\frac{t_r}{\tau} = \frac{\pi - \phi}{\sin\phi} \quad \ldots (3.9)$$

$$\emptyset = \frac{\delta}{\sqrt{1-\delta^2}}$$

where t_R is the time it takes the output to reach the final steady-state value for the first time (**see Fig. 3.2**).

These estimation methods are simple and easy to use. They can provide a rough model that is adequate for many engineering purposes. For example, an approximate model can be used to get preliminary values for controller settings.

3.5 Frequency Response Analysis

The frequency response of a control system presents a qualitative picture of the transient response; the correlation between frequency and transient responses is indirect, except for the case of the second-order system. In analyzing a closed-loop system, we adjust the frequency response characteristics of the open-loop transfer function by using analysis criteria in order to obtain acceptable transient response characteristics for the system.

If we have indicated the relative stability by frequency response method we must calculate the gain margin and phase margin for the control system when the loop is open [69].

3.5.1 Frequency Response Plotting

In this work, frequency response plotting by Bode plots, which compute the magnitude and phase of the frequency response of linear models.

The magnitude is plotted in decibels (dB), and the phase in degrees. Bode plots are used to analyze system properties such as the gain margin, phase margin, and stability.

3.5.2 Phase Margin and Gain Margin

The phase margin is the amount of additional phase angle at the gain crossover frequency required to bring the system to instability. The gain crossover frequency is the frequency at which the magnitude of the open-loop transfer function is unity. The phase margin is 180º plus the phase angle of the open-loop transfer function at the gain crossover frequency. For a minimum-phase system to be stable, the phase margin must be positive, while the gain margin is the reciprocal of the magnitude at the frequency at which the phase angle is -180º. The phase crossover frequency is the frequency at which the phase angle of the open-loop transfer function equals -180º gives the gain margin.

The gain margin expressed in decibels is positive if it is greater than unity and negative if it is smaller than unity. Thus a positive gain margin (in decibels) means that the system is stable, and the negative gain margin (in decibels) means that the system is unstable [69].

3.5.3 Frequency response data

Frequency response data are complex numbers, by varying the input frequency over a range, to obtain a group of frequency response data; these data can be manipulated and displayed in a variety of ways, such as bode diagram that shown below:

Bode plots: Two plots the magnitude of *G (jw)* is in decibels; phase is in degrees. The horizontal axes are logarithm of angular frequency. Bode plots are most frequently used in control systems analysis and design, **fig. 3.3** [69].

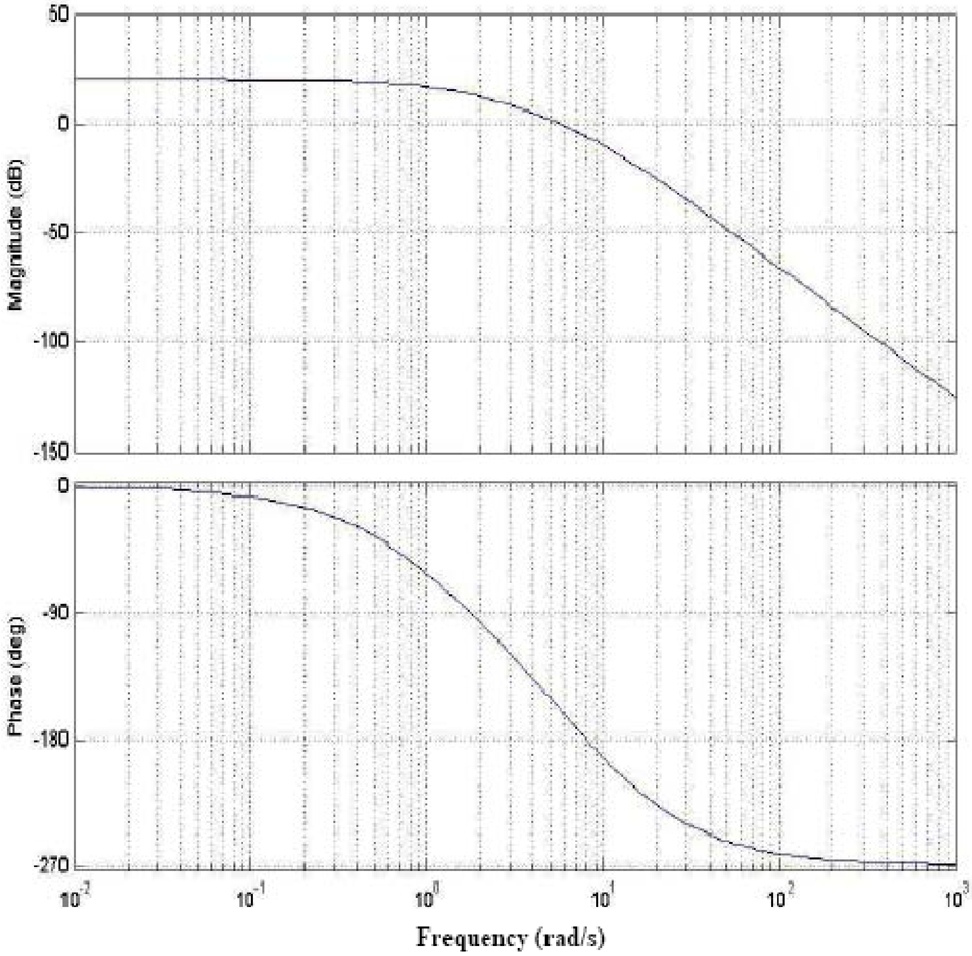


Figure 3.3 Bode plots [69].

3.6 Control Strategies [71]

3.6.1. Types of Control

Process controls are instruments used to control a parameter, such as temperature, level, and pressure. PID controllers are a type of continuous controller because they continually adjust the output vs. an on/off controller, when looking at feed forward or feed backward conditions.

3.6.1.1 Proportional (P) Control

P-control linearly correlates the controller output (actuating signal) to the error (difference between measured signal and set point). This P-control behavior is mathematically illustrated in Equation (3.10).

$$c(t) = K_c e(t) + b \qquad \dots (3.10)$$

c (t) = controller output

K_c = controller gain

e(t) = error

b = bias

In this equation, the bias and controller gain are constants specific to each controller. The bias is simply the controller output when the error is zero.[71]

3.6.1.2 Integral (I) Control

I-control correlates the controller output to the integral of the error. The integral of the error is taken with respect to time. It is the total error associated over a specified amount of time. This I-control behavior is mathematically illustrated in **Equation (3.11)**.

$$c(t) = \frac{1}{T_i} \int e(t)dt + c(t_0) \qquad \dots (3.11)$$

c(t) = controller output

τ_i = integral time

$e(t)$ = error

$c(t_0)$ = controller output before integration

In this equation, the integral time is the amount of time that it takes for the controller to change its output by a value equal to the error.[71]

3.6.2.3 Derivative (D) Control

D-control correlates the controller output to the derivative of the error. The derivative of the error is taken with respect to time. It is the change in error associated with change in time. This D-control behavior is mathematically illustrated in **Equation (3.12)**.

$$c(t) = T_d \frac{de}{dt} \quad \dots (3.12)$$

$c(t)$ = controller output

T_d = derivative time constant

de = change in error

dt = change in time

Mathematically, derivative control is the opposite of integral control. Although I-only controls exist, D-only controls do not exist. D-controls measure only the change in error. [71]

3.6.2.4 Proportional-Integral (PI) Control

PI-control correlates the controller output to the error and the integral of the error. This PI-control behavior is mathematically illustrated in **Equation (3.13)**.

$$c(t) = K_c e(t) + \frac{1}{Ti} \int e(t)dt + C$$

… (3.13)

c(t) = controller output

K_c = controller gain

T_i = integral time

e(t) = error

C = initial value of controller

In this equation, the integral time is the time required for the I-only portion of the controller to match the control provided by the P-only part of the controller. [71]

3.6.2.5 Proportional-Derivative (PD) Control

As mentioned, PD-control correlates the controller output to the error and the derivative of the error. This PD-control behavior is mathematically illustrated in **Equation (3.14)**.

$$c(t) = K_c e(t) + T_d \frac{de}{dt} + C$$

… (3.14)

c(t) = controller output

K_c = proportional gain

e = error

C = initial value of controller

The equation indicates that the PD-controller operates like a simplified PID-controller with a zero integral term. Alternatively, the PD-controller can also be seen as a combination of the P-only and D-only control equations. [71]

3.6.2.6 Proportional-Integral-Derivative (PID) Control

PID-control correlates the controller output to the error, integral of the error, and derivative of the error. This PID-control behavior is mathematically illustrated in **Equation (3.15)**.

$$c(t) = K_c e(t) + \frac{1}{T_i} \int e(t)dt + T_d \frac{de}{dt} + C \qquad \dots (3.15)$$

c(t) = controller output

K_c = controller gain

e(t) = error

T_i = integral time

T_d = derivative time constant

C = intitial value of controller

As shown in the above equation, PID control is the combination of all three types of control. The P-, I-, and D- controllers can be combined in other ways. These alternative combinations are simplifications of the PID-control. [71]

3.6.2 Summary Tables

A summary of the advantages and disadvantages of the three controls is shown below in **Table 3.1**.

Table 3.1 Advantages and disadvantages of controls [71]

	Proportional (P)	Integral (I)	Derivative (D)
Advantages	+ Fast response time + Minimizes fluctuation	+ Contains small offset + Returns system to steady state	+ Keeps system at consistent setting + Controls processes with rapidly changing outputs
Disadvantages	- Contains large offset - Does not bring system to desired set point	- Slow response time	- Slow response time - Requires combined use with another controller

A guide for the typical uses of the various controllers is shown below in **Table 3.2**

Table 3.2 Typical uses of P, I, D, PI, and PID controllers [71]

Controller	Estimates	When to use	Examples
P	Present	Systems with slow response, systems tolerant to offset	Float valves, thermostats, humidistat
I	Back	Not often used alone, as is too slow	Used for very noisy systems
D	Forward	Not used alone because it is too sensitive to noise and does not have set point	None
PI	Present & back	Often used	Thermostats, flow control, pressure control
PID	All time	Often used, most robust, but can be noise sensitive	Cases where the system has intertia that could get out of hand: i.e. temperature and concentration measurements on a reactor to avoid runaway.

3.7 CONTROLLER TUNING

3.7.1 The Ziegler-Nichols (ZN)

The Ziegler-Nichols (ZN) controller settings are pseudo-standards in the control field. They are easy to find and to use and give reasonable performance on some loops. The ZN settings are benchmarks against which the performance of other controller settings is compared in many studies. They are often used as first guesses, but they tend to be too under damped for most process control applications. Some on-line tuning can improve control significantly. But the ZN settings are useful as a place to start.

The ZN method consists of first finding the ultimate gain K_u the value of gain at which the loop is at the limit of stability with a proportional-only feedback controller.

The period of the resulting oscillation is called the *ultimate period,* P, (minutes per cycle). The ZN settings are then calculated from K_u and P_u by the formulas given in table 3.3 for the three types of controllers. Notice that a lower gain is used when integration is included in the controller (PI) and that the addition of derivative permits a higher gain and faster reset. [67]

Table 3.3 Ziegler-Nichols (ZN) settings [67]

	P	PI	PID
K_c	$K_u/2$	$K_u/2.2$	$K_u/1.7$
τ_i	-	$P_u/1.2$	$P_u/2$
τ_d	-	-	$P_u/8$

3.7.2 Tyreus-Luyben

The Tyreus-Luyben method procedure is quite similar to the Ziegler-Nichols method but gives more conservative settings (higher closed loop damping coefficient) and is more suitable for chemical process control applications.

The method uses the ultimate gain K_u and the ultimate frequency w_u. The formulas for PI and PID controllers are given in **Table 3.4**[67]

Table 3.4 Tyreus-Luyben [67]

	P	PI	PID
K_c	-	$K_u/3.2$	$K_u/2.2$
τ_i	-	$2.2P_u$	$2.2P_u$
τ_d	-	-	$P_u/6.3$

3.7.3 Cohen-Coon

Cohen-Coon method requires an open-loop first-order-plus-time-delay transfer function model of the process. This can be obtained from a process reaction curve. From the identified effective gain, time constant and dead time (K_c, τ_i, τ_d), and one can computed controller using the rules which are summarized in **table 3.5**[63]

Table 3.5 Cohen-coon settings [63]

	P	PI	PID
Kc	$K_c = (\frac{1}{K})(\frac{\tau}{t_d})(1+\frac{t_d}{3\tau})$	$K_c = (\frac{1}{K})(\frac{\tau}{t_d})(\frac{9}{10}+\frac{t_d}{12\tau})$	$K_c = (\frac{1}{K})(\frac{\tau}{t_d})(\frac{4}{3}+\frac{t_d}{4\tau})$
τ_i	-	$\tau_I = t_d(\frac{30+3t_d/\tau}{9+20t_d/\tau})$	$\tau_I = \tau_d(\frac{30+3\tau_d/\tau}{9+20\tau_d/\tau})$
τ_d	-	-	$\tau_D = t_d(\frac{4}{11+2t_d/\tau})$

3.8 Relative Gain Array (RGA)

The RGA is a matrix of numbers. The *i* jth element in the array is called *ßij*. It is the ratio of the steady-state gain between the ith controlled variable and the jth manipulated variable when all other manipulated variables are constant, divided by the steady-state gain between the same two variables when all other controlled variables are constant.

The RGA has the advantage of being easy to calculate and requires only steady-state gain information. [67]

$$\beta_{ij} = \frac{[Y_i/m_j]_{m_k}}{[Y_i/m_j]_{Y_k}} \quad \ldots (3.16)$$

For example, suppose we have a 2 **X** 2 system with the steady-state gains K_{Pij}.

$$Y_1 = K_{p11}m_1 + K_{p12}m_2 \quad \ldots (3.17)$$

$$Y_2 = K_{p21}m_1 + K_{p22}m_2 \quad \ldots (3.18)$$

For this system, the gain between Y_1 and m_1 when m_2 is constant is

$$[Y_1/m_1]_{m2} = K_{p11} \quad \ldots (3.19)$$

The gain between Y_1 and m_1 when Y_2 is constant (Y_2 = 0) is found from solving the equations

$$Y_1 = K_{p11}m_1 + K_{p12}m_2 \quad \ldots (3.20)$$

$$0 = K_{p21}m_1 + K_{p22}m_{2k}$$

$$Y_1 = K_{p11}m_1 + K_{p12}[-K_{p21}m_1/K_{p22}]$$

$$Y_1 = \left[\frac{K_{p11}K_{p22} - K_{p12}K_{p21}}{K_{p22}}\right] m_1 \quad \ldots (3.21)$$

$$[Y_1/m_1]_{\overline{Y}_2} = \left[\frac{K_{p11}K_{p22} - K_{p12}K_{p21}}{K_{p22}}\right]$$

Therefore, the β_{11} term in the RGA is:

$$\beta_{11} = \frac{1}{1 - \frac{K_{p_{12}} K_{p_{21}}}{K_{p_{11}} K_{p_{22}}}} \quad \ldots (3.22)$$

3.8.1 Uses and limitations

The elements in the RGA can be numbers that vary from very large negative values to very large positive values. If the RGA is close to 1, there should be little effect on the control loop by closing the other loops in the multivariable system. Therefore, there should be less interaction, so the proponents of the RGA claim that variables should be paired so that they have RGA elements near 1. Numbers around 0.5 indicate interaction. Numbers that are very large indicate interaction. Numbers that are negative indicate that the sign of the controller may have to be different when other loops are on automatic.

The problem with pairings to avoid interaction is that interaction is not necessarily a bad thing. Therefore, the use of the RGA in deciding how to pair variables is not an effective tool for process control applications. Likewise, the use of the RGA in deciding what control structure (choice of manipulated and controlled variables) is best is not effective. What is important is the ability of the control system to keep the process at set point in the face of load disturbances. Thus, load rejection is the most important criterion for deciding what variables to pair and what controller structure is best.

The RGA is useful for avoiding poor pairings. If the diagonal element in the RGA is negative, very large values of the RGA indicate that the system can be quite sensitive to changes in the parameter values [67].

3.9 Decoupling

Some of the earliest work in multivariable control involved the use of decouplers to remove the interaction between the loops. **Fig. 3.11** gives the basic structure of the system. The decoupling matrix $D_{(s)}$ is chosen such that each loop does not affect the other. **Fig. 3.12** shows the details of a 2 **x** 2 system. The decoupling element D_{ij} can be selected in a number of ways. One of the most straightforward is to set $D_{11} = D_{22} = 1$ and design the D_{12} and D_{21} elements so that they cancel (in a feed forward way) the effect of each manipulated variable in the other loop. For example, suppose $Y_{(1)}$ is not at its setpoint but $Y_{(2)}$ is. The G_{c1} controller changes m_1 to drive Y_1 back to Y_1^{set}. But the change in m_l disturbs Y_2 through the G_{M21} transfer function.

If, however, the D_{21} decoupler element is set equal to $(-G_{m21}/G_{m22})$, there is a change in m_2 that comes through the G_{m22} transfer function and cancels out the effect of the change in m_1 on Y_2 [67].

$$D_{21} = \frac{-G_{M21}}{G_{M22}} \qquad \text{... (3.23)}$$

Using the same arguments for the other loop, the D_{12} decoupler could be set equal to

$$D_{12} = \frac{-G_{M12}}{G_{M11}} \qquad \text{... (3.24)}$$

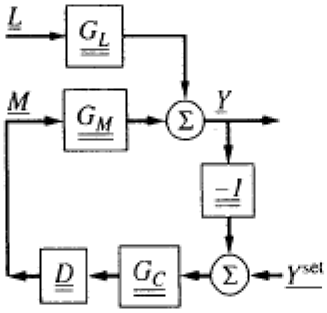


Figure 3.4 [67]
Decoupling.

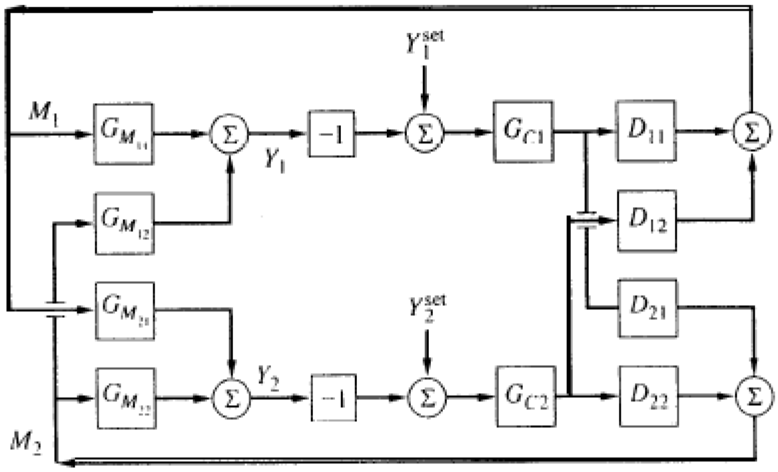


Figure 3.5 Block diagram for 2x2 system with decoupler [67]

This "simplified decoupling" splits the two loops so that they can be independently tuned. Note, however, that the closed loop characteristic equations for the two loops are not $1 + G_{M11} G_{C1} = 0$ and $1 + G_{M22}G_{C2} = 0$. The presence of the decouplers changes the closed loop characteristic equations to:

$$1 + G_{C1}\frac{G_{M11}G_{M22}-G_{M12}G_{M21}}{G_{M22}} = 0 \qquad \ldots (3.25)$$

$$1 + G_{C2}\frac{G_{M11}G_{M22}-G_{M12}G_{M21}}{G_{M11}} = 0 \qquad \ldots (3.26)$$

Other choices of decouplers are also possible. However, since decoupling may degrade the load rejection capability of the system, the use of decouplers is not recommended except in those cases where set point changes are the major disturbances.

3.10 Fuzzy Logic Control (FLC)

Fuzzy logic is a super set of traditional logic, according to Zadeh, [72] who invented this concept in 1965. The logic of approximate reasoning and it is unlike other branches of artificial intelligence (AI) that use conventional logic. Since then, the theory of mathematics has gained more recognition from many researches in a wide range of scientific fields.

Fuzzy mathematics is attractive not only because it is based on the very intuitive idea of Fuzzy sets, but because it is capable of generating many structures that provide today's scientists and engineers with new insights into interesting, significant and often-debated problems in both science and engineering. [73, 74]

The theory of Fuzzy sets has one of its aims, the development of a methodology for the formulation and solution of problems that are too complex or too ill-defined to be analyzed by conventional techniques.

Hence the theory of Fuzzy sets is likely to be recognized as a natural development in the evaluation of scientific thinking [75].

3.10.1 Application of Fuzzy Logic Control System

There are five types of systems where fuzziness is necessary and these systems are: [76]

a. Complex systems and those are difficult to model.

b. Systems controlled by human expertise.

c. Systems with complex and continuous change in inputs and outputs.

d. Systems that use human observation as inputs or as the basis for rules.

e. Systems which are naturally vague, such as those in behavioral and social sciences.

3.10.2 Advantages of Fuzzy Logic Control System:

a. It relates output to input without having understood all the variables, permitting the design of a system that may be more accurate and stable than are with a conventional control system.
b. Rapid prototyping is possible because a system designer does not have to know everything about the system before starting.
c. It is cheaper to make than conventional systems because it is easier to design.
d. It has increased robustness.
e. It simplifies knowledge acquisition and representation [77,78].

3.10.3 Fuzzy Set Basic Operations

The space which Fuzzy set are working in is called the (universe of discourse). Then a Fuzzy subset (A) of a universe of discourse (ν) is characterized by a membership function [$\mu_{A(\rho)}$] which is assigned to each element [$\rho \varepsilon \nu$]. A membership [$\mu_{A(\rho)}$] in the interval (0 to 1) represents the grade of membership in a Fuzzy subset (A). The three basic operations used are defined as follows: [78]

A. The *union* of the Fuzzy subsets (A) and (B) of the universe of discourse (ν) is denoted by:

[A $\cup$ B] with a membership function defined by:

$$\mu_{A \cup B}(e) = \max [\mu_A(e); \mu_B(e)] \qquad \text{... (3.27)}$$

This corresponds to the connective (OR).

B. The *Interaction* of the Fuzzy subsets (A) and (B) of the universe of discourse (ν) is denoted by:

[A $\cap$ B] with a membership function defined by:

$$\mu_{A \cap B}(e) = \min [\mu_A(e); \mu_B(e)] \qquad \text{... (3.28)}$$

This corresponds to the connective (AND).

C. The complement of a Fuzzy subset (A) of the universe of discourse (ν) is denoted by:

[A] with a membership function denoted by:

$$\mu_A(e) = 1 - \mu_A(e) \qquad \text{... (3.29)}$$

This corresponds to the connective (NOT).

3.10.4 Design of Fuzzy Logic Controller

The purpose of any plant controller is to relate the state variables to action variables. Now the controller of a physical system need not itself be physical but may be purely logic. Furthermore, where known relationships are vague and qualitative, a Fuzzy logic controller may be constructed to implement the known heuristic. Thus in such a controller the variables are equated to non-Fuzzy universe given the possible range of measurement or action magnitudes. These variables, however, take on linguistic values which are expressed as Fuzzy subset of the universe. The complete procedure of the Fuzzy controller design can be described as follows: [79]

1. Choose a suitable scaled universe of discourse (ν) of $-L \le (E_i, CE_i) \le L$, Where: L and –L represent the positive and negative ends respectively of this universe which is quantized into equally spaced levels in

between those two ends. E_i and CE_i represent the error and its rate of change for the same instant (i).

2. Define the non-Fuzzy set intervals (the quantized levels scaled values) for E_i, CE_i and control action (U). Each level has a value (I) lying between ($-XG \leq I \leq XG$) where: XG and –XG represent the controller gain and they are regarded as the values of the universe of discourse limits (L and –L) respectively.
3. The theory of Fuzzy sets deals with a subset (A) of the universe of discourse (ν), where the transition between the full membership ($\mu=1$) and on membership ($\mu=0$), is gradual rather than abrupt.[41]

   The Fuzzy-sets definitions in control for E, CE and U are used to have these forms:

   PB=positive Big NB=Negative Big Z=Zero

   PS=Positive Small NS=Negative Small

   The grades of membership, based on normal distribution, for these Fuzzy sets are declared as below:

   -Interval Center +Interval

   PB=(0,0,0,0,0,0,0,0,0,0,0,0,0,0,0.1,0.2,0.3,0.4,0.5,0.8,1)

   PS=(0,0,0,0,0,0,0,0,0,0,0.1,0.3,0.6,0.8,1,0.8,0.6,0.3,0.1,0,0)

   Z= (0,0,0,0,0,0,0.1,0.3,0.6,0.8,1,0.8,0.6,0.3,0.1,0,0,0,0,0,0)

   NS=(0,0,0.1,0.3,0.6,0.8,1,0.8,0.6,0.3,0.1,0,0,0,0,0,0,0,0,0,0)

   NB= (1,0.8,0.6,0.3,0.1,0,0,0,0,0,0,0,0,0,0,0,0,0,0,0,0)

The above three points of design procedures will form a look-up table of N rows and M columns, where N represents the number of intervals and M represents the number of Fuzzy sets definitions for variable (E,CE and U).

4. The Fuzzy decision rules are developed linguistically to do a particular control task and are implemented as set of Fuzzy conditional statements of the form:

IF E IS PB AND CE IS NB THEN ZERO ACTION

This form can be translated with the help of Fuzzy sets definition into a new statement.

IF PEB IS NCB AND ZU

Table 3.6 25-Rule Fuzzy Logic Controller [79]

	NCB	*CS*	*ZC*	*PCS*	*PCB*
NEB	*PUB*	*PUB*	*PUB*	*PUS*	*ZU*
NES	*PUB*	*PUS*	*PUS*	*ZU*	*NUS*
ZE	*PUB*	*PUS*	*ZU*	*NUS*	*NUB*
PES	*PUS*	*ZU*	*NUS*	*NUS*	*NUB*
PEB	*ZU*	*NUS*	*NUB*	*NUB*	*NUB*

The derivation of the Fuzzy rules can be obtained directly from the phase-plane of error and its rate of change. **Table 3.6** shows the Fuzzy rules conclusions. The five Fuzzy sets definition generates (25) rules Fuzzy controller. To read these one can obtain the following translation of the first three rules.

IF PEB AND PCB THEN NUB

IF PEB AND PCS THEN NUB

IF PEB AND ZC THEN NUB...and so on

It is worthy to know that in any control system:

E_i= (Set values) – (Measured values)$_i$ ... (3.30)

CE_i= (Instant error) – (pervious error) ... (3.31)

But in certain Fuzzy applications

E_i= (Measured values)$_i$ - (Set values) ... (3.32)

To clear this difference, consider the initial condition state for a system subjected to a unit step change in input.

For conventional controller,

E (0) =+1 , CE (0) =0 ... (3.33)

And according to **Table 3.6** the Fuzzy rule will be:

IF BEB AND ZC THEN NUB

The action will be negative and the output will follow it. To overcome this problem we must use the **Equation (3.30)**.

E (0) =-1 , CE (0) =0 ... (3.34)

And the Fuzzy rule will be:

IF NEB AND ZC THEN PUB

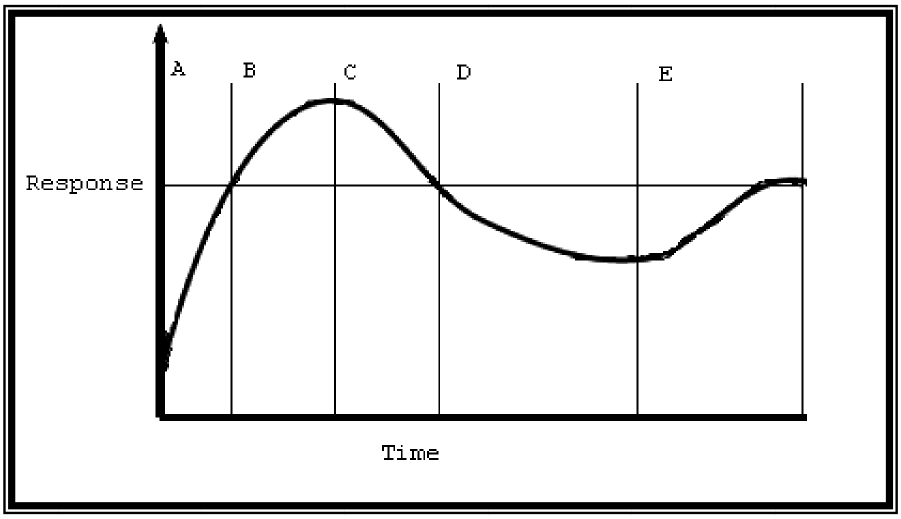


Figure 3.6 Transient Response of Closed Loop [79]

So the action will be positive and the output will follow it. This Fuzzy definition E and CE will be considered in this work.

According to **Equations (3.33), (3.34)** and **Figure 3.13** the signs of error and the change of error will be as follows in **Table 3.4**:

Table 3.7 Signs Distribution in Fuzzy logic controller [79]

Section	*Signs of error*	*Signs of change of error*
A	-	+
B	+	+
C	+	-
D	-	-
E	-	+

5. Both Ei and CEi are multipled by the scale factor of the universe of discourse to ensure mapping their values into suitable intervals that belong to each one, also this scale factor helps to simplify handling of the numerical values of all variables.

6. Control algorithm: the following steps show the algorithm design of a Fuzzy logic controller for SISO system which is shown in Figure (3.14).

a. For the error (E_i), the rate of change of error (CE_i) defines the Fuzzy subsets with their discrete membership functions.

b. Find the degree of full filament (DOF) by implementation of the part of all Fuzzy rules BY ANDing the membership of both E_i and CE_i.

c. Calculate the control vector (UA_j) for each rule by ANDing DOF_j with the control action subset elements (An) as shown below.

$$UA_j=\min(DOF_j, A_n) \quad \text{... (3.35)}$$

Where:

j=1, 2, 3....R, N=1, 2, 3......N

R is a number of rules and N is a number of intervals.

d. Compute the net control action vector (Ua_{net}) by Oring the vectors as UA_j as follows:

$$UA_{net} = \max [\min (DOF_j, A_n)] \qquad \text{... (3.36)}$$

7. Calculate the scalar control action (U_s), using the center of gravity method on which the selected deterministic output has a vector value that divides the area under a Fuzzy set into two equal halves.[46,47]

$$Us \frac{\sum_{n=1}^{N} I_n * (weight)_n}{\sum_{n=1}^{N} (weight)_n} \qquad \text{... (3.37)}$$

Where:

(Weight) represents the elements (membership) of the net control action vector.

(I) represents the value on the interval n.

8. An integral procedure (an algebraic sum) is required to obtain the effective control action scalar for each instant (i).

$$Us_{i+1}=U_{si}+Us_{i-1} \qquad \text{... (3.38)}$$

9. A scalar factor is used to remove the first scalar factor in order to put the values into real one.

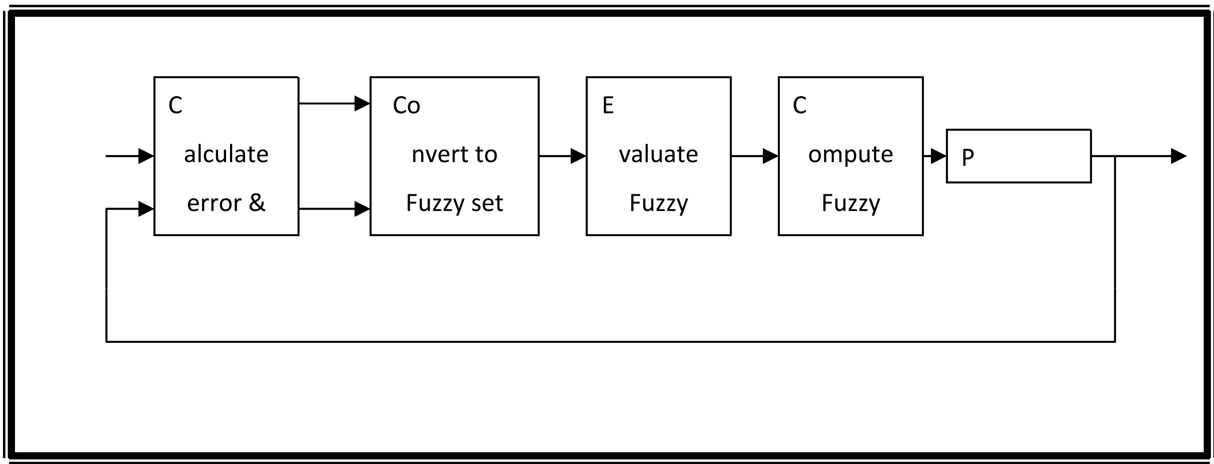


Figure 3.7 Block diagram of a control system using Fuzzy Logic Control [79]

3.10.5 Fuzzy Logic Control Procedure for MIMO System

Figure 3.15 describes a 2 × 2 Fuzzy controlled process. The Fuzzy control procedure for MIMO system is similar to the one for SISO process. All Fuzzy control functions are defined and calculations are made except that the fuzzy rules will be divided for each controlled variable taking into account the other controlled variables with (ANY membership) which gives a membership ($\mu = 1$) whenever it appears. To clarify the idea, the following Fuzzy rules are examined:

IF E1 IS PEB AND CE1 IS PCS AND E2 IS ANY AND CE2 IS ANY THEN NUB

The same shape of rules will be fulfilled for other controlled variable as shown below:

IF E1 IS ANY AND CE1 IS ANY AND E2 IS PEB AND CE2 IS PCS THEN NUB

And so on for all rules. From the definition of AND (min), (ANY) membership will have no effect on the control procedure.

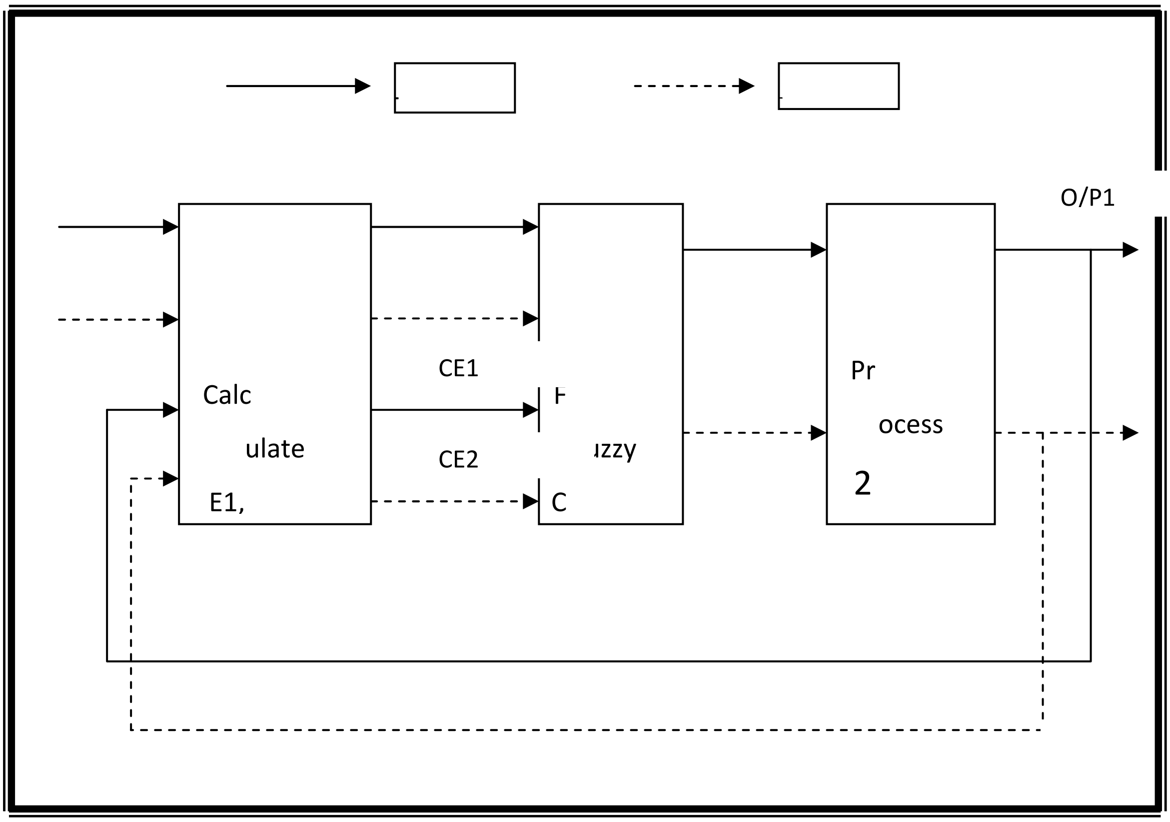


Figure 3.8 block diagram of fuzzy logic controller for 2×2 process [79]

3.10.6 Fuzzy Control Tuning

To modify the Fuzzy controlled response, three parameters are to be taken into account: [79]

a. *Gain tuning*: This is achieved by varying the gain and fixing other parameters.

b. *Interval tuning*: This can be done by varying the quantized level (interval) and fixing other parameters.

c. *Fine tuning*: This can be achieved by using more than one digit.

3.10.7 Controller Selection

To choose the suitable controller, the following points must be taken into account:

1. The controller ability to give a reasonable response, which depends on
    a. Number of rules.
    b. Number of intervals.
    c. Interval values.
    d. Fuzzy sets definition.
2. For real-time applications, the computer execution time required for performing the Fuzzy algorithm must be within the sampling period so as to give the appropriate control action.

Chapter Four

Results and Discussions

4.1 Introduction

This chapter illustrates the result of system using Simulink and $MATLAB_{7.6}$ programs to show the composition response.

Take a multi component distillation column with side stream as a system. The influences of manipulated variables (R and L_S) on controlled variables (X_D and X_S) were studied.

The matrix of transfer function was obtained by Alatiqi and Luyben [56] as shown below:

$$\frac{X_D}{R} = \frac{4.09e^{-1.3s}}{(33s+1)(8.3+1)} \quad \ldots (2.1)$$

$$\frac{X_D}{L_S} = \frac{-4.17e^{-5s}}{45s+1} \quad \ldots (2.2)$$

$$\frac{X_S}{R} = \frac{-0.49e^{-6s}}{(22s+1)^2} \quad \ldots (2.3)$$

$$\frac{X_S}{L_S} = \frac{1.53e^{-3.8s}}{48s+1} \quad \ldots (2.4)$$

As a matrix:

$$\begin{matrix} X_D \\ X_S \end{matrix} = \begin{bmatrix} \frac{4.09e^{-1.3s}}{(33s+1)(8.3+1)} & \frac{-4.17e^{-5s}}{45s+1} \\ \frac{-0.49e^{-6s}}{(22s+1)^2} & \frac{1.53e^{-3.8s}}{48s+1} \end{bmatrix} \begin{matrix} R \\ L_S \end{matrix} \quad \ldots (3.1)$$

$R_{(max)}(R_{(total)})$ =100% CO = 725.8lb.mol/hr	$Ls_{(max)}$ ($L_{S(total)}$)= 100%CO=201.69 lb.mol/hr
R_{min}= 0% co = 400 lb.mole/hr	Ls_{min}= 0% CO =70lb.mole/hr
$R_{st.st}$ = 61% CO = 445.14 lb.mole/hr	$Ls_{st.st}$=66% CO = 106.25lb.mole/hr

Where CO : controller output

4.2 open loop process

The results of the transient response based on open loop system are shown in **fig. 4.1**.

10% step change of manipulating variables which is reflux flow rate (R) and side stream flow rate (L_S) on controlled variable which is distillate composition (X_D) and side stream composition (X_S) were applied. The results are expressed as perturbation values.

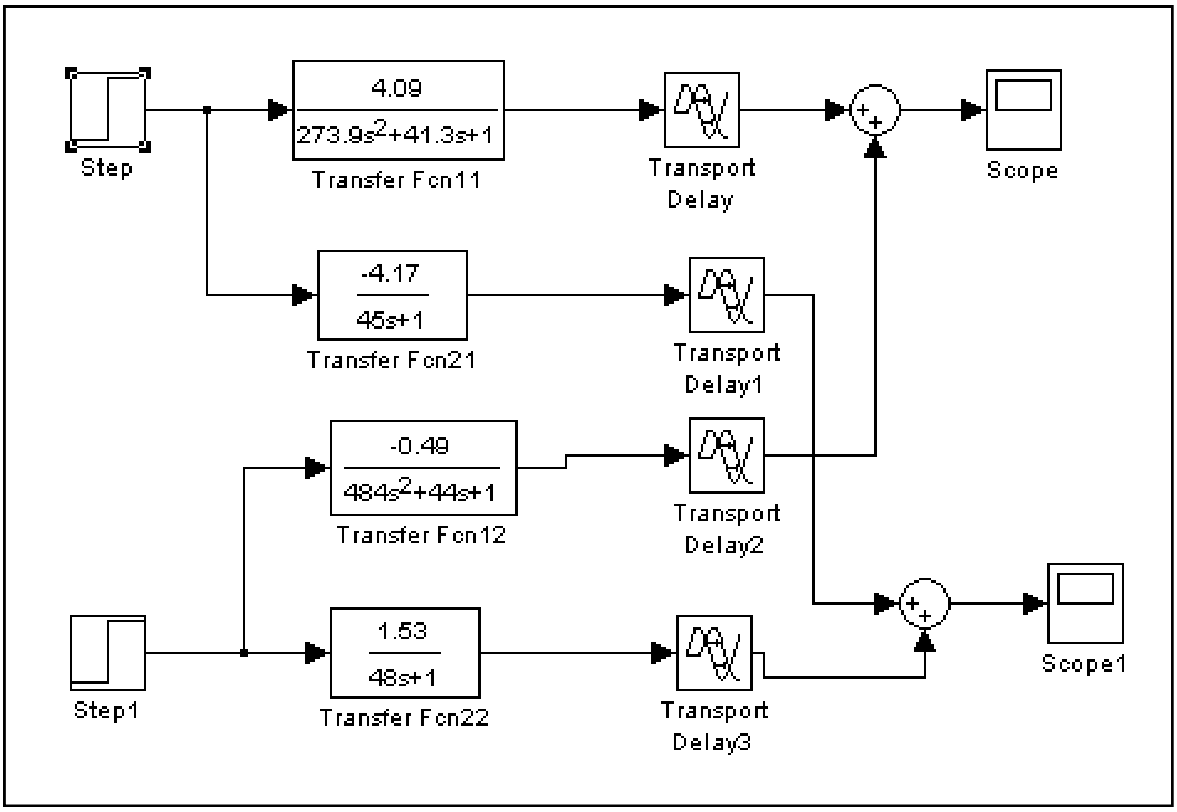


Figure 4.1.a Block diagram of open loop process

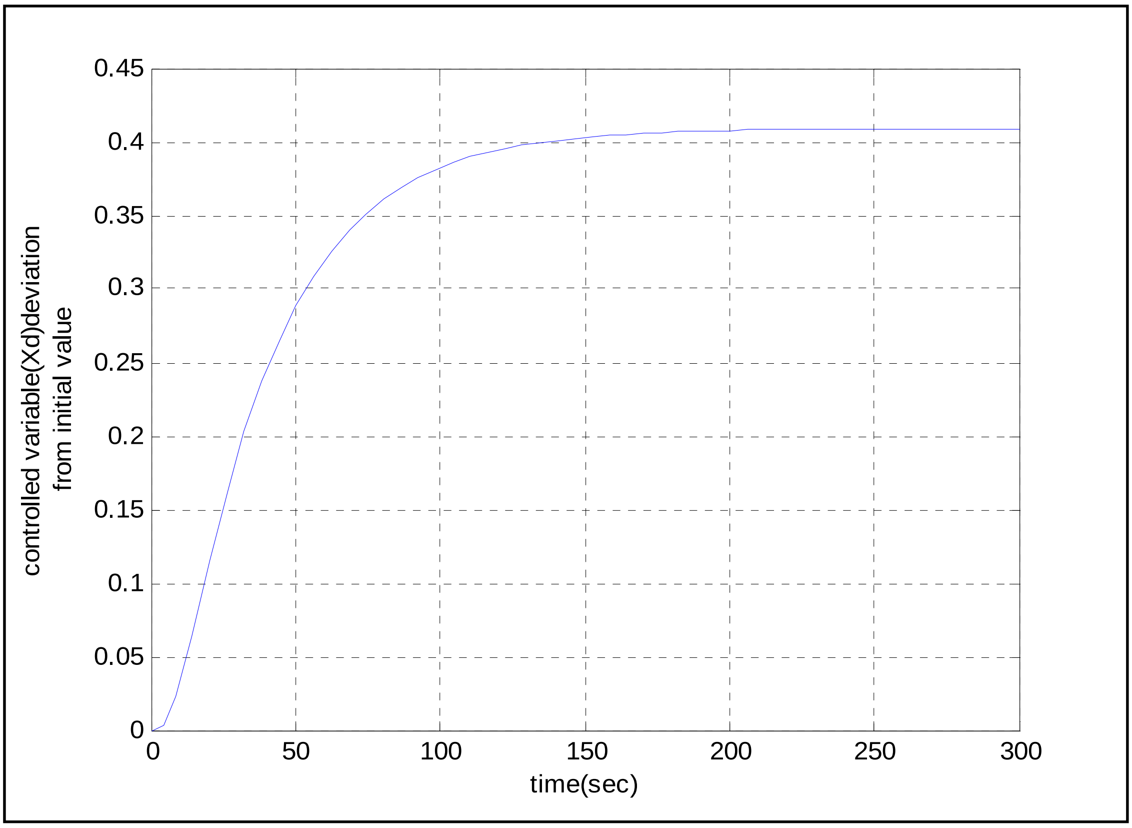


Figure 4.1.b.1 effect of R on X_D for 10% step change for open loop process

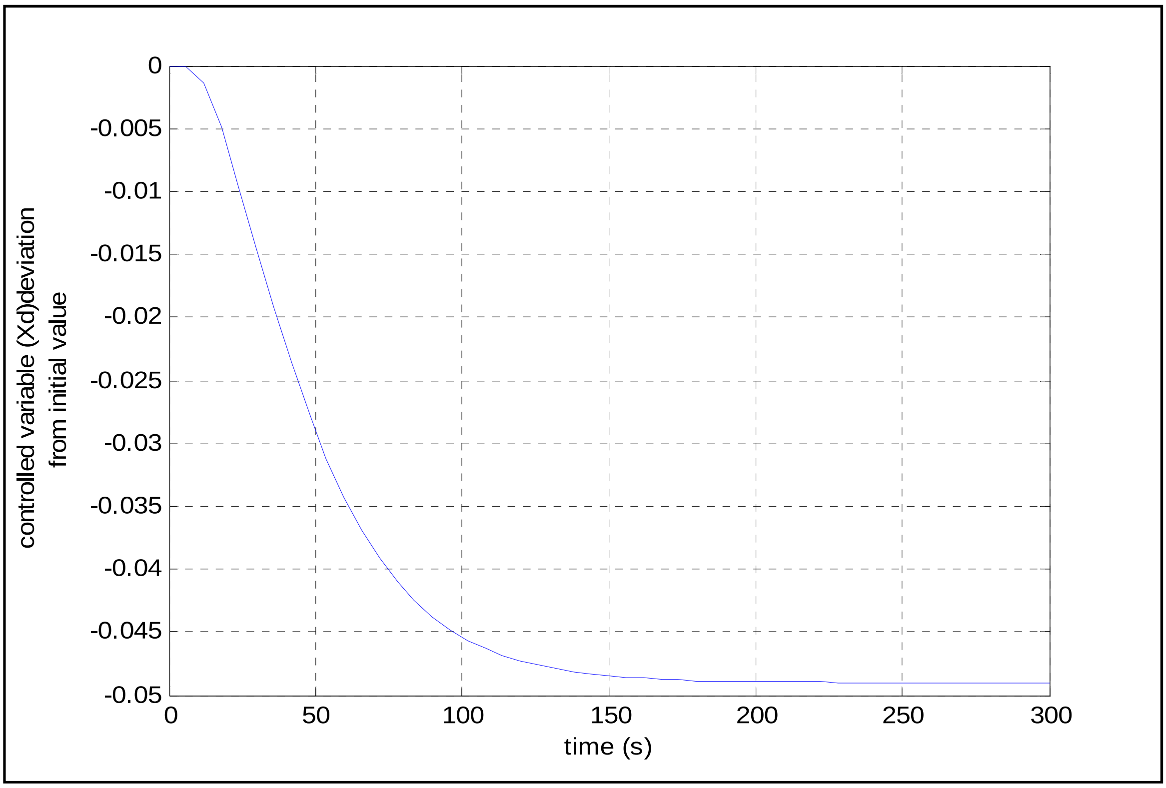


Figure 4.1.b.2 effect of L_S on X_D for 10% step change for open loop process

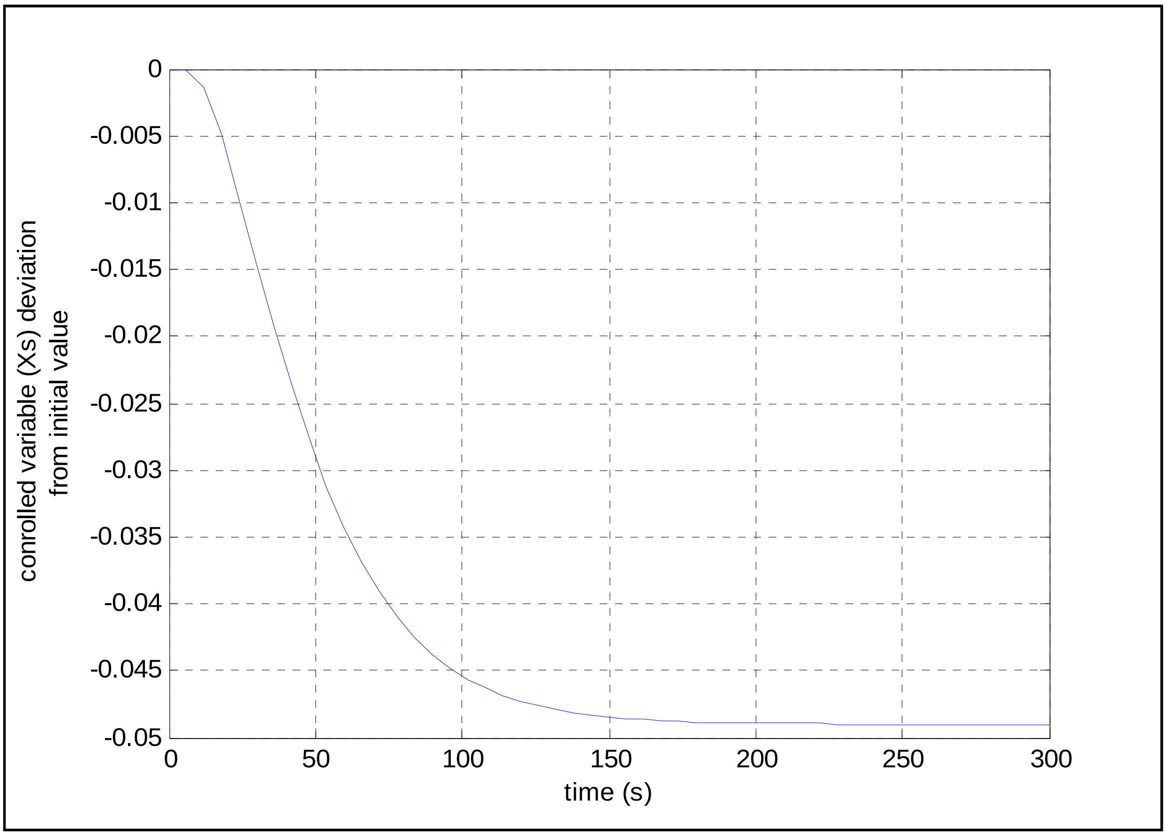


Figure 4.1.b.3 effect of R on X_S for 10% step change for open loop process

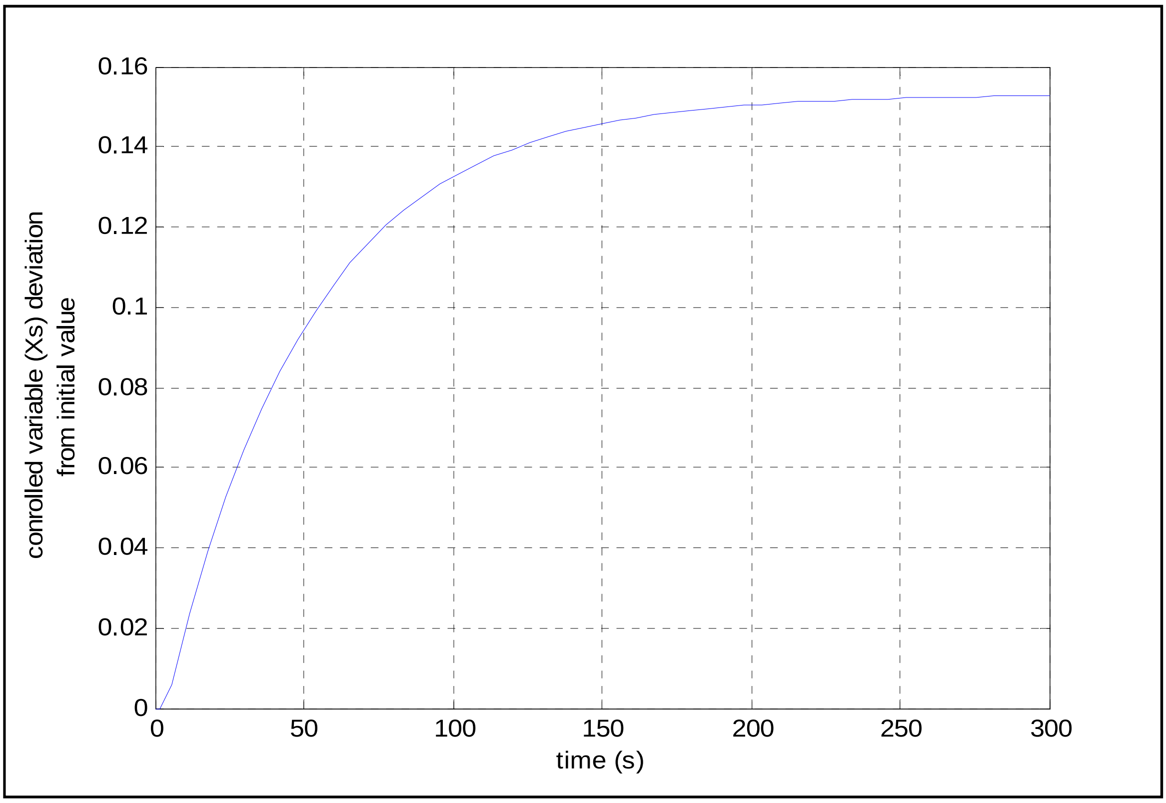


Figure 4.1.b.4 effect of L_S on X_S for 10% step change for open loop process

Figure 4.1.b.1 shows the response of distillate composition (X_D) via applying 10% step change on reflux flow rate (R). The result shows that distillate composition (X_D) increase with increasing reflux flow rate (R) and then rapidly reaches the steady state value.

Figure 4.1.b.2 shows the response of distillate composition (X_D) via applying 10% step change on side stream flow rate (L_S). The result shows that distillate composition (X_D) decrease with increasing side stream flow rate (L_S) and then rapidly reaches the steady state value.

Figure 4.1.b.3 shows the response of side stream composition (X_S) via applying 10% step change on reflux flow rate (R). The result shows that side stream composition (X_S) decrease with increasing reflux flow rate (R) and then rapidly reaches the steady state value.

Figure 4.1.b.4 shows the response of side stream composition (X_S) via applying 10% step change on side stream flow rate (L_S). The result shows that side stream composition (X_S) decrease with increasing side stream flow rate (L_S) and then rapidly reaches the steady state value.

4.3 close loop process without controller

The results of the transient response based on close loop system are shown in figure (4.2).

10% step change of manipulating variables which is reflux flow rate (R) and side stream flow rate (L_S) on controlled variable which is distillate composition (X_D) and side stream composition (X_S) was applied.

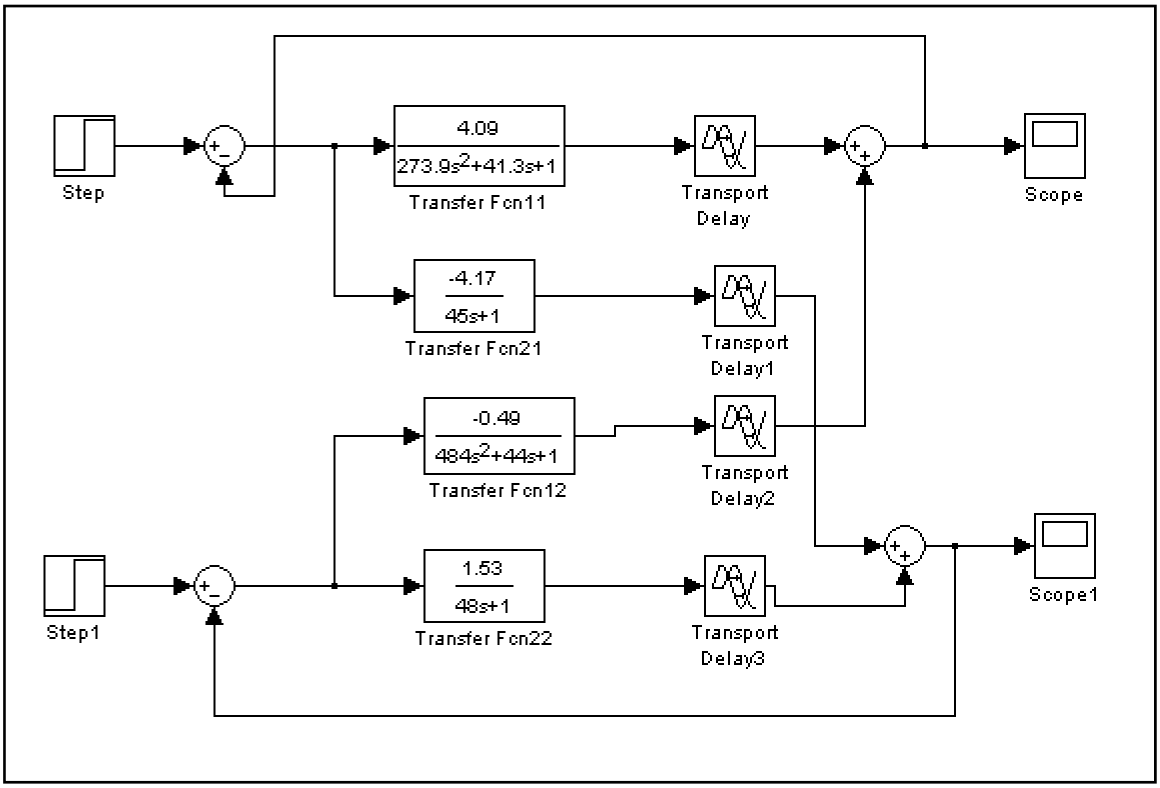


Figure 4.2.a Block diagram of close loop process

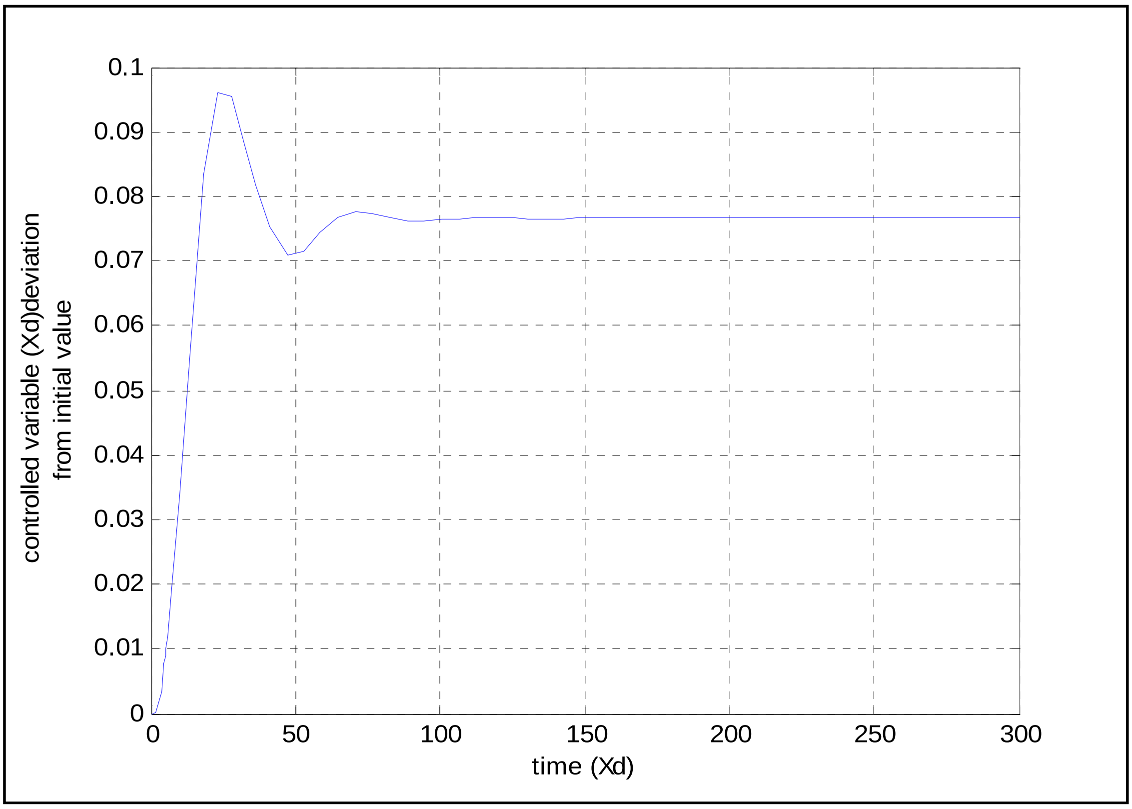


Figure 4.2.b.1 effect of R on X_D for 10% step change for close loop process

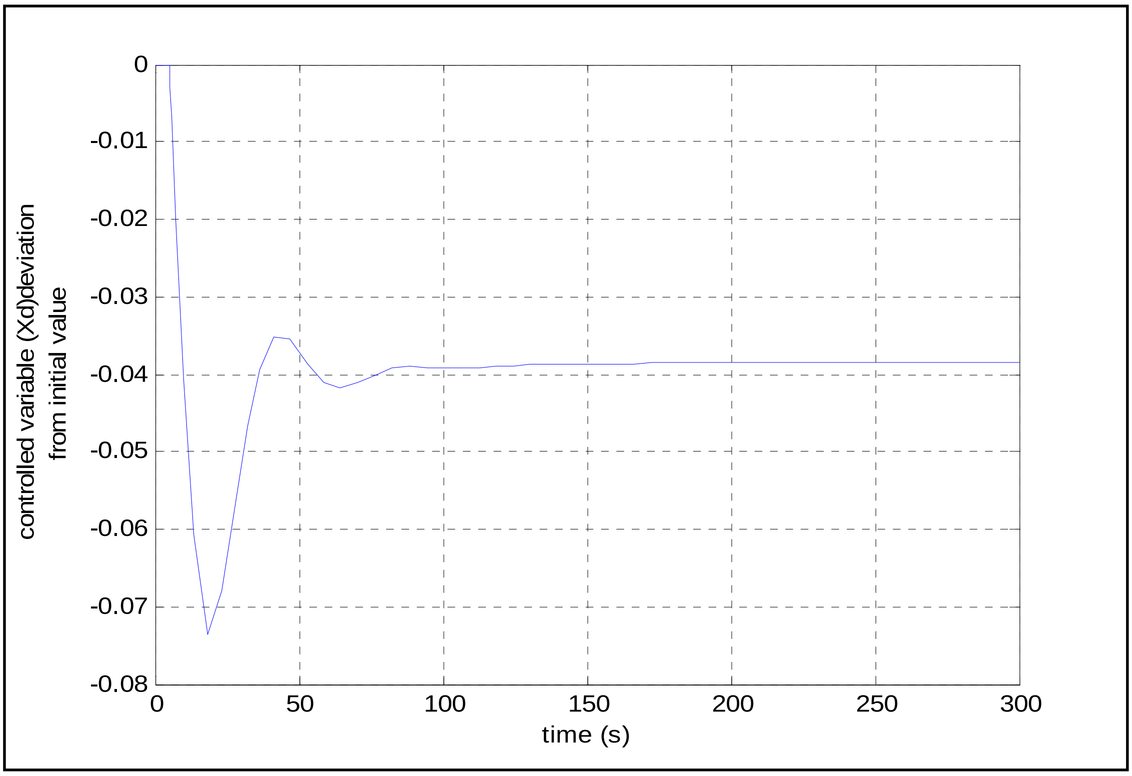


Figure 4.2.b.2 effect of L_S on X_D for 10% step change for close loop process

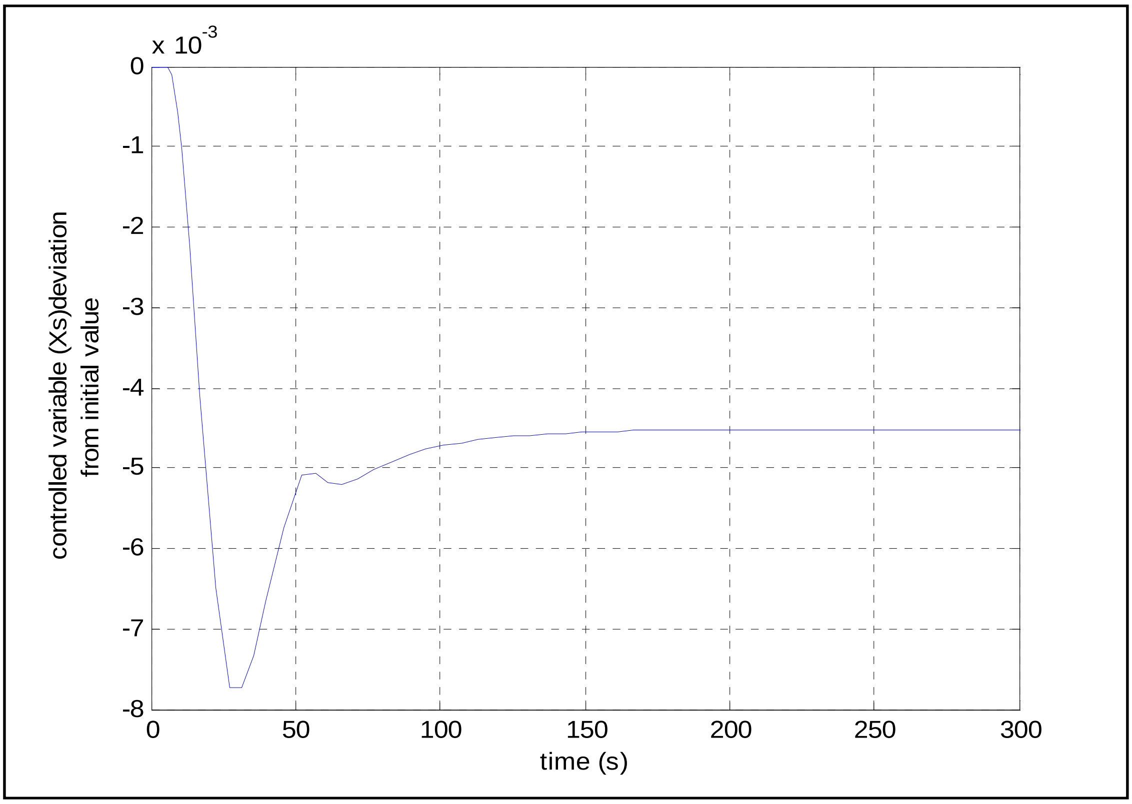


Figure 4.2.b.3 effect of R on X_S for 10% step change for close loop process

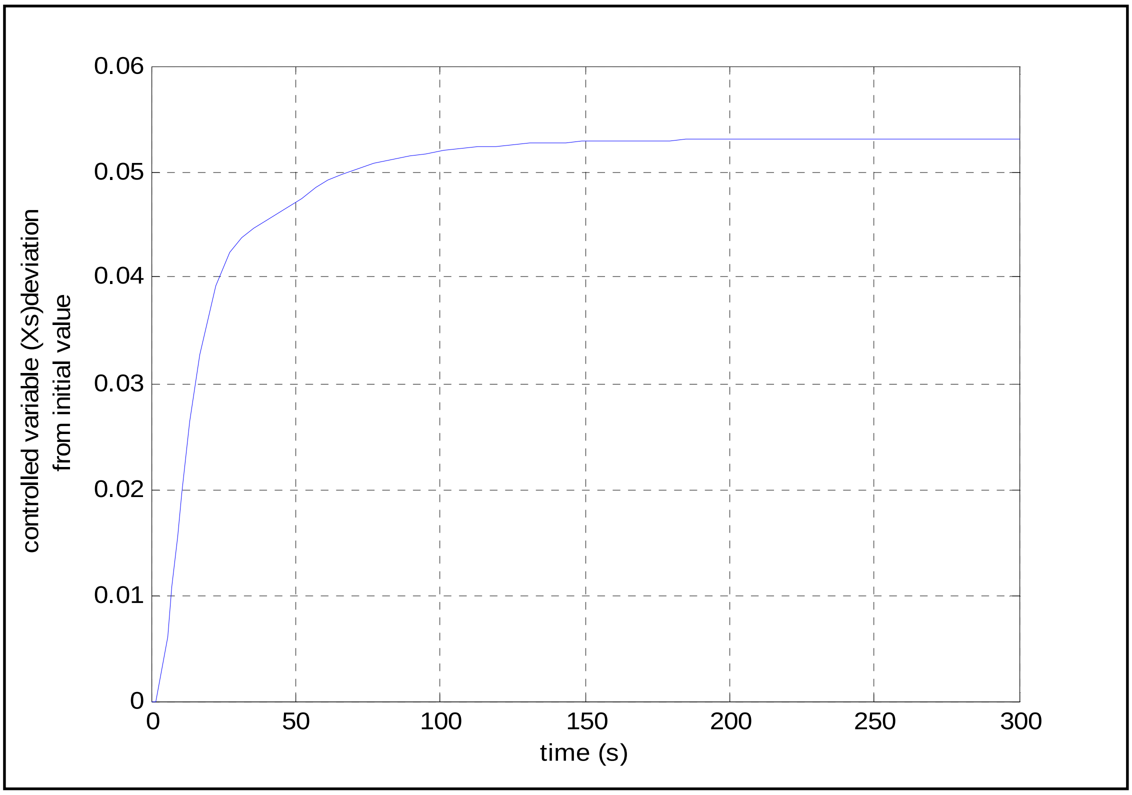


Figure 4.2.b.4 effect of L_S on X_S for 10% step change for close loop process

Figure 4.2.b.1 shows the response of distillate composition (X_D) via applying 10% step change on reflux flow rate (R). The result shows that distillate composition (X_D) increase with increasing reflux flow rate (R) and then rapidly reaches the steady state value.

Figure 4.2.b.2 shows the response of distillate composition (X_D) via applying 10% step change on side stream flow rate (L_S). The result shows that distillate composition (X_D) decrease with increasing side stream flow rate (L_S) and then rapidly reaches the steady state value.

Figure 4.2.b.3 shows the response of side stream composition (X_S) via applying 10% step change on reflux flow rate (R). The result shows that side stream composition (X_S) decrease with increasing reflux flow rate (R) and then rapidly reaches the steady state value.

Figure 4.2.b.4 shows the response of side stream composition (X_S) via applying 10% step change on side stream flow rate (L_S). The result shows that side stream composition (X_S) decrease with increasing side stream flow rate (L_S) and then rapidly reaches the steady state value.

4.4 Frequency Response Analysis

By using Bode plots we can estimate the frequency response of the open loop and close loop systems and its stability, as shown in the following figures

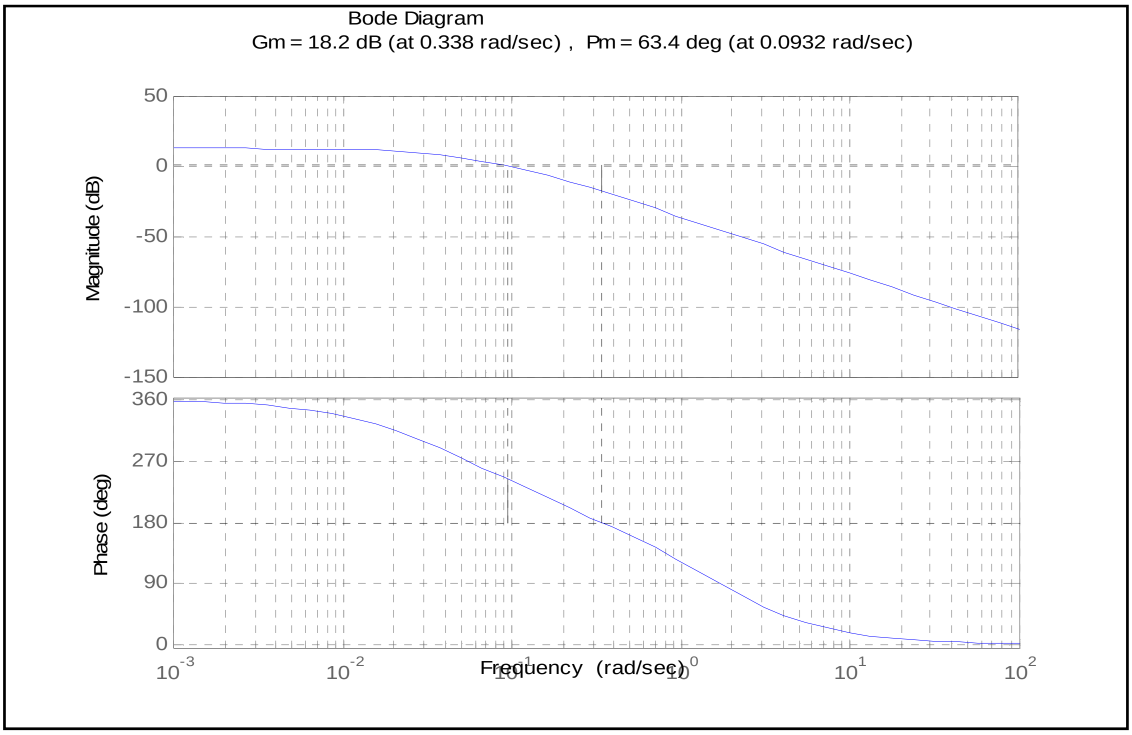


Figure 4.3 The frequency response of distillate composition to reflux flow rate.

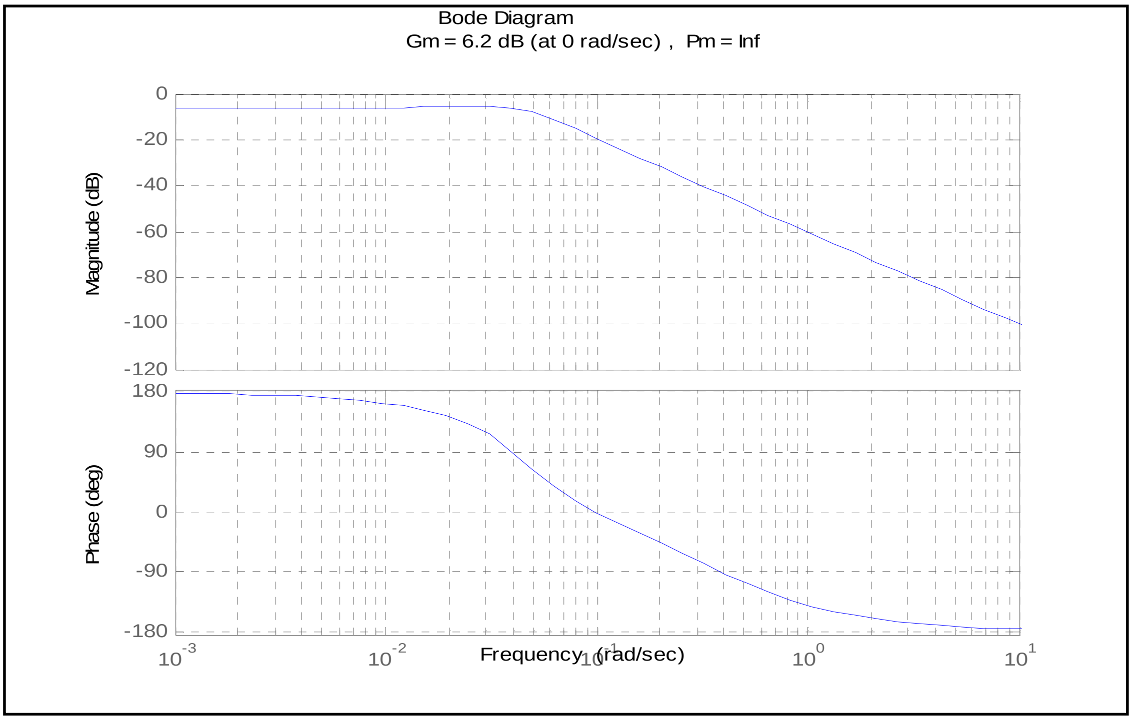


Figure 4.4 The frequency response of distillate composition to side stream flow rate.

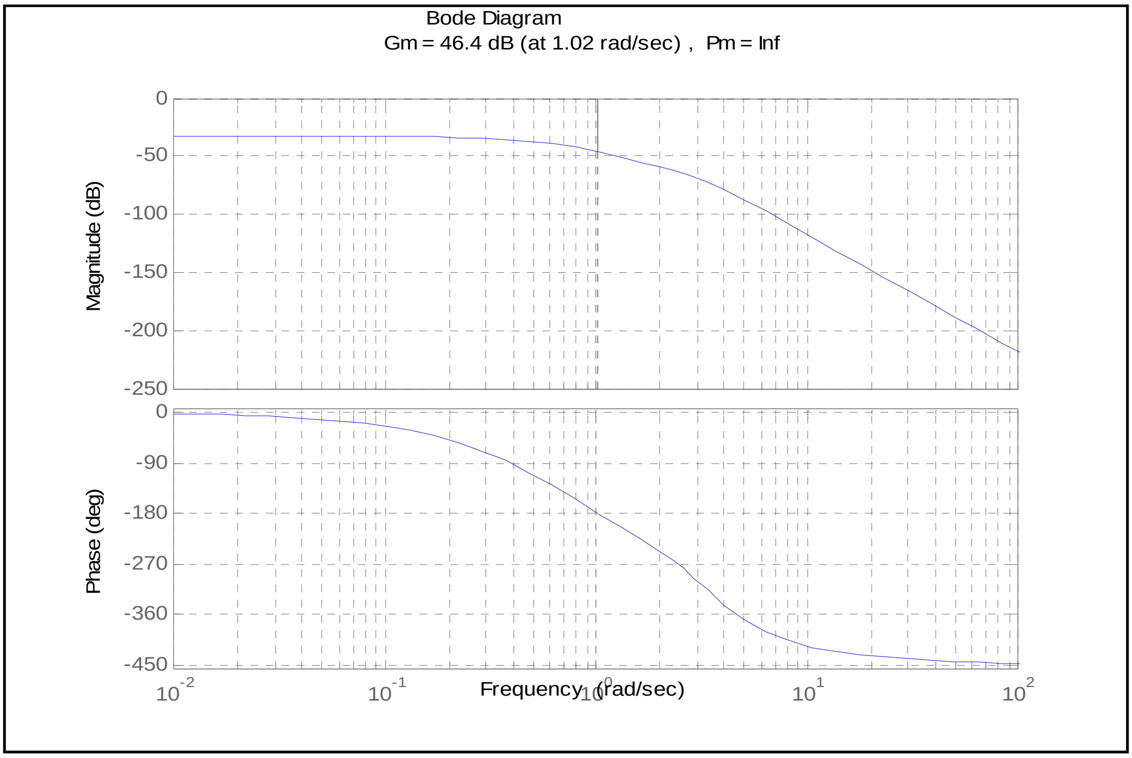


Figure 4.5 The frequency response of side stream composition to reflux flow rate.

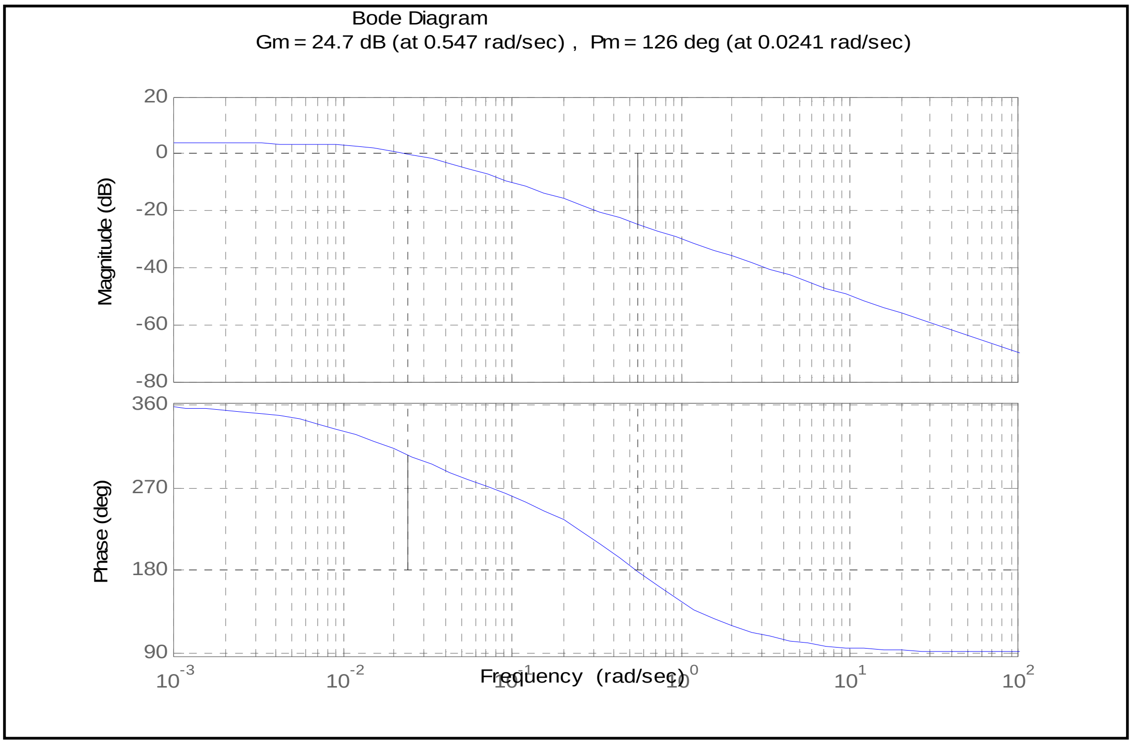


Figure 4.6 The frequency response of side stream composition to side stream flow rate.

The results of these figures shown in **table 4.1**

Table 4.1 The characteristics of the system

	gain margin(db)	phase margin (degree)	phase crossover (frequency)	gain crossover (frequency)
distillate composition to reflux flow	18.2	63.4	0.0932	0.338
distillate composition to side stream flow	6.2	Inf	non	0
side stream composition to reflux flow	46.4	Inf	non	1.02
side stream composition to side stream flow	24.7	126	0.0241	0.547

These results show that the open-loop system is stable due to positive gains margin and phases margin [69].

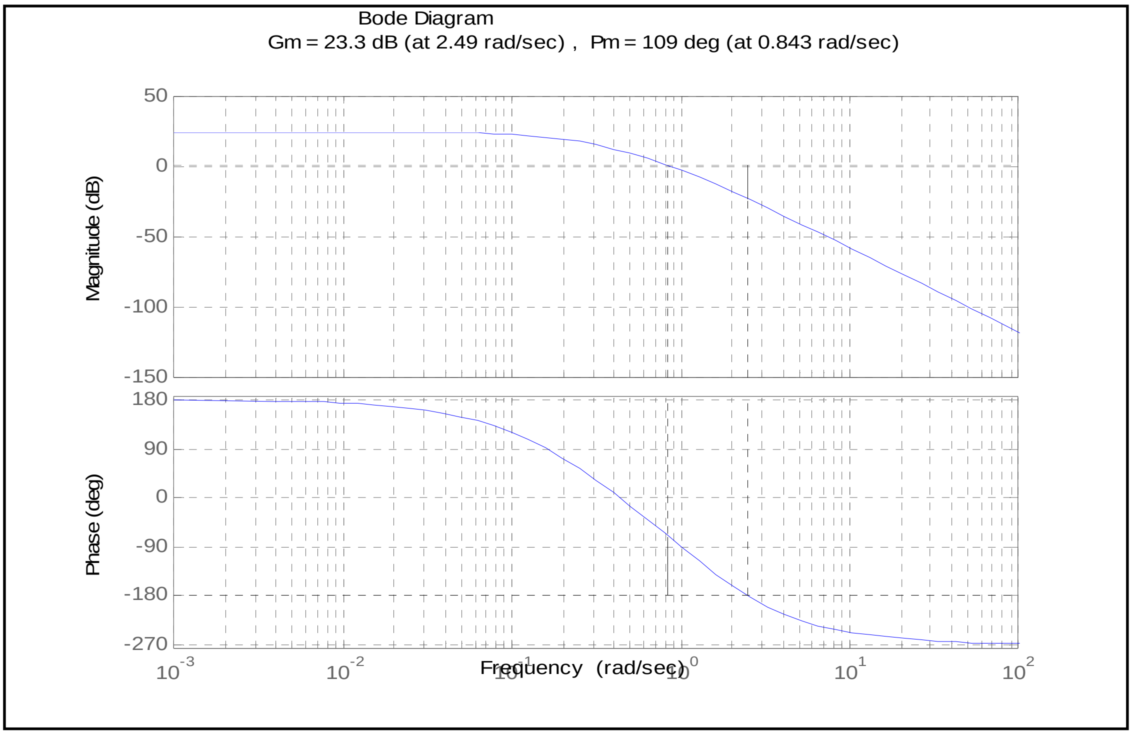


Figure 4.7 The frequency response of distillate composition to reflux flow rate.

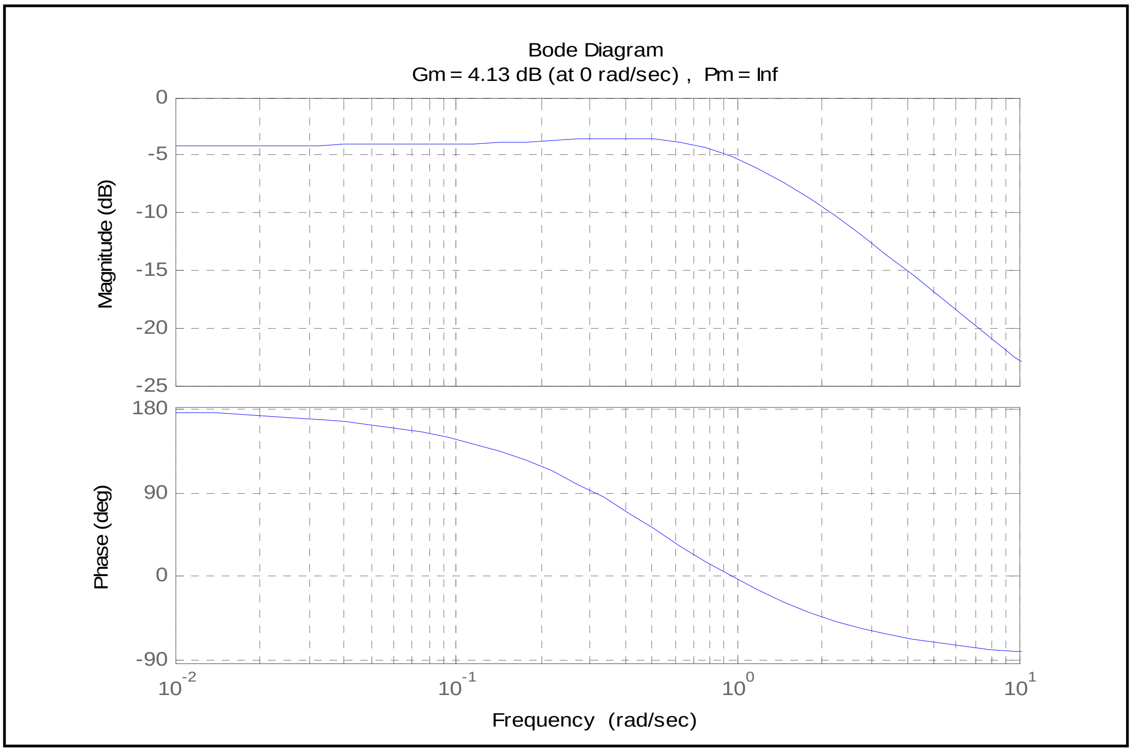


Figure 4.8 The frequency response of distillate composition to side stream flow rate.

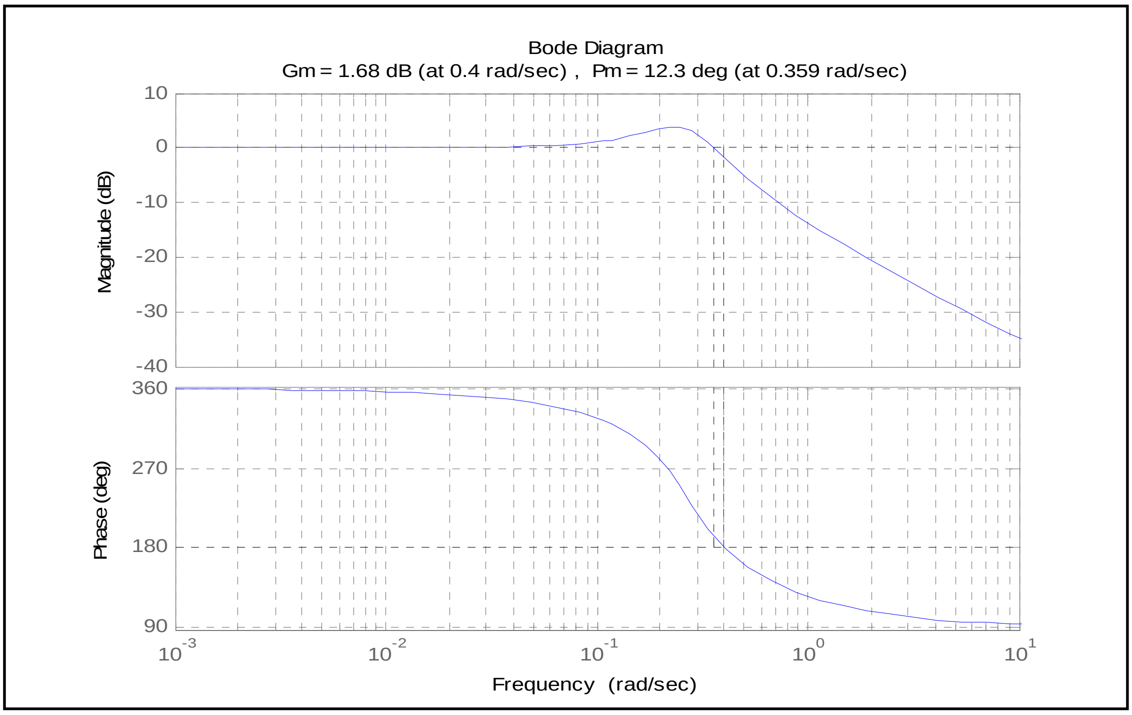


Figure 4.9 The frequency response of side stream composition to reflux flow rate.

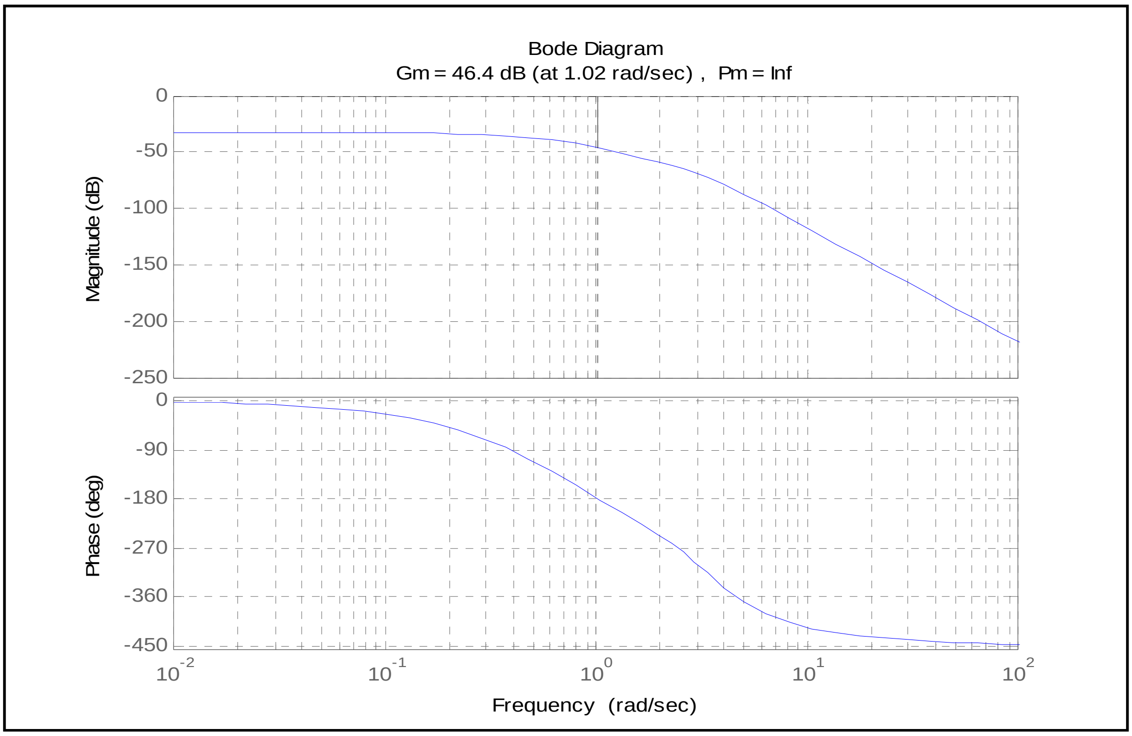


Figure 4.10 The frequency response of side stream composition to side stream flow rate.

The results of these figures shown in **table 4.2**

Table 4.2 The characteristics of the system

	gain margin(db)	phase margin (degree)	phase crossover (frequency)	gain crossover (frequency)
distillate composition to reflux flow	23.3	109	0.843	2.49
distillate composition to side stream flow	4.13	Inf	Non	0
side stream composition to reflux flow	1.68	12.3	0.359	0.4
side stream composition to side stream flow	46.4	Inf	Non	1.02

These results show that the close-loop system is stable due to positive gains margin and phases margin [69].

4.5 interaction control loops

Whenever a single manipulated variable can significantly affect two or more controlled variables, the variables are said to be coupled and there is interaction between loops, this interaction can be troublesome. Some variables are difficult enough to be controlled because of being subjected to upsets from other loops. The following figures show the response of the interaction between loops when applying PID controller on the system.

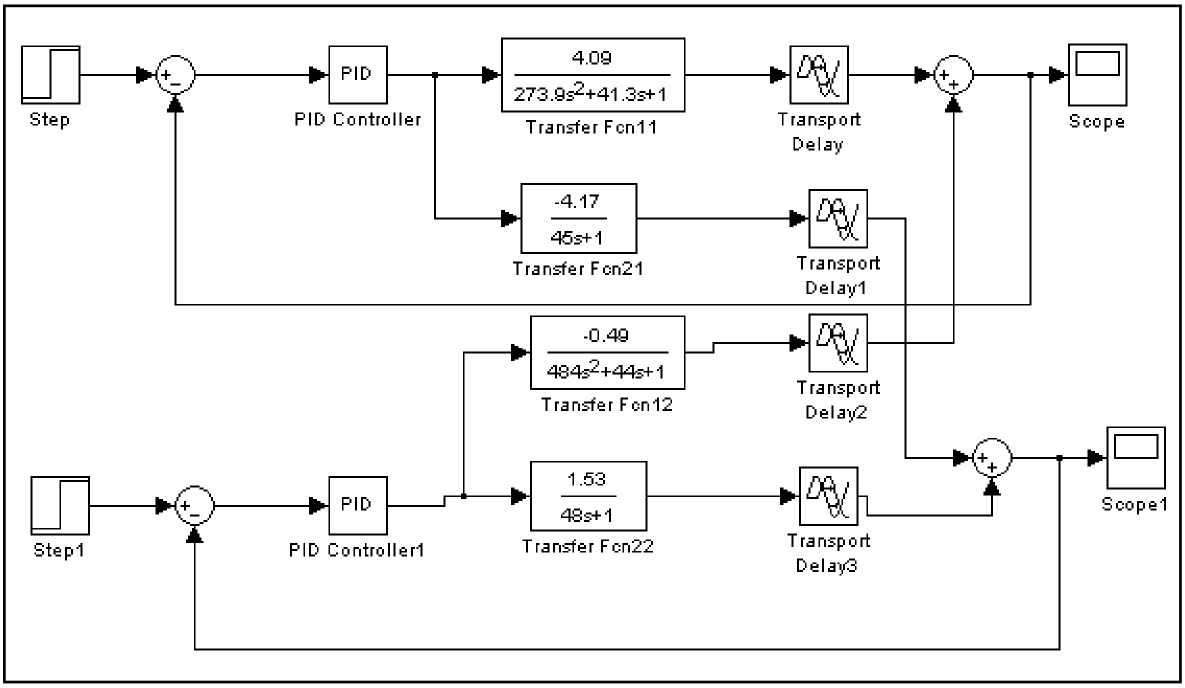


Figure 4.11.a Block Diagram for Interaction control system

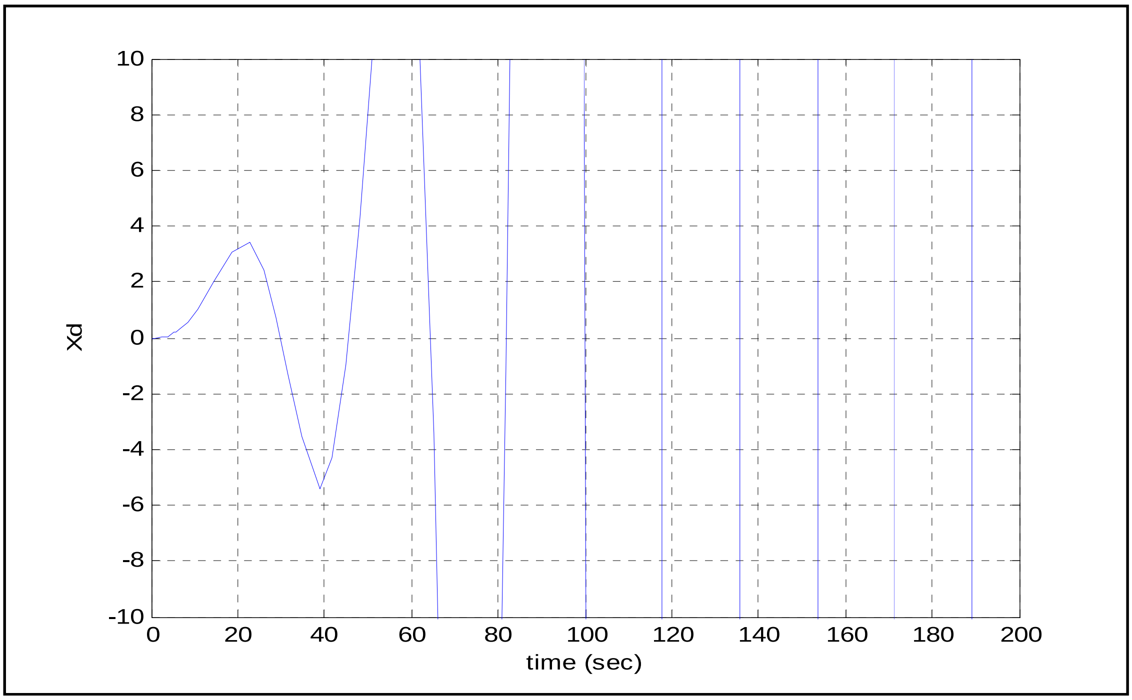


Figure 4.11.b.1 transient response of X_D with respect to R with interaction

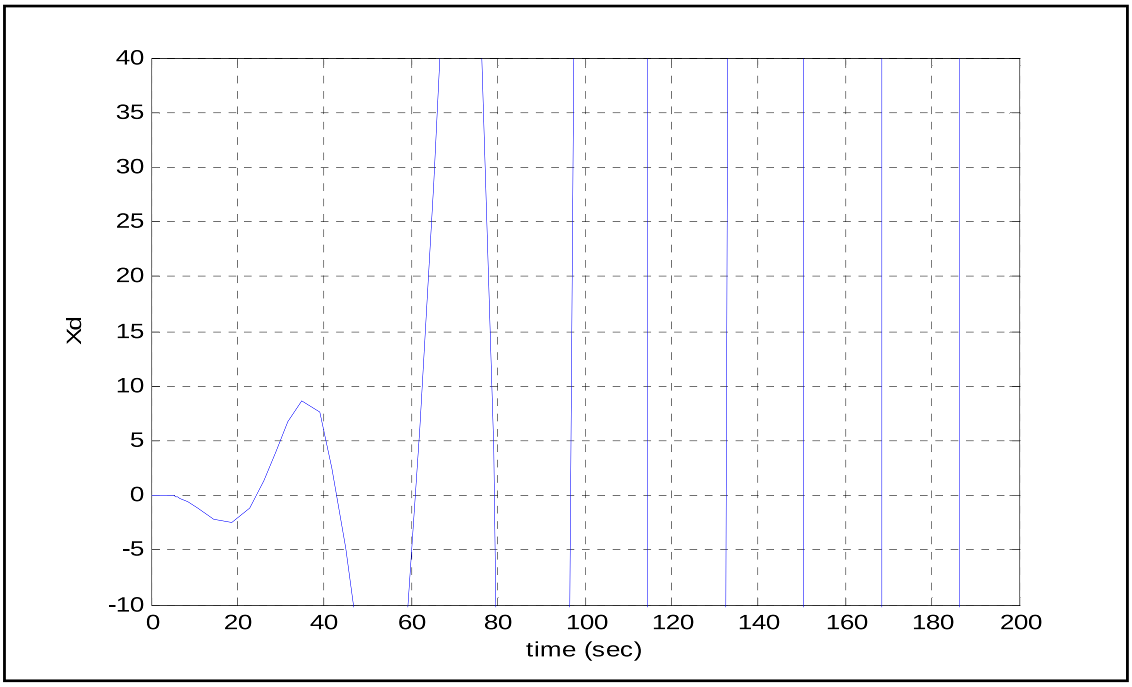


Figure 4.11.b.2 transient response of X_D with respect to L_S with interaction

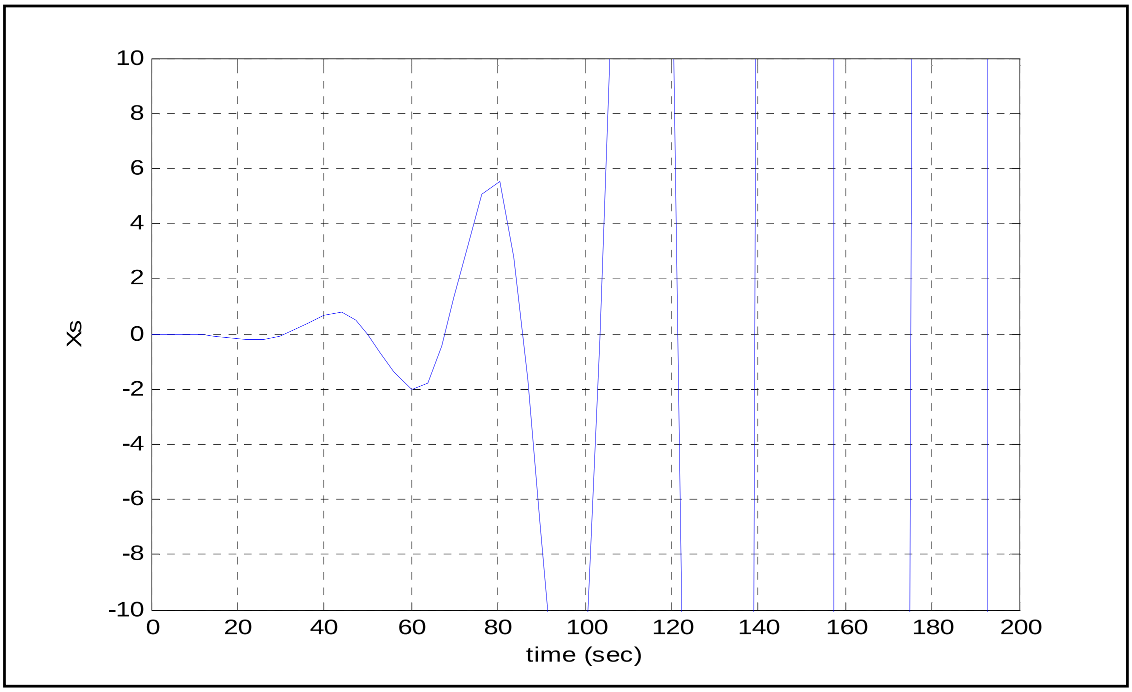


Figure 4.11.b.3 transient response of X_S with respect to R with interaction

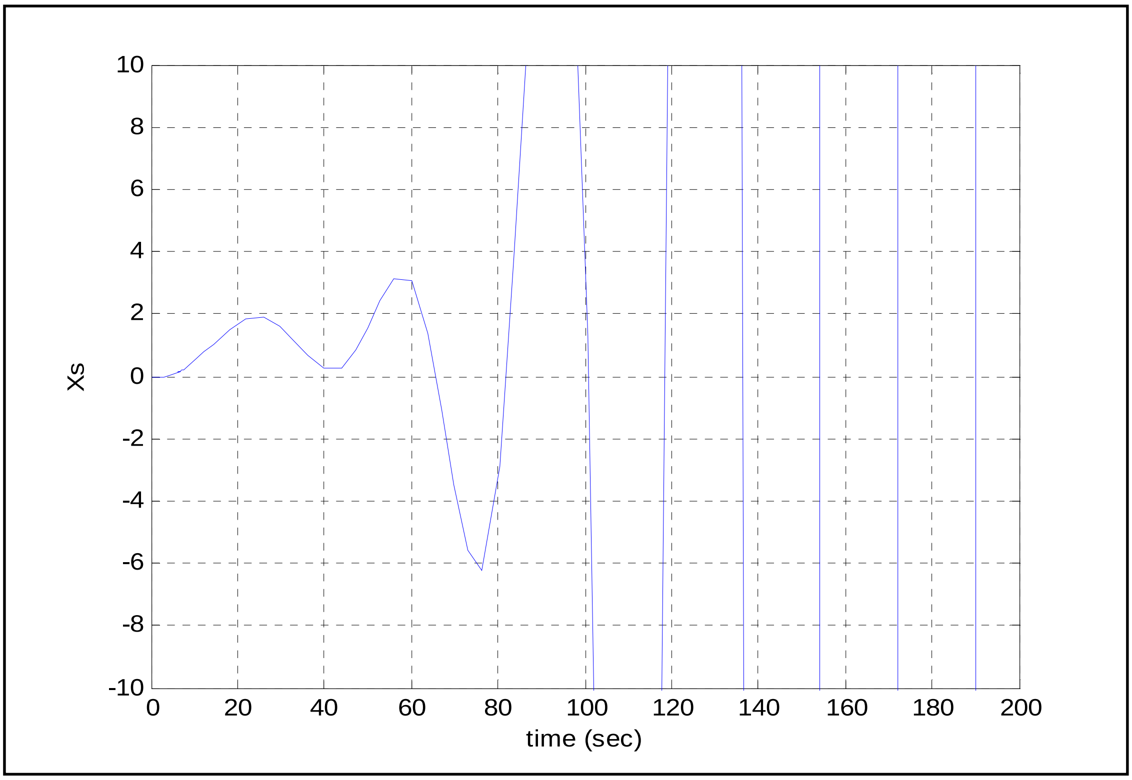


Figure 4.11.b.4 transient response of X_S with respect to L_S with interaction

As shown above in **figure 4.11** there is a wide interaction between loops so decoupling control system is recommended to remove the interaction.

4.6 Relative Gain Array (RGA) Calculations

Relative Gain Array (RGA) must be calculated to choose the best pairing of the two controlled variables (X_D and X_S) and the two manipulated variables (R and L_S) before applying the control techniques. In this work, the results of RGA calculation were obtained by using computer simulation program.

$$RGA=\begin{bmatrix}\lambda_{11} & \lambda_{12}\\ \lambda_{21} & \lambda_{22}\end{bmatrix} = \begin{bmatrix}1.4848 & -0.4848\\ -0.4848 & 1.4848\end{bmatrix} \qquad \ldots (4.6)$$

So the best coupling are obtained by pairing the distillate composition (X_D) with side stream flow rate (L_S), and the side stream composition (X_S) with reflux flow rate (R), since λ_{11} has the largest positive number of the array [67].

Frey et al. examined the control variable pairing by using the relative gain array. Their results indicated that decoupling of the column was possible by proper selection and pairing of the manipulated and control variables. It was shown that the column decoupled at steady-state compositions independently to design parameters [32].

4.7 Decoupler Design

The decoupler of loop1 ($D_{12}(s)$) was designed to eliminate the effect of interaction of loop2 on loop1 by using Equation (3.23) on substitution the values of G_{m11} and G_{m12} the decoupler shows the following value:

$$D_{12}(s) = \frac{-315.37s^3+86.64s^2+19.09s+0.49}{4653s^3+2403s^2+189.61s+4.09} \quad \ldots (4.7)$$

The value of $D_{12}(s)$ is coupled with the value of the main reflux ratio (R) to get the final value, after each time interval.

In the same way, the decoupler of loop2 ($D_{21}(s)$) was designed to eliminate the effect the loop1 on loop2. After applying the value of G_{21} and G_{m22} the value is:

$$D_{21} = \frac{-120s^2+197.7s+4.17}{41.3s^2+69.76s+1.53} \quad \ldots (4.8)$$

The decoupler was obtained to justify the main value of side stream flow rate (L_S).

So the decoupling system becomes:

$$\begin{bmatrix} 1 & \frac{-315.37s^3+86.64s^2+19.09s+0.49}{4653s^3+2403s^2+189.61s+4.09} \\ \frac{-120s^2+197.7s+4.17}{41.3s^2+69.76s+1.53} & 1 \end{bmatrix} \quad \ldots (4.9)$$

Bode plot was used to check the stability of the system as show below:

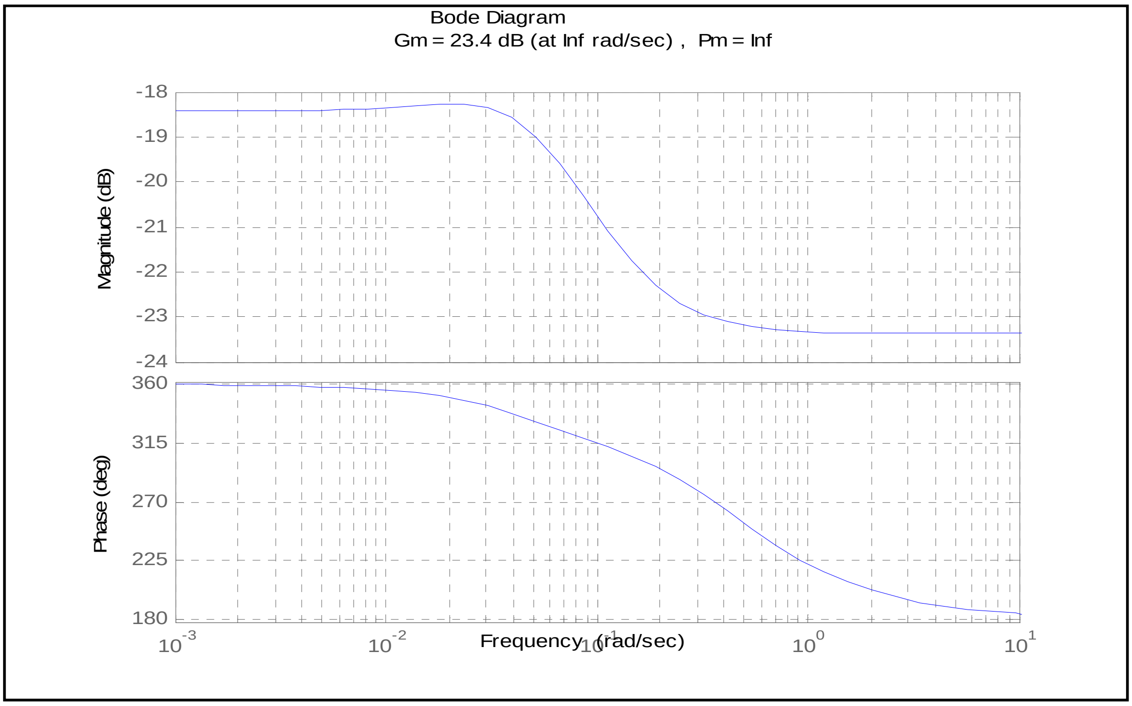


Figure 4.12 Bode plot for decoupler 12

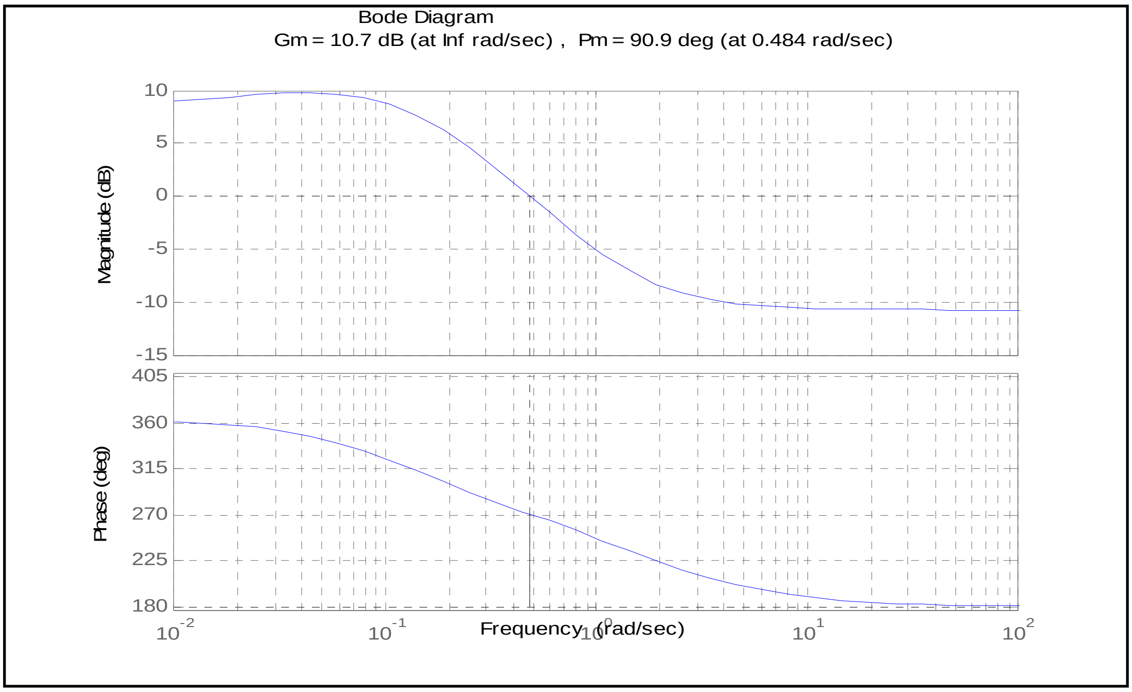


Figure 4.13 Bode plot for decoupler 21

The results of these figures shown in table 4.3

Table 4.3 The characteristics of the decoupling system

	gain margin(db)	phase margin (degree)	phase crossover (frequency)	gain crossover (frequency)
Decoupler 12	23	Inf.	Non	Inf.
Decoupler 21	10	90.9	0.484	Inf.

These results show that the decoupling system is stable due to positive gains margin and phases margin.

Figures (4.14) and (4.15) show the response of decouplers and it's found that the decouplers are greatly improving the response of the system.

Luyben presented a quantitative study of two types of decoupling elements to achieve a non-interacting feedback control of overhead and bottom compositions in binary distillation, ideal and simplified decoupling. It was concluded that in ideal decoupling, leading to unstable feedback loops, while simplified decoupling was effective and stable and appeared to be easily implemented with commercial control instrumentation [24].

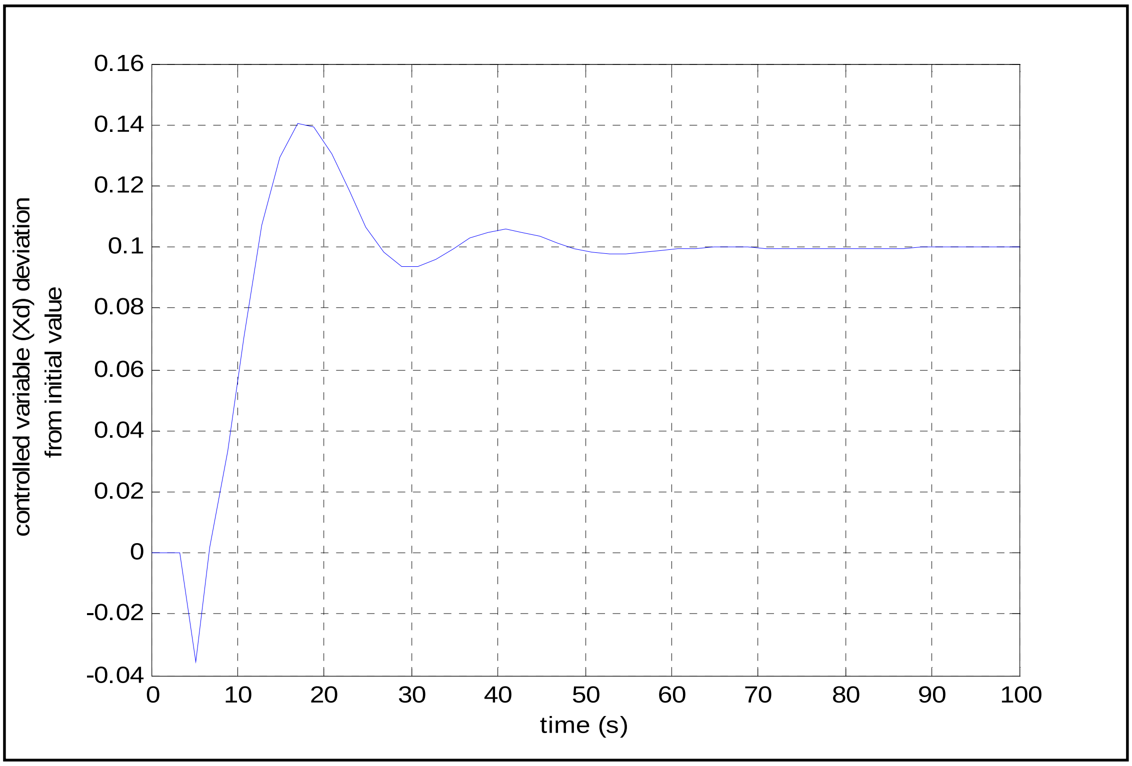


Figure 4.14 transient response of decoupler 12

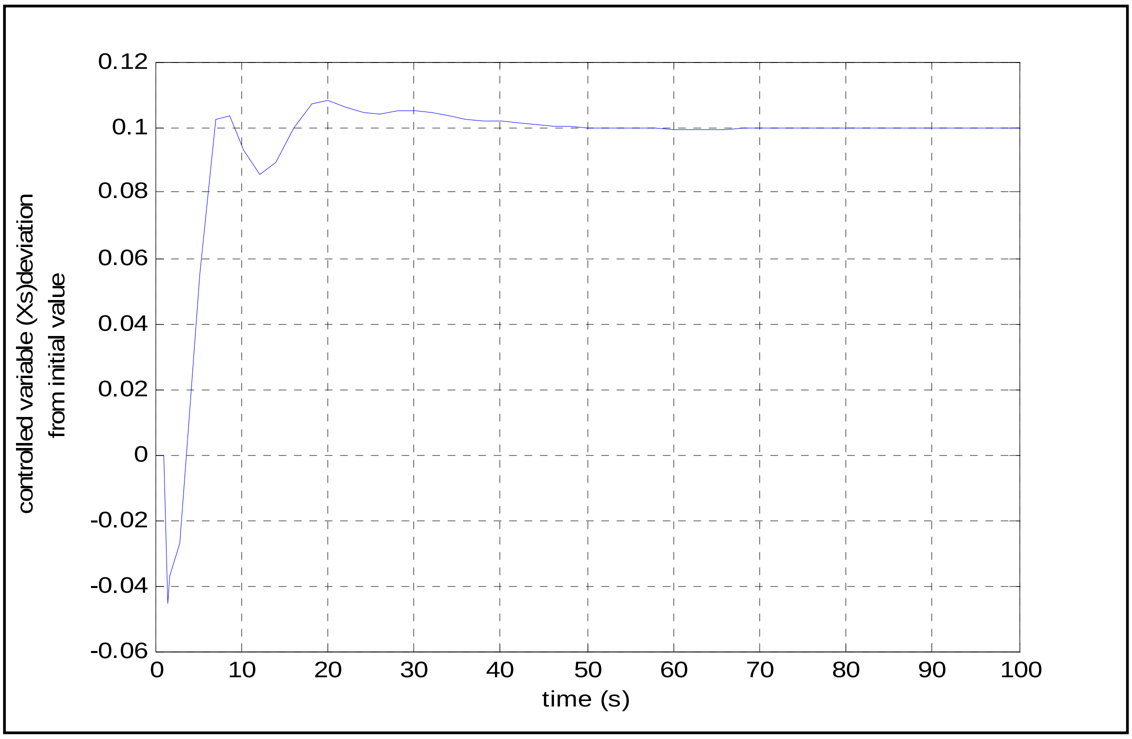


Figure 4.15 transient response of decoupler 21

4.8Control Strategies

In this section two different control strategies were used, the Feedback control and Fuzzy logic control.

4.8.1 Feedback control

P, PI and PID controller modes were used in Feedback system to control the composition of both distillate and side stream; therefore, tuning the control parameters (proportional gain (K_c), time integral (τ_I) and time derivative (τ_D)) must be applied.

Table4.4 control parameters for X_D with R with feedback controller

Method	P-Controller	PI-Controller		PID-Controller		
	K_c	K_c	τ_I	K_c	τ_I	τ_D
Ziegler-Nicols	3.5	3.18	15.83	4.11	9.5	2.35
Tyreus-luyben	_	2.187	41.8	3.18	41.8	3.015
Cohen-Coon	3.194	2.82	3.74	4.21	3.1	0.466

Table4.5 control parameters for X_D with L_S with feedback controller

Method	P-Controller	PI-Controller		PID-Controller		
	K_c	K_c	τ_I	K_c	τ_I	τ_D
Ziegler-Nicolas	-27.5	-25	24.16	-32.35	14.5	3.625
Tyreus-luyben	_	-17.18	63.8	-25	63.8	4.6
Cohen-Coon	-9.52	-6.9	12.78	-10.48	13.27	2.078

Table4.6 control parameters for X_S with R with feedback controller

Method	P-Controller	PI-Controller		PID-Controller		
	K_c	K_c	τ_I	K_c	τ_I	τ_D
Ziegler-Nicolas	-7.2	-6.54	5	-8.47	3	0.75
Tyreus-luyben	_	-4.5	13.2	-6.545	13.2	0.952
Cohen-Coon	-2.23	-1.96	13.51	-2.93	11.76	1.78

Table4.7 control parameters for X_S with L_S with feedback controller

Method	P-Controller	PI-Controller		PID-Controller		
	K_c	K_c	τ_I	K_c	τ_I	τ_D
Ziegler-Nicolas	20.5	18.6	5	24.11	3	0.75
Tyreus-luyben	_	12.8	13.2	18.6	13.2	0.95
Cohen-Coon	8.47	7.48	10.85	10.48	9.05	1.362

As shown in the **Tables (4.4)-(4.7)**, the control tuning was found in three different methods therefore; it can be seen that the tuning by using the Ziegler-Nicolas method is better than process reaction curve (Cohen-Coon) method because Ziegler-Nicolas method depends on closed loop system while process reaction curve depends on open loop system.

The figures below show the response of the three controllers when applied on the system to choose the best one among them:

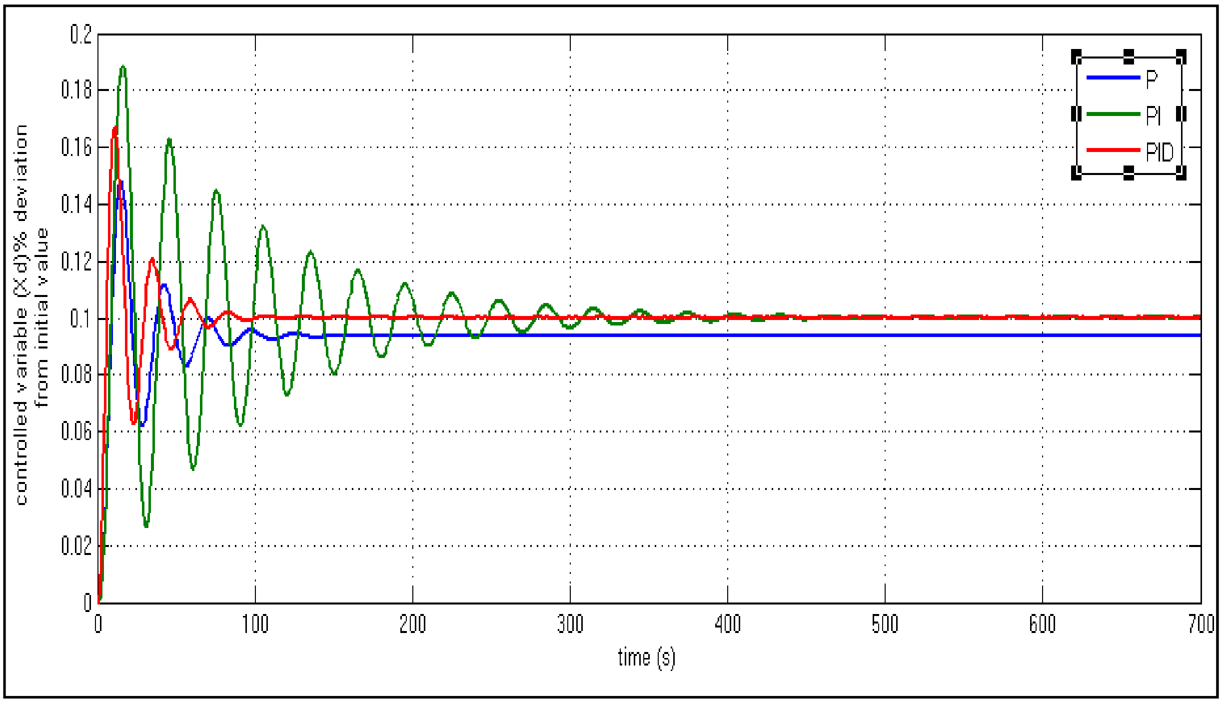


Figure 4.16 Transient response of distillate composition with respect to reflux flow rate using Ziegler-Nicolas parameters

Table 4.8 time responses of distillate composition with respect to reflux flow rate using Ziegler-Nicolas parameters

Controller	Rise Time	Settling Time	Offset
P	15	213	0.015
PI	14	450	0
PID	10	225	0

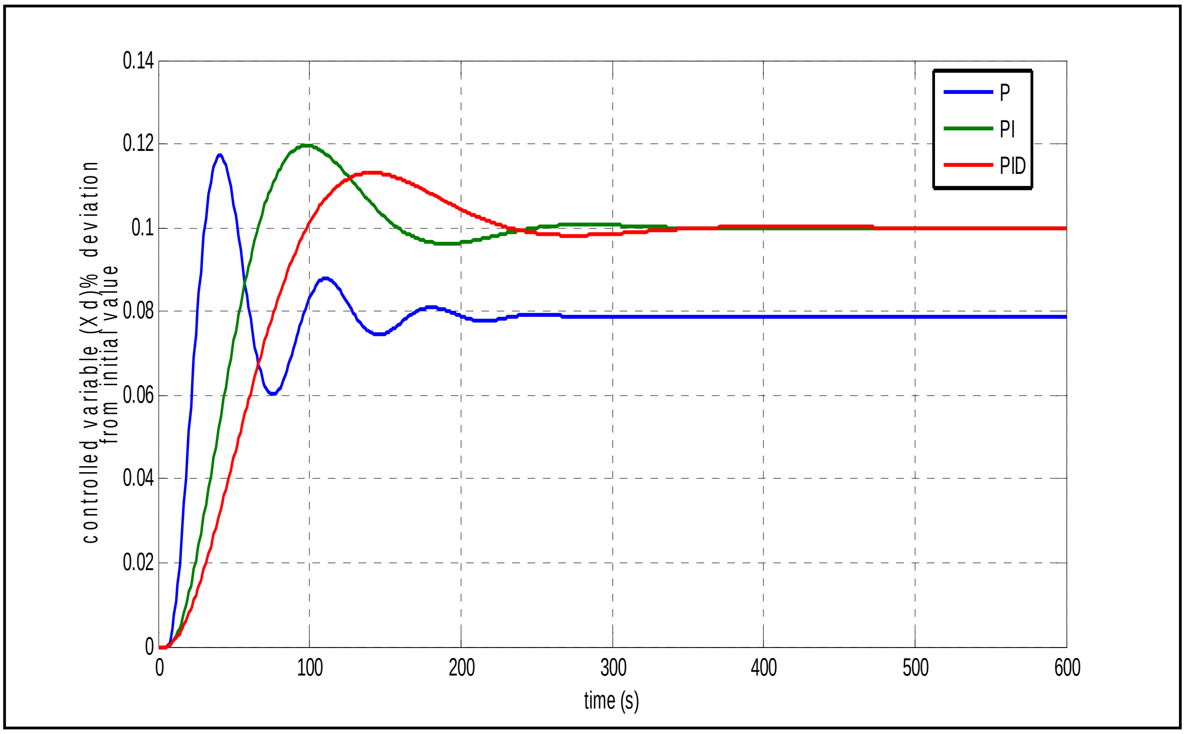


Figure 4.17 Transient response of distillate composition with respect to side stream flow rate using Ziegler-Nicolas parameters

Table 4.9 Time responses of distillate composition with respect to side stream flow rate using Ziegler-Nicolas parameters

Controller	Rise Time	Settling Time	Offset
P	40	290	0.021
PI	85	400	0
PID	135	350	0

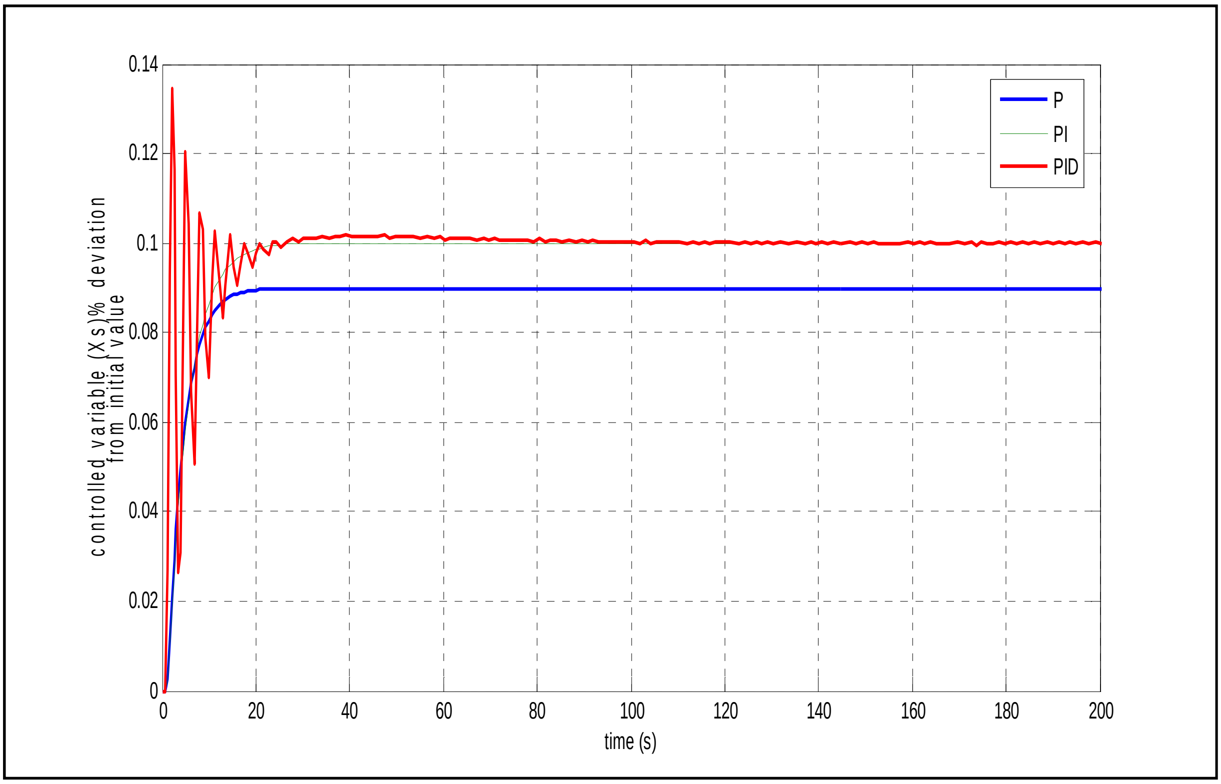


Figure 4.18 Transient response of side stream composition with respect to reflux flow rate using Ziegler-Nicolas parameters

Table 4.10 Time responses of side stream composition with respect to reflux flow rate using Ziegler-Nicolas parameters

Controller	Rise Time	Settling Time	Offset
P	25	25	0.01
PI	18	30	0
PID	5	90	0

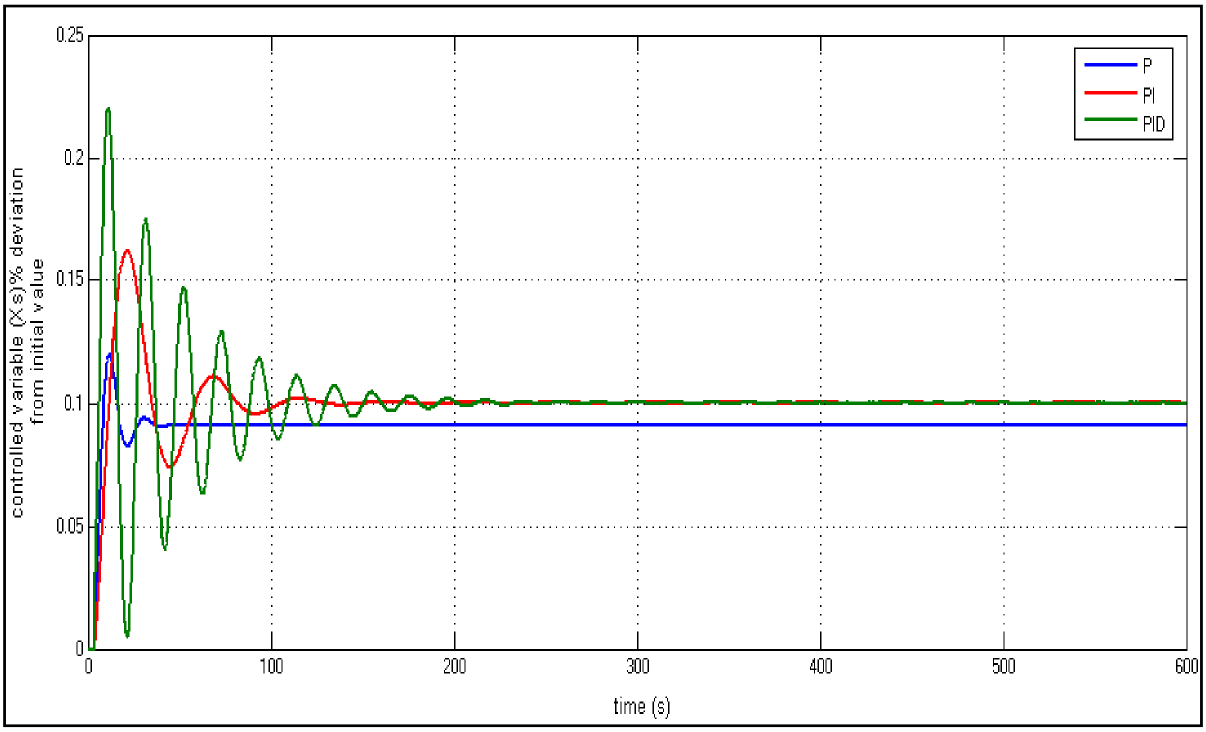


Figure 4.19 Transient response of side stream composition with respect to side stream flow rate using Ziegler-Nicolas parameters

Table 4.11 Time responses of side stream composition with respect to side stream flow rate using Ziegler-Nicolas parameters

Controller	Rise Time	Settling Time	Offset
P	25	25	0.01
PI	18	30	0
PID	5	90	0

Three modes were used in Feedback control; it is clear that PI mode is better than the others because of the good tuning of adjusted parameters values in PI mode which gives the smaller overshoot and makes the system with smaller oscillation and reaches the new steady state value in a shorter time.

A simple practical approaches to the problem of finding reasonable *PI* controller settings of the *N* single-input single-output controllers in an *N*th order typical industrial multivariable process was presented by Luyben [17]. The procedure was straight-forward extension of the familiar Nyquist method and required only nominal computing power. The method was tested on ten multivariable distillation column examples taken from the literature. The resulting settings gave reasonable and stable responses.

4.8.2 Fuzzy Logic Controller (FLC) Behavior

The control tuning of the FLC depends on the trial and error to find the value of the controller gain; therefore, this method was used with *MIMO* system. The optimum values of the controller gains were tuned by using computer simulation program, as shown in Appendix (A).

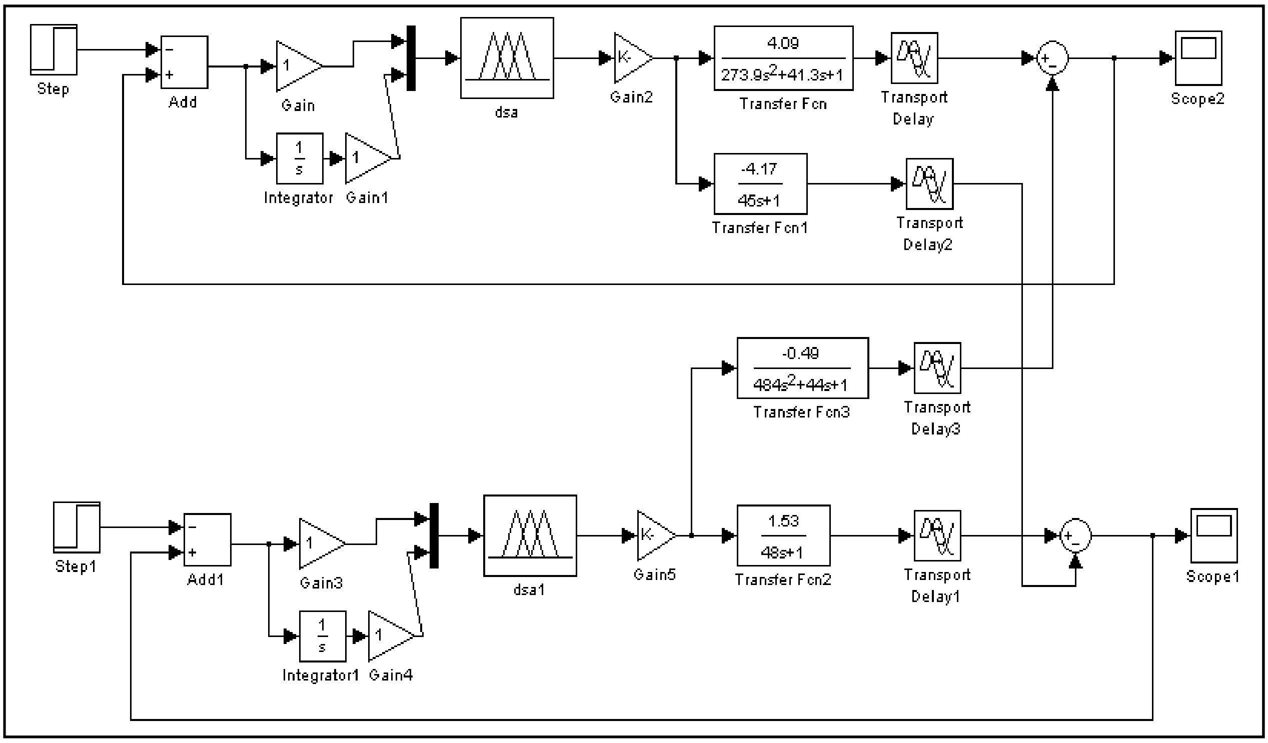


Figure 4.20 block diagram of fuzzy logic system for MIMO system

Table 4.12 Control tuning of fuzzy logic control for *MIMO* system

	Controller gain K_C
distillate loop	0.043
side stream loop	0.045

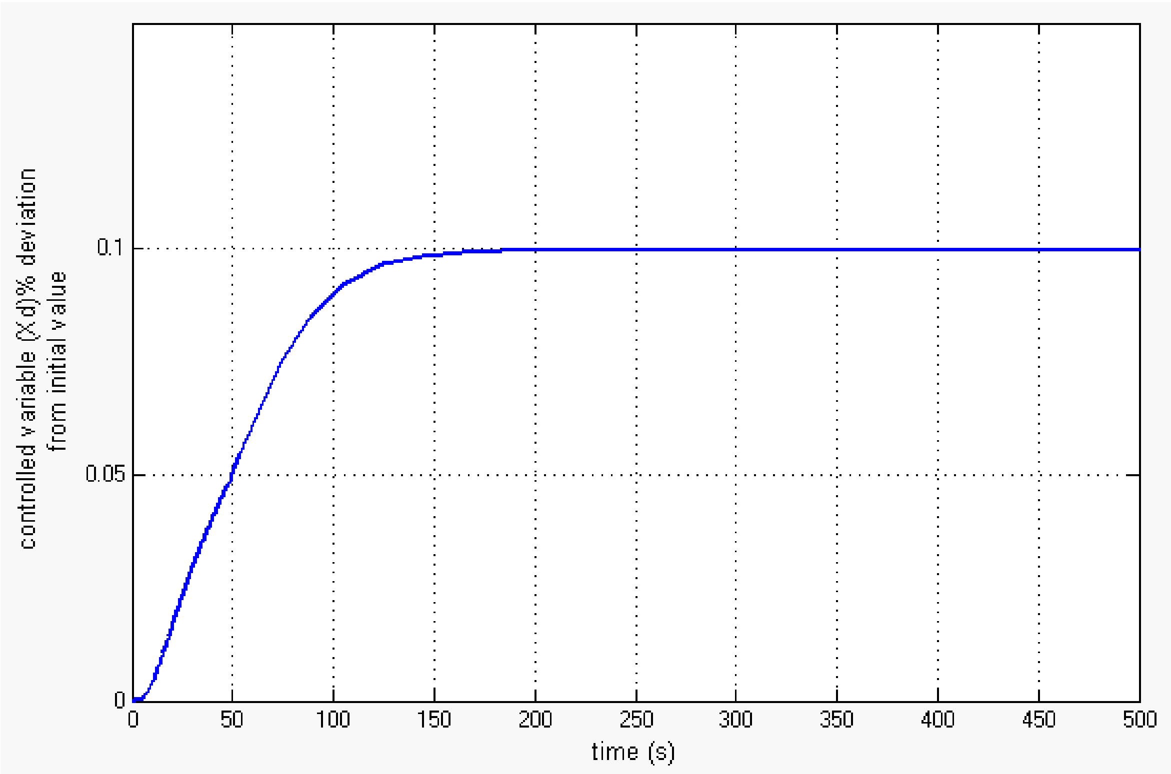


Figure 4.21 Transient response of distillate composition with respect to reflux flow rate in fuzzy logic controller using *MIMO* system

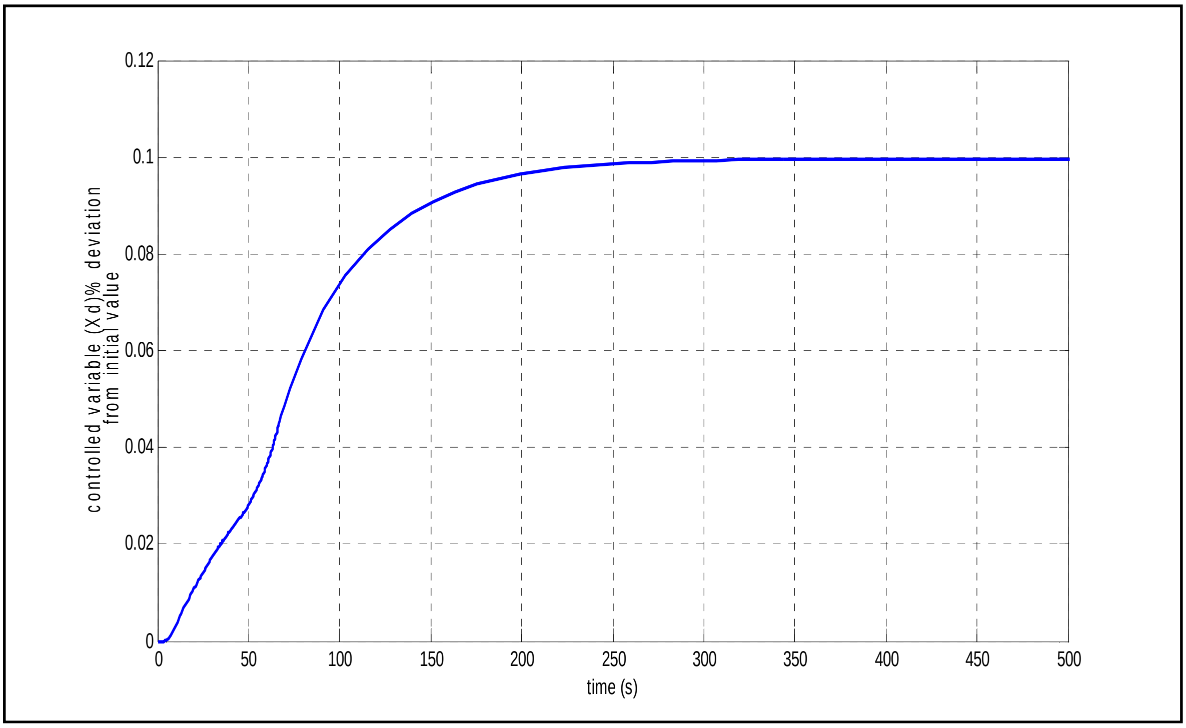


Figure 4.22 Transient response of distillate composition with respect to side stream flow rate in fuzzy logic controller using *MIMO* system

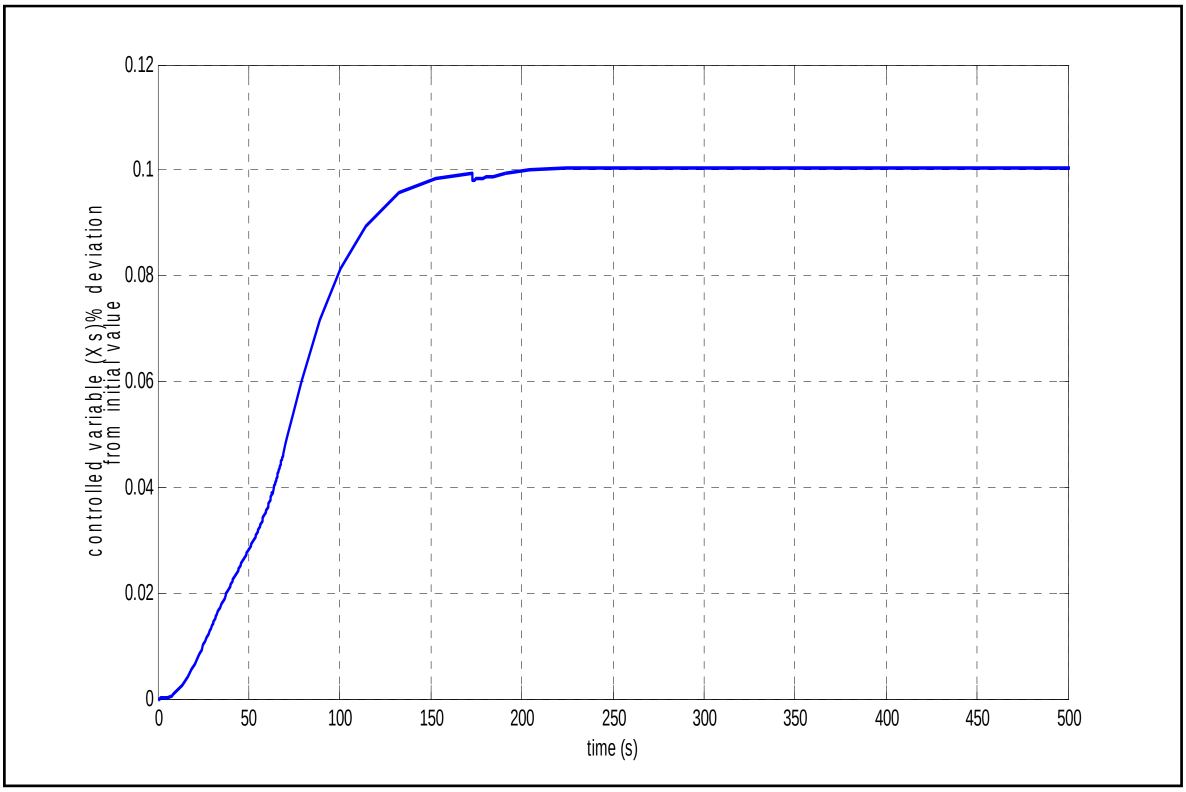


Figure 4.23 Transient response of side stream composition with respect to reflux flow rate in fuzzy logic controller using *MIMO* system

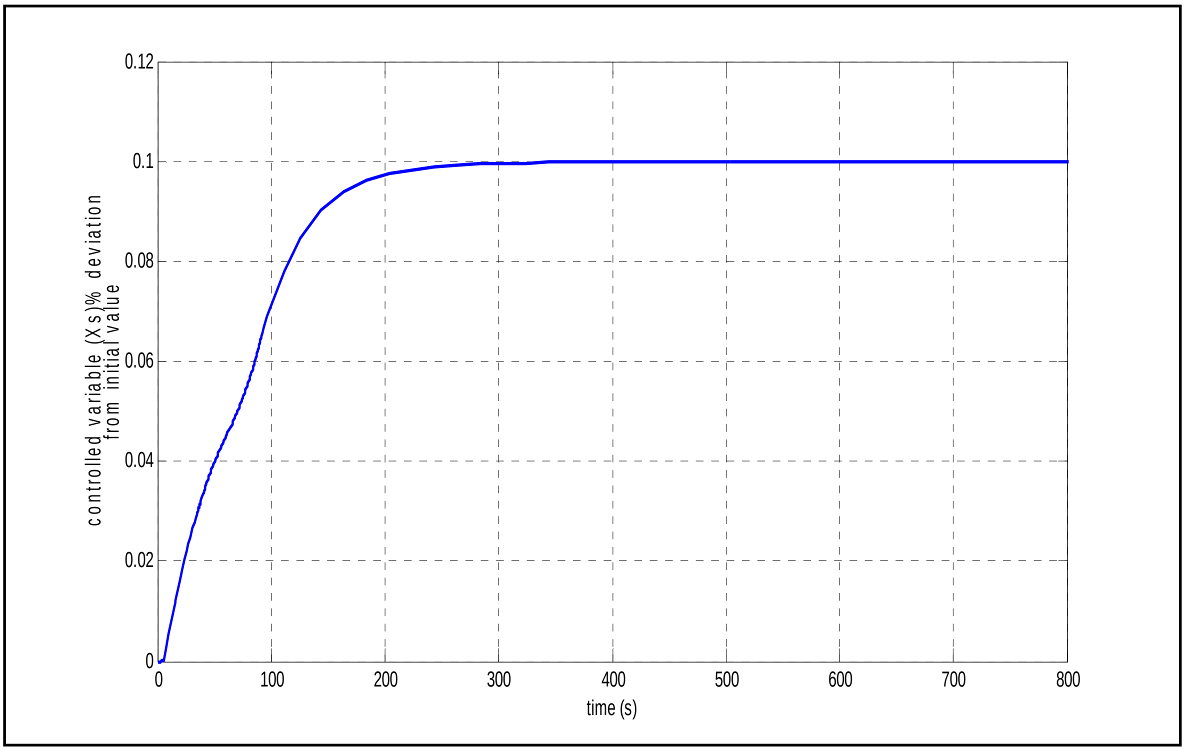


Figure 4.24 Transient response of side stream composition with respect to side stream flow rate in fuzzy logic controller using *MIMO* system

Figures (4.21)-(4.24) show the transient response using controller gain tuning method of distillate and side stream composition for *MIMO* system.

Naoum [48] designed a rule-based fuzzy logic controller for a binary distillation column separating methanol-water mixture. The results obtained when applying fuzzy logic controller to the control of top and bottom temperature were as good, if not better, than those obtained using PI controllers. Fuzzy decouplers were proposed as an alternative to other types of decouplers namely simple steady state decouplers and the results were very promising compared with PI controllers and fuzzy controller without decoupler.

4.8.3 Comparison between PID and Fuzzy Logic Controllers

PID controller was considered for comparison study with FLC because it is still the widely used strategy in industry. To make a clear comparison between these controllers all controllers (FLC and PID Controller) were tuned to the approximately best settings. In general FLC gives better results than PID controller, where the advantage of the FLC is that it does not need a model to build the control settings as in the case of PID controller. Hence the FLC is the effective one for all defined processes as shown in the figures below:

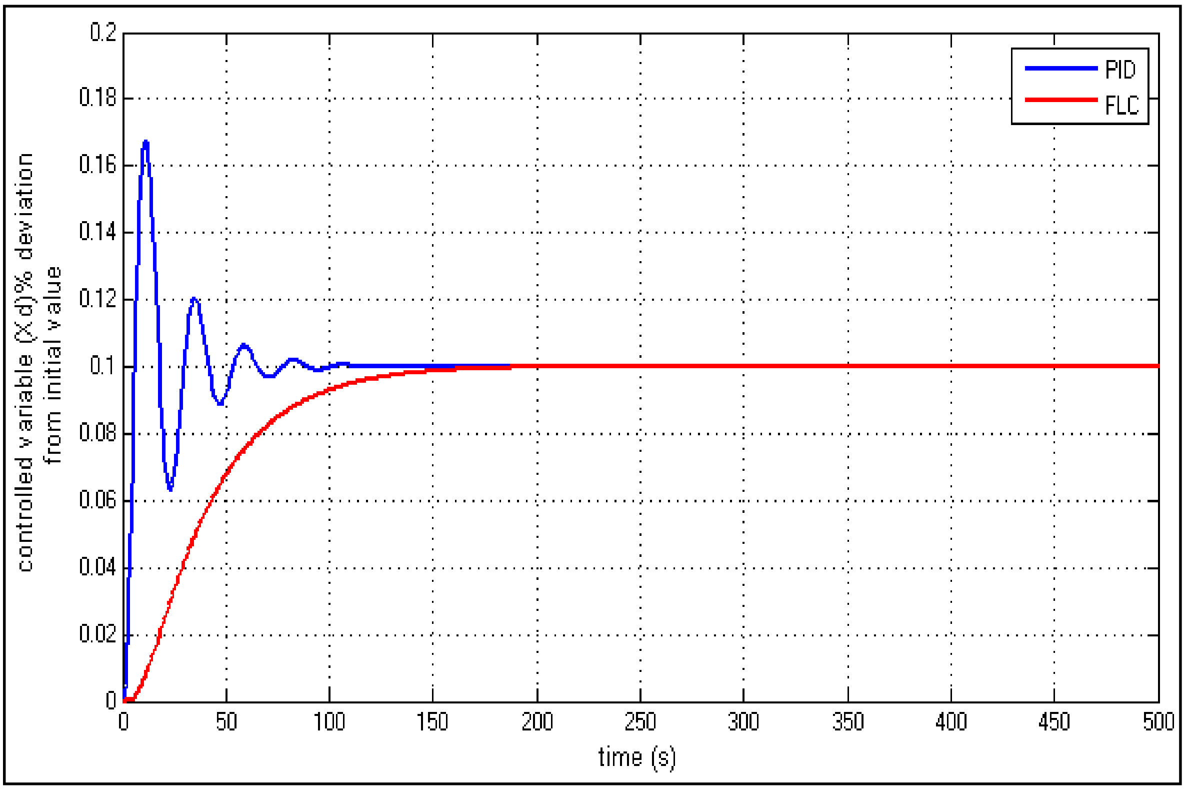


Figure 4.25 Transient response of distillate composition with respect to reflux flow rate in fuzzy logic and PID controllers

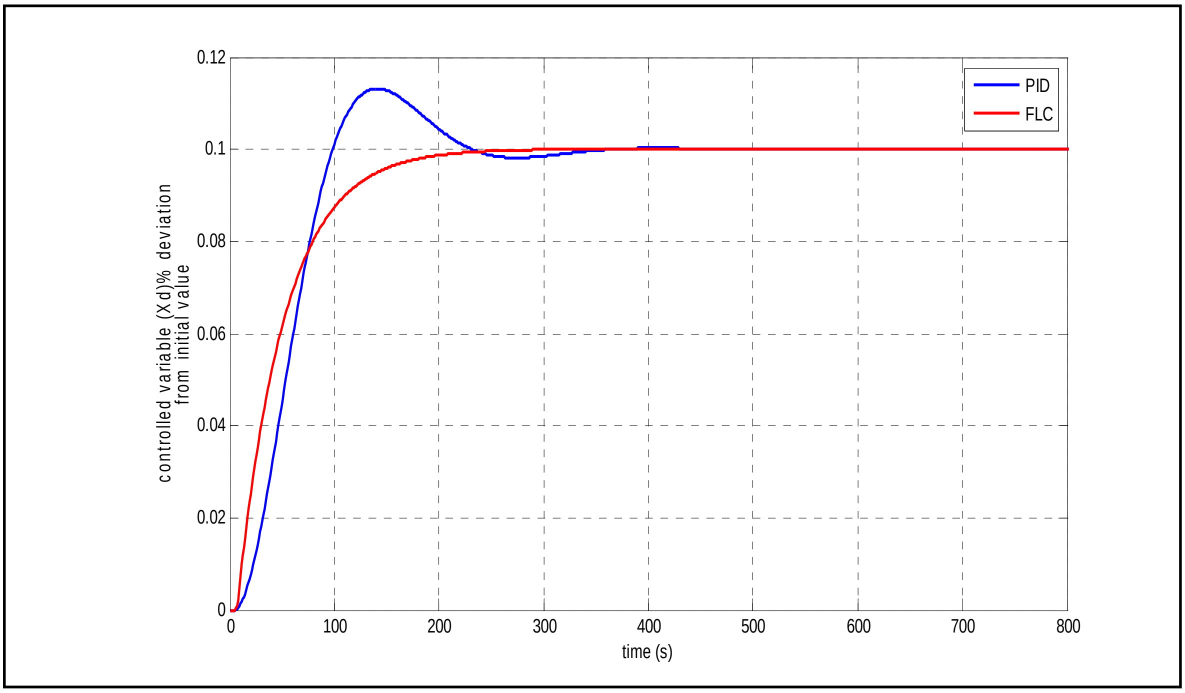


Figure 4.26 Transient response of distillate composition with respect to side stream flow rate in fuzzy logic and PID controllers

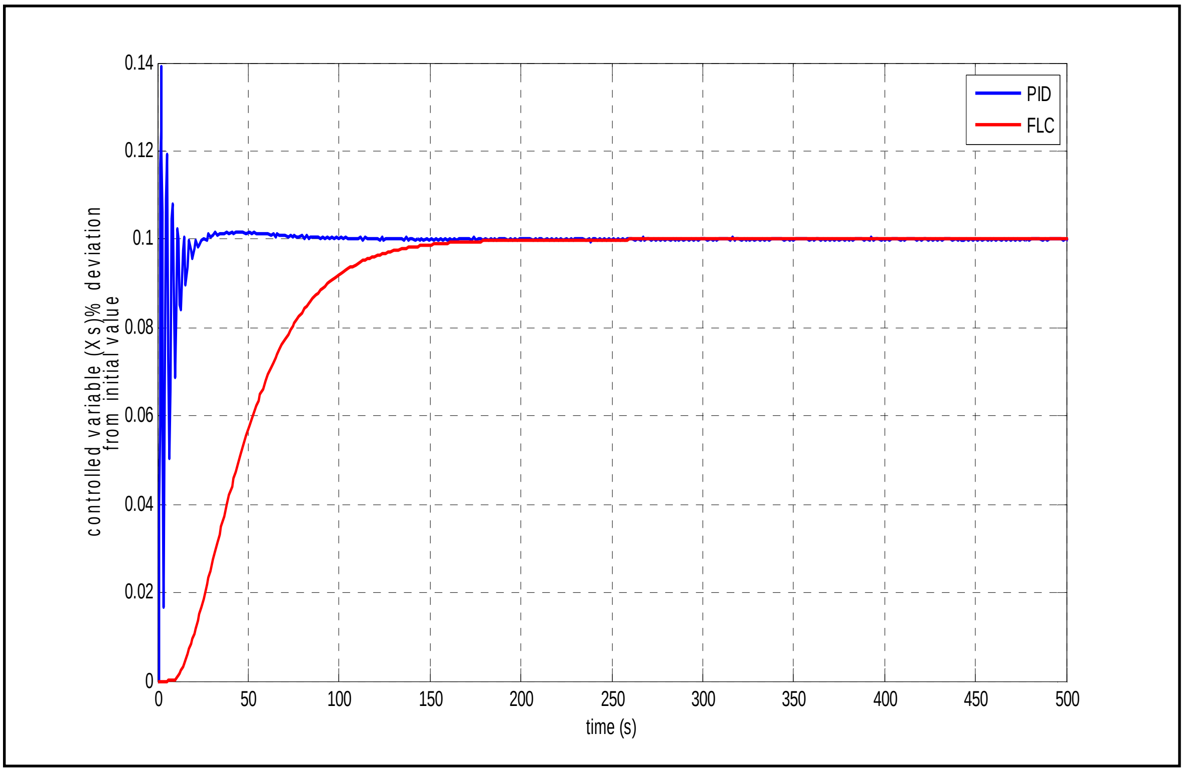


Figure 4.27 Transient response of side stream composition with respect to reflux flow rate in fuzzy logic and PID controllers

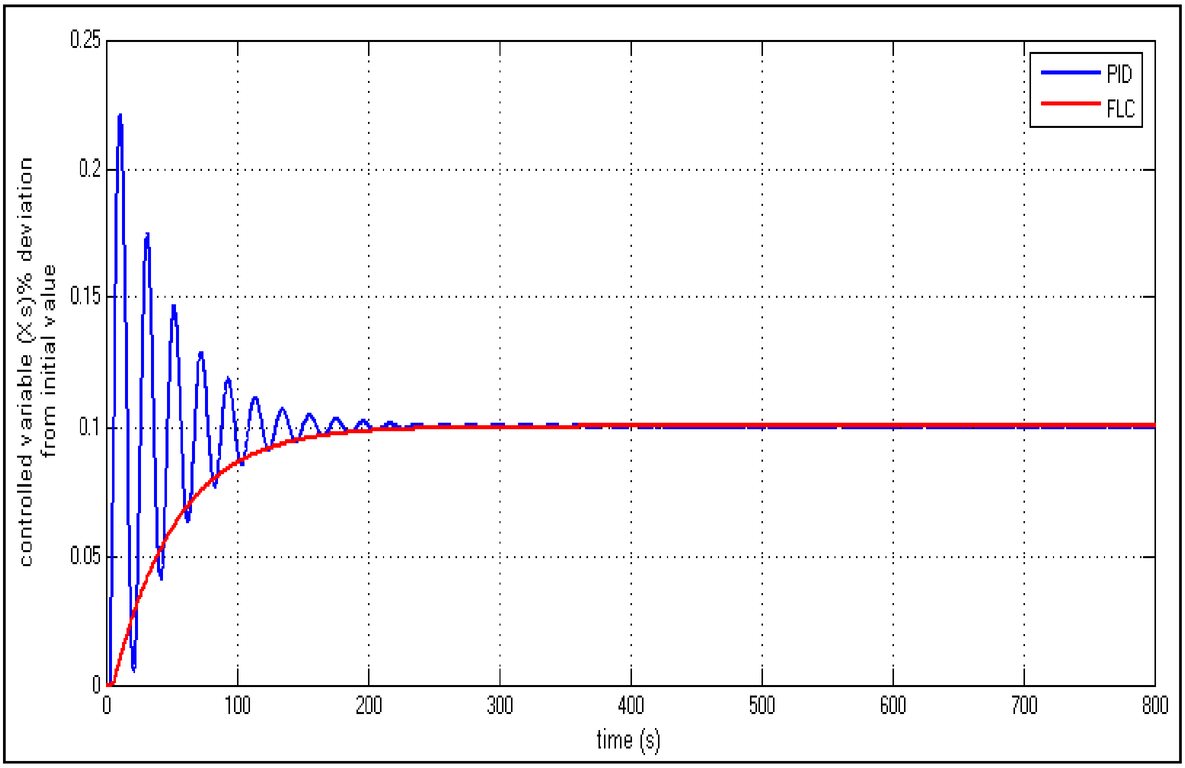


Figure 4.28 Transient response of side stream composition with respect to side stream flow rate in fuzzy logic and PID controllers

Chapter Five

Conclusions and Recommendations for Future Work

5.1 Conclusions

1. The response of distillation column with open loops is stable because of the overall individual loops are stable, if we have one closed loop of distillation column (top or bottom loop) with or without controllers then the system will be stable, and with both closed loops with or without controllers it will also be stable.

2. Interaction between the two loops of distillation column was recommended, so we use ideal decoupling in order to eliminate the interaction between loops and to make the system stable.
3. The decouplers are greatly improved the response of the system.
4. Fuzzy logic controller gave a marked improvement over Feedback controller. However the Fuzzy logic controller is preferable since it does not require an accurate mathematical model for the process to be controlled, while feedback control strategy requires very wide knowledge about the dynamic behavior and an accurate mathematical model of the process.
5. The system without time delay is more appropriate for stability of the system than with system having time delay

5.2 Recommendations for Future Work

The following suggestions for future work can be considered:

1. Applying the neural network to optimize the result of this work.
2. Adding other control strategies like adaptive, cascade control, etc.
3. The same procedure of this work is useful for another distillation column that is different in distillation specifications or using the same procedure for other controlled and manipulated variables.

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Appendix A

A.1 Fuzzy Logic Control Program

The following diagrams show the method of applying the Fuzzy logic control in the MATLAB program.

Figure A.1 shows the main window for Fuzzy logic control and the main data must put in it such as the error, change of error and the action.

Figure A.2 shows the window that concern with the error, this window displays the range of error and the range of NB, NS, Z, PS and PB.

Figure A.3 shows the window that concern with the change of error, this window displays the range of change of error and the range of NB, NS, Z, PS and PB of it.

Figure A.4 shows the window that concern with the action, this window displays the range of action and the range of NB, NS, Z, PS and PB of it.

Figure A.5a, b, c shows the window that concern with the 25 rules.

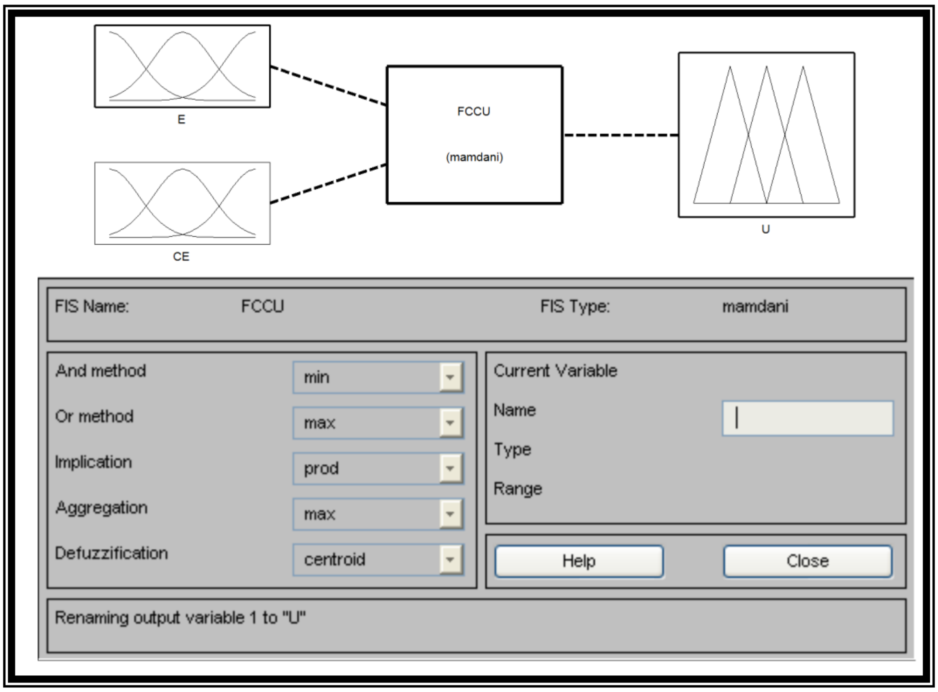

Figure A.1 Main Diagram of Fuzzy Logic Controller

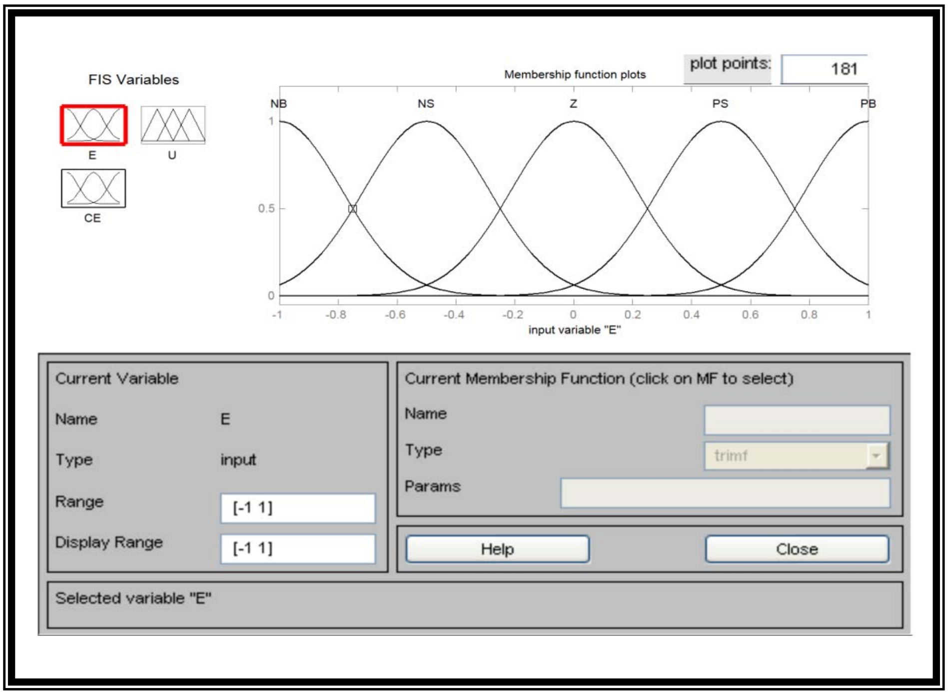

Figure A.2 Diagram of Error

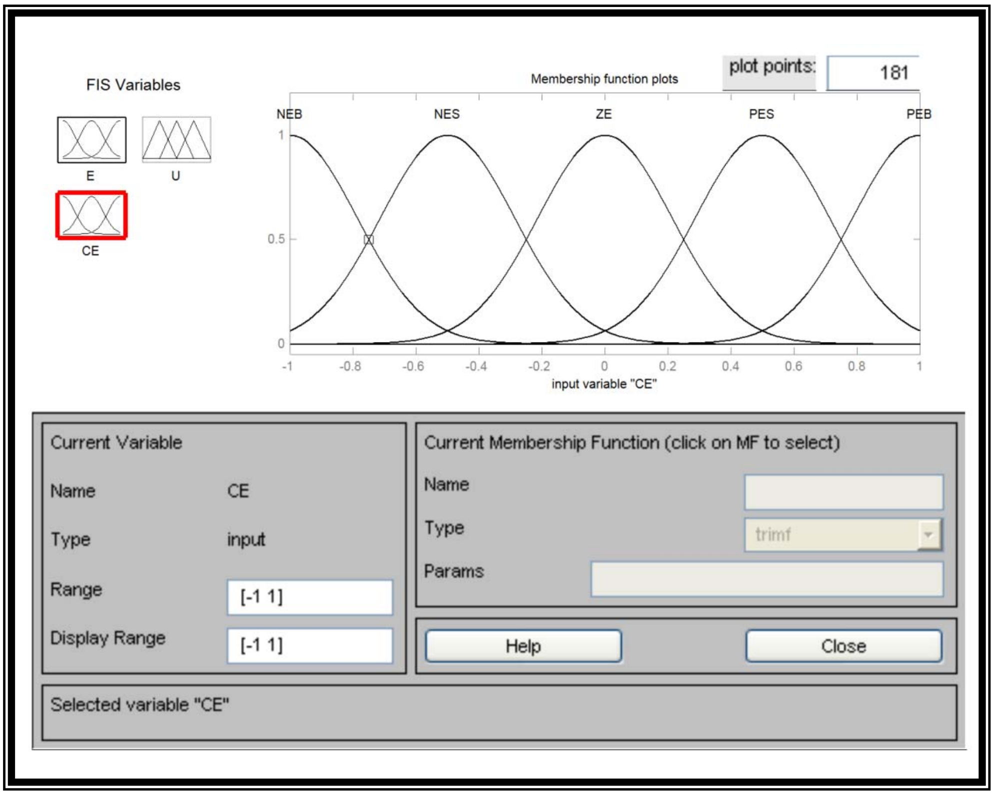

Figure A.3 Diagram of Change of Error

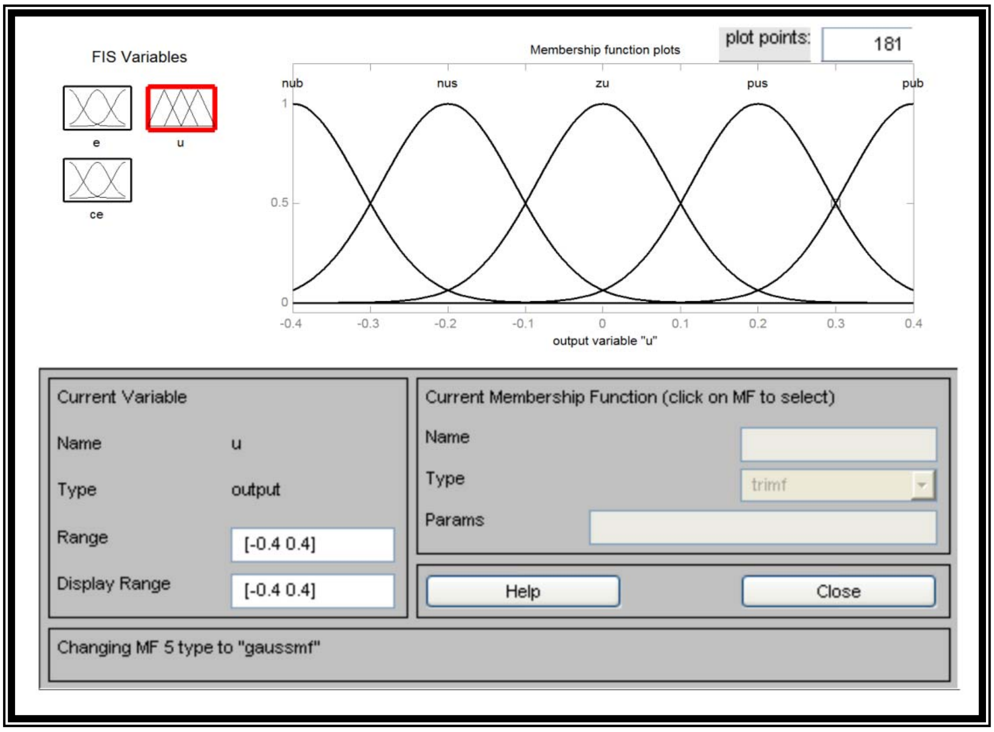

Figure A.4 Diagram of Action

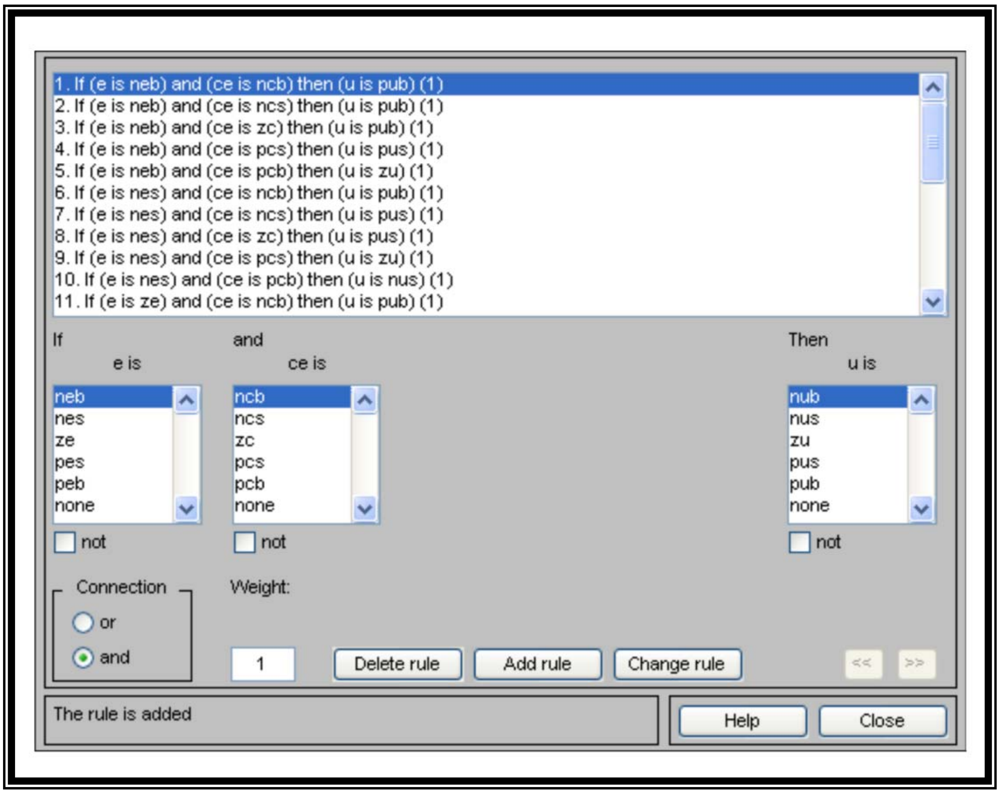

Figure A.5a diagram of Mamdani

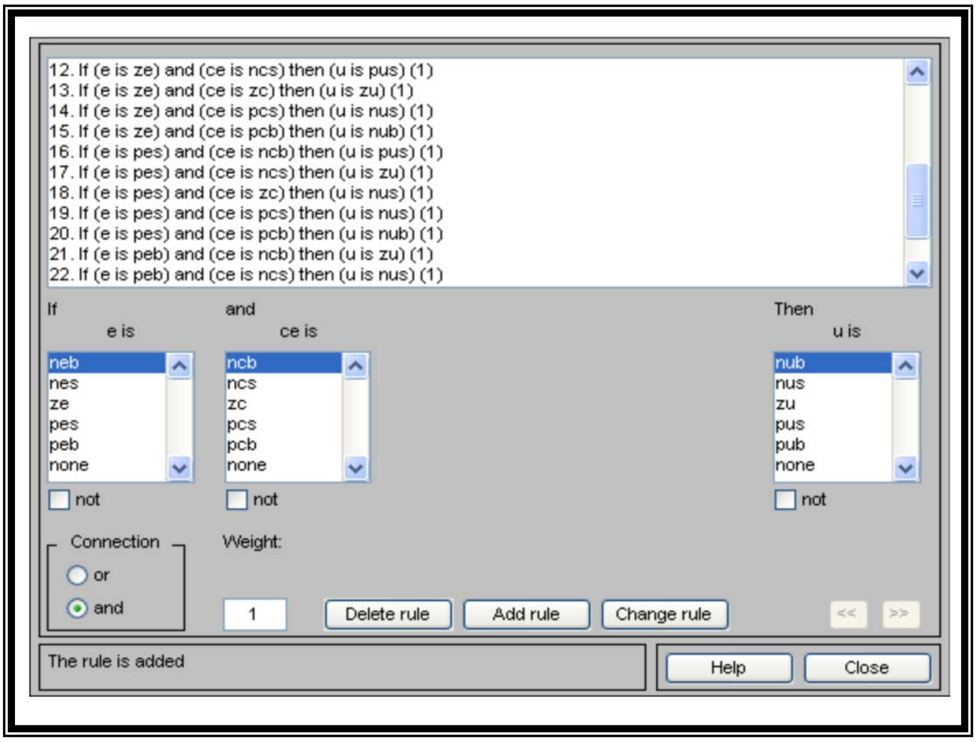

Figure A.5b Diagram of Mamdani

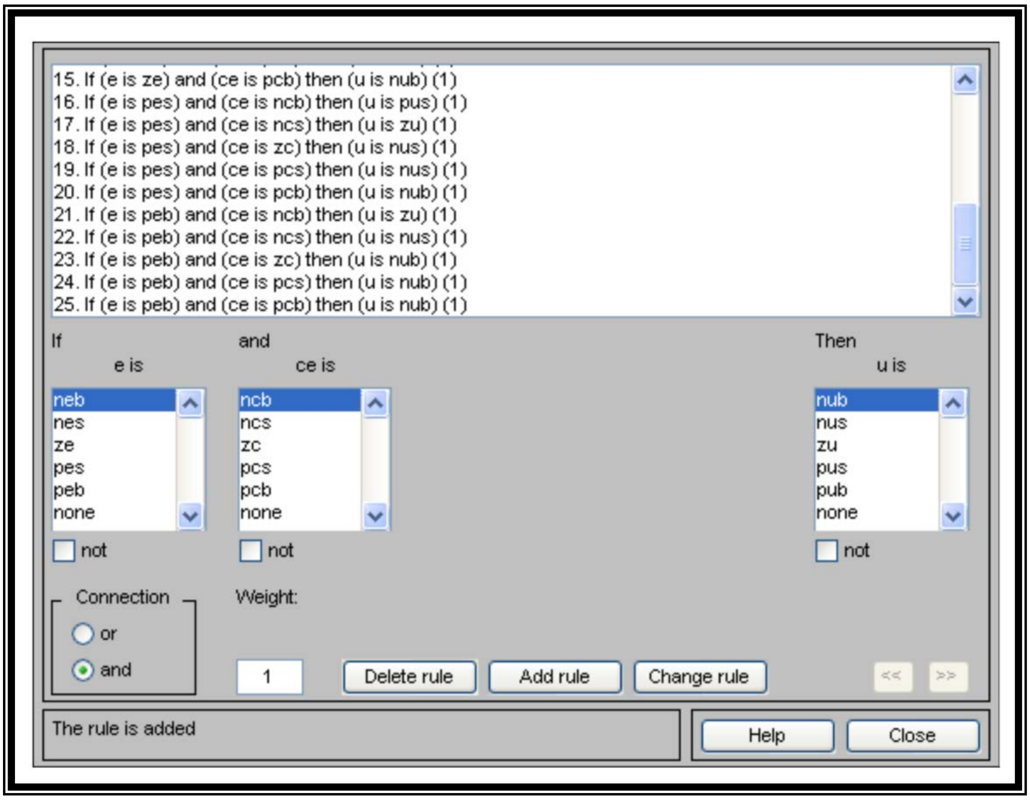

Figure A.5c Diagram of Mamdani

الخلاصة

برج التقطير هو من اهم متعلقات عمليات الفصل في الصناعات الكيمياوية و البتروكيمياوية لذلك نحتاج الى السيطرة بواسطة عمليات التصميم وقد تم تطوير نظام السيطرة على برج التقطير متعدد المكونات في هذا البحث.

في نظام السيطرة هذا اختير تركيب المادة المقطرة والجدول الجانبي (X_D, X_S) كمتغيرت مقاسة واعتبر كل من نسبة التدفق الجزرية ونسبة التدفق في الجدول الجانبي (R, L_S) متغيرات معالجة.

تم توصيف مؤشرات السيطرة بثلاث طرق وهي: (Zeigler-Nicolas Method ، Tyreus-Luyben methodو Process reaction curve) لايجاد افضل قيم للمعاملات (K_C و τ_i و τ_D).

تم تطبيق صيغ P ، PI و PID في استراتيجية السيطرة للنظام.

حددت درجة التداخل بالاعتماد على مصفوفة الكسب النسبي**(RGA)** مما يتطلب الى نظام **Decoupling**.

تم تصميم نظام **Decoupling** لالغاء تاثير التداخل في دوائر السيطرة.

و قد تم استخدام نظام السيطرة المنطقي كاستراتيجية اخرى للمقارنة مع نظام السيطرة التقليدي.

طريقة الاستجابة الترددية (مخطط بود) استخدمت في هذا البحث لدراسة اتزان النظام.

وقد استخدم مخطط بود (طريقة الاستجابة الترددية) في هذا البحث لدراسة اتزان النظام

استخدم برنامج **MATLAB** كأداة في الحل لجميع الحالات المستخدمة في هذا البحث.

شكر وتقدير

اود أن أعبر عن خالص شكري وتقديري وأمتناني العميق للمشرف الفاضل الدكتور خالد مخلف موسى لما قدمه لي من توجهات قيمة ونصائح سديدة طوال فترة أنجاز البحث.

واود التقدم بالشكر الى رئيس القسم المحترم وكافة منتسبي القسم ممن كان له الفضل في اتمام البحث.

كما اوجه شكري أيضاً الى زملائي طلبة الدراسات العليا وأصدقائي جميعاً لما ابدوه من مساعدة خلال فترة البحث.

ولا أنسى أن أتقدم بالشكر والامتنان الى من لازمني طوال فترة البحث وخلال أصعب الظروف، الى أغلى الناس وأعز من في الوجود الى أبي وأمي وأخوتي وجميع أهلي فلهم جزيل الشكر والتقدير.

سامر عطا قاسم

المحاكاة على الديناميكية والسيطرة لبرج التقطير

رسالة

مقدمة الى كلية الهندسة في جامعة النهرين وهي جزء من متطلبات نيل درجة
ماجستير علوم في الهندسة الكيمياوية

من قبل

سامر عطا قاسم

(بكالوريوس علوم في الهندسة الكيمياوية ٢٠٠٦)

ذو الحجة ١٤٣٠

كانون الاول ٢٠٠٩