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Faculty of Engineering- Department of Chemical Engineering

Simulation, Modeling and Control for Crude Distillation Unit

المحاكاة والنمذجة والتحكم في وحدة تقطير النفط الخام

A Thesis

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By

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Verse

قال تعالى:

((إقرأ بإسم ربك الذي خلق (1) خلق الإنسان من علق (2)

اقرأ وربك الأكرم (3) الذي علم بالقلم (4)

علم الإنسان ما لم يعلم(5)))

صدق الله العظيم

سورة العلق

Dedication

Especially to my: mother, father, husband, babies, sisters,

brother and friends

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Thank Allah for his kindness provided me with good health and full strength to finish this research, in time of difficulties. I believe He is always there to help me and give me an internal encouragement.

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Abstract

The control of the distillation unit in Khartoum refinery, by keeping the operating conditions in certain limits is essential to adjust disturbances which lead to deviation from the desired values. The aim of this work is to select the suitable controller and the appropriate tuning method. The control loops are designed depending upon controllability and performance. These loops were Control of furnace temperature, Control of interface level at reflux drum, Control of level at stripper one, Control of level at bottoms of the column and Control of pressure at the top of the column. The mathematical model for each loop was determined using MATLAB toolbox system identification by designing Graphical User Interface (GUI). The same was used to identify the transfer function for each loop. Having identified the transfer functions, each loop was closed and the characteristic equation was obtained from the overall transfer function. Using Z-N tuning method to get the ultimate

gain and ultimate period, from which the adjustable parameters were determined using Z-N criterion. These were used to investigate the offset upon a unit step change in the set point using proportional only (P), proportional integral (PI) and proportional integral derivative controller (PID).The controller that gave the minimum offset is found to be proportional integral derivative controller (PID) and has been selected. The procedure is repeated using Root-Locus and decay ratio criterions. In each case the controller that gave the lowest offset was selected. The three methods of Z-N, Roots Locus and decay ratio were compared according to their offset and found to be within good accuracy. Stability performance was examined using Bode, Nyquist and Root locus Criteria. Having closed the loops in conventional system, the same was transformed to digital computer control. The analysis of the offset investigation and system stability were performed on (Z-domains). The results are in agreement with conventional analysis. In conclusion a complete control system of a distillation unit was designed, drawn in block diagram and interfaced to a digital computer control.

المستخلص

يعتبر التحكم في وحدة التقطير لمصفاة الخرطوم ضروري وذلك بالتحكم في ظروف التشغيل عند قيم معينة لتجنب التشويشات.الهدف من البحث هو اختيار الحاكمية الأفضل للنظام.تم اقتراح دوال الانتقال بناءاً على تميز أداء المنظومة على النحو التالي: دالة درجة الحرارة عند الفرن وهى عباره عن درجة حرارة الخام الذى سيتم تقطيره،المنسوب عند أسفل برج التقطير، المنسوب عند المجرد الأول والضغط عند أعلى برج التقطير. كما تم بناء النموذج الرياضي لكل عملية باستخدام برنامج MATLAB للتنبوء بدالة الانتقال لكل دائرة تحكم وبعد تحديد دالة الانتقال تم الحصول على المعادلة المميزة من الدالة الكلية وباستخدام طريقة (Z-N) Zeighler-Nycolas ثم تم أيجاد المكسب الحرج وزمن الدورة الحرج واستخدمت هذه البيانات لفحص الانحراف بناءاً على إدخال خطوة أحادية في القيمة المطلوبة باستخدام الحاكمات ,Proportional integral (PI) ,proportional (P) و proportional integral derivative (PID). أجريت تحاليل للحاكمات و تم اختيار الحاكمة Proportional integral derivative التي أعطت انحرافاً يعادل الصفر. استخدمت طريقة Root Locus وطريقة decay ratio لإجراء نفس التحاليل وبعد المقارنة وجد أن الطرق الثلاثة أعطت نتائج متفاوتة نسبياً . وتم إختبار إستقراريه للدوائر التى تم التحكم فيها بواسطه الطرق الآتيه Bode, Nyquist and Root locus Criteria

وأجريت نفس التحاليل باستخدام الحاسوب لفحص قيمة الانحراف وكذلك فحص استقرار النظام في مجالZ المتقطع (Discrete). النتائج المتحصل عليها متوافقة مع التحليل التقليدي.وأخيراً تم تصميم نظام تحكم متكامل لوحدة التقطير كما تم رسم الصناديق لدوائر التحكم وحول نفس النظام الى مجال Z المتقطع وتم توصيله الى الحاسب الرقمي.

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Nomenclature

P proportional.

I integral.

D derivative.

k_c proportional gain.

τ_i Integral time.

τ_d Derivative time.

API American Petroleum Institute.

e(t) error.

Z-N Ziegler Nycolas.

Ku Ultimate gain.

Rlocus Root locus.

S.P. Set Point.

Wco Cross over frequency

Pu Ultimate Period.

ADC analog to digital converter.

DAC digital to analog converter.

G(s) transfer function.

Ʈ time constsnt.

ζ decay ratio.

T Temperature. ^{0}C

P pressure, MPa.

L Level, m.

F flow rate, ton/hr.

TT temperature transmitter.

TC temperature controller.

LT level transmitter.

LC level controller.

PT pressure transmitter.

PC pressure controller.

Abbreviation

KRC Khartoum refinery.

CDU Crude oil Distillation Unit.

GUI   Graphical User Interphase.

CHAPTER ONE

INTRODUCTION

Chapter one

Introduction

Petroleum is a complex mixture of organic liquids called crude, which occurs naturally in the ground and was formed millions of years ago. Crude oil varies from oilfield to oilfield in color and composition, from a pale yellow low viscosity liquid to heavy black 'treacle' consistencies.

As crude oil comes from the well it contains a mixture of hydrocarbon compounds and relatively small quantities of other materials such as oxygen, nitrogen, sulphur, salt and water. In the refinery, most of these non - hydrocarbon substances are removed and the oil is broken down into its various components, and blended into useful products.

An oil refinery is an organized and coordinated arrangement of manufacturing processes designed to produce physical and chemical changes in crude oil to convert it into everyday products like petrol, diesel, lubricating oil, fuel oil and bitumen. [10]

1 .1 Topping plant unit:

Topping plant unit is one of the most important units in petroleum refinery operation.

Topping plant process is divided into:

1- Crude oil blending unit:

The process of the topping plant starts from the crude oil blending unit.

In this unit, a blending occurs between different crude oils coming from various fields; in order to obtain physical properties required for refinery processing.

Although blending process is quite simple, yet it needs to be tightly and automatically controlled. The blending unit is very vital as it eliminates any offset and disturbances in the topping unit.

***2-* Desalting unit:**

After blending and having a crude oil mixture of constant physical properties, the blend is to be desalted in a desalter. Desalting of crude oil is a treatment process made to remove salts from crude oil.

This is an essential unit where in the salt, water and demulsifying agent needs to be automatically controlled.

3-Distillation unit:

The desalted crude blend will then be introduced to a fractionation column for separation of top products, side streams and bottom products or residue. These are mainly LPG, light naphtha, heavy naphtha, kerosene and residue.

Control is necessary for the Topping plant process and refinery operation.

1.2 Importance of process control

1.2.1 Overview

In recent years the performance requirements for process plants have become increasingly difficult to satisfy. Stronger competition, tougher environmental and safety regulations, and rapidly changing economic conditions have been key factors in tightening product quality specifications. A further complication is that modern plants have become more difficult to operate because of trend towards complex integrated processes.

Process control has become increasingly important in the process industries as a consequence of global competition, rapidly changing economic conditions, and more stringent environmental and safety regulations. Process control is also a critical concern in the development of more flexible and more complex processes for manufacturing high value added products. Tuning control is one of the complex and difficult in process control. Control tuning is the major key issue to operate the plant. Process tuning is a key role in ensuring that the plant performance satisfies the operating objectives.

1.2.2 External disturbances

One of the most common objectives of a control in chemical plants to suppress the influence of external disturbances. The disturbances, which denote the effect of the surrounding on the processes of which are usually out of the reach of human operator.

The control mechanism is needed to make the proper changes on the process to cancel the negative impact of the disturbances.

1.2.3 Process sa E4 R4TGHBJtability

If, a process variable; such as temperature, pressure, concentration or flow rate disturbed by external factors returns to its initial value; it is called: "self regulation process"; which needs no external interventions for stabilization.

If the process variable does not return to the initial value after disturbed by external influences, it is called: "unstable process". This requires a control for the stabilization of system behavior.

1.2.4 Performance Optimization

The plant operation should be able to change in such a way that an economic objective – profit is always maximized.

Control is necessary for the topping plant unit in refinery operation. The main objectives of applying Control on processes are: maintaining safety (operation conditions to be within allowable limits), production specifications (final products to be in the right amounts), compliance with environment, operational condition (some equipment's have constrains to be adhered with through its operation in a plant) and economic consideration (plant operation must conform with the market conditions: availability of raw materials, demand of products and utilization of energy, capital and human labor).

1.3 Statement of the research problems:

The aims of this study are

1- To investigate the impact of the decrease in the flow rate of crude oil on control systems performance

2- Apply advance process control, utilizing technology varying from simple multivariable control to the most advance control system.

3- Design of these units is one of objective of this proposal.

1.4 Control systems

In this study, the control systems of a crude oil topping plant are:

1- Control of furnace temperature.
2- Control of interface level at reflux drum.
3- Control of level at stripper one.
4- Control of level at bottoms of the column.
5- Control of pressure at the top of the column.

1.5 Objectives

1.5.1 General Objectives

- Develop a mathematical model using graphical user inter-phase (GUI). In order to test the performance of distillation unit at Khartoum. Refinery
- Tuning and Analysis of the control performance and stability of the system.
- Design of a computer control system in continuous analog and discrete digital format.

1.5.2 Specific Objectives

- To determine the transfer function from model using system identification.

- To determine the optimum parameters; K_c, τ_I and τ_D.
- To tune the feedback controllers used in distillation unit.
- To compare the accuracy of Bode, Root Locus and Nyquist criterions.

CHAPTER TWO

LITERATURE REVIEW

Chapter two

Literature Review

Crude oil is extracted from the ground, on land or under the oceans, by sinking an oil well and is then transported by pipeline and/or ship to refineries where their components are processed into refined products. [10]

2.1 Importance of Refining

An oil refinery is an organized and coordinated arrangement of manufacturing processes designed to produce physical and chemical changes in crude oil to convert it into everyday products like gasoline, diesel, lubricating oil and fuel oil.

As crude oil comes from the well it contains a mixture of hydrocarbon compounds and relatively small quantities of other materials such as oxygen, nitrogen, sulphur, salt and water. In the refinery, most of these non - hydrocarbon substances are removed and the oil is broken down into its various components, and blended into useful products.

Every refinery begins with the separation of crude oil into different fractions by distillation.

The quantities of petrol available from distillation alone were insufficient to satisfy consumer demand. Refineries began to look for ways to produce more and better quality petrol. [10]

2.2 Crude oil Distillation Unit (CDU)

Crude oil Distillation unit (CDU) is one of the most important units in petroleum refinery operation.

CDU unit is divided into:

2.2.1 Crude oil blending unit

The process of the CDU starts from the crude oil blending unit.

In this unit, a blending occurs between different crude oils coming from various fields; in order to obtain physical properties required for refinery processing.

2.2.2 Crude oil desalting unit

After blending and having a crude oil mixture of constant physical properties, the blend is to be desalted in a desalter. Desalting of crude oil is a treatment process made to remove salts from crude oil.

There are three types of desalting methods:

1- Electrical desalting method.

2- Chemical desalting method.

3- Filtration desalting method. [14]

2.2.3 Crude oil Distillation Unit(topping column)

Because crude oil is a mixture of hydrocarbons with different boiling temperatures, it can be separated by distillation into groups of hydrocarbons that boil between two specified boiling points. Two types of distillation are performed: atmospheric and vacuum.

1- Atmospheric distillation: takes place in a distilling column at or near atmospheric pressure liquid fractions are drawn from the trays and removed. [1]

2- Vacuum distillation: recover additional heavy distillates from this residue; it may be piped to a second distillation column where the process is repeated under vacuum.

This allows heavy hydrocarbons with boiling points of 450°C and higher to be separated without partly cracking into unwanted products such as coke and gas. [1]

2.3 Introduction about distillation column:

Distillation columns are very widely used in the chemical and petroleum industries to separate chemical components into more or less pure product streams. This separation is based on differences in "volatilities" among various chemical components. In a distillation column, the more volatile, or lighter, components are removed from the top of the column, and the less volatile, or heavier, components are removed from the lower part of the column. [17]

The vast majority of industrial distillation columns are equipped with trays or plates (sometimes "decks" in the petroleum industry). These trays facilitate good contact between the phases. Vapor-liquid contacting is achieved by a variety of devices. [16]

The most widely used trays in recent years have been the sieve and valve trays because of their simplicity and low cost.

Sieve trays are simple flat plates with a large number of small holes. Vapor flows up through the holes, preventing the liquid from falling through. Liquid flows across each tray, passes over a weir, and drops into a "down comer", which provides liquid for the tray below through an opening at the base of the down comer. [17]

Distillation columns can be used to separate chemical components when there are differences in the concentrations of these components in the liquid and vapor phases.

2.3.1 Basic distillation equipment

Distillation columns are made up of several parts, each of which is used either to transfer heat energy or enhance material transfer. [19]

A typical distillation column contains several major components:

- A vertical shell where the separation of liquid components is carried out.
- Column internal such as trays or packing which are used to enhance component separations.
- A reboiler to provide the necessary vaporization of the liquid.
- A condenser to remove some of the energy to cool and condense the vapor leaving the top of the column.
- A reflux drum to hold the condensed vapor from the top of the column so that reflux can be recycled backed to the column.
- Side streams.
- Pump around.

Control is necessary for distillation column in CDU unit to avoid any problems and disturbances.

The main objectives of applying Control on processes are: maintaining safety (operation conditions to be within allowable limits), production specifications (final products to be in the right amounts), compliance with environment regulations, operational condition (some equipment's have constrains to be adhered with through its operation in a plant) and economic consideration (plant operation must conform with the market conditions: availability of raw materials, demand of products and utilization of energy, capital and human labor). [3]

2.4 Control Problems

Many control problems arise from the design of engineering systems. Such problems typically large- scale and fuzzy.

The generalized design problem reduces to the selection of a suitable controller, which could ensure the designed control law, and adjustments of the controller's parameters to suit the plant's required dynamic and static responses.

2.4.1 Design Aspects of a process control system

1- Classifications of the variables

The variables (temperature t, pressure p, concentration CA.) associated with a chemical plant, they are divided to:

a) Input variables, which denote the effect of the surroundings on the chemical process, they are divided to:

1- Manipulated (adjustable), if their values can be adjusted freely by the human operator or a control mechanism.

2-Disturbances, if their values aren't adjustable by an operator or control system.

b) Output variables, which denote the effect of the process on the surroundings, they are divided to:

1- Measured output variables if their values are known by directly measuring them.

2-Unmeasured output variables, if they are not or cannot are measured directly.

Figure (2.3) explain Classifications of the variables

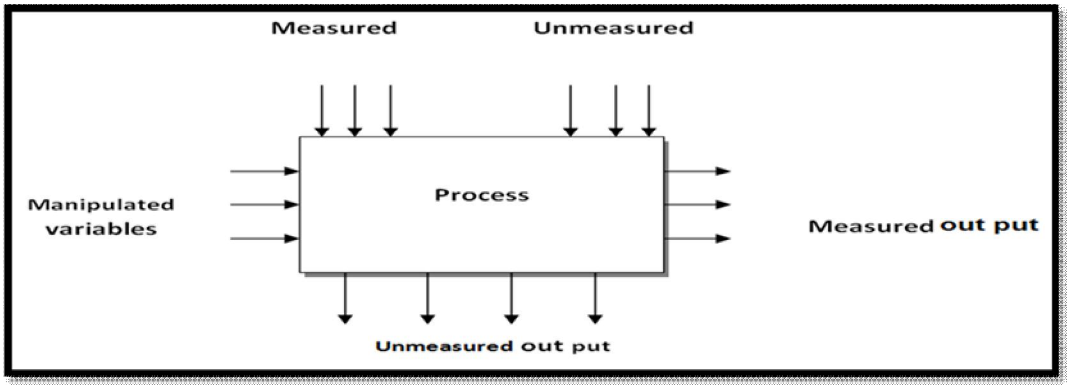


figure (2.3): Classifications of the variables

2.4.2 Elements of a control system

1. Define control objective, to design a control system that satisfies the chemical process, the operational objectives of the system should be determined.
2. Select measurement of the chemical process, certain variables should be measured (T, P, F….) which are representing the control objectives on the operation performance. This is done whenever it's possible. Such measures are calling primary measurements.

2.4.3 Select manipulated variables

In the process there are a number of input variables, which can be adjusted freely, some will select to use as manipulated variables by determination of degrees of freedom, as the choice will affect the quality of the control action.

2.4.4 Control configuration

Depending on how many controlled outputs in the plant .it can be distinguish as either single input single output (SISO) or multiple outputs (MIMO).

Feedback control configuration uses direct measurements of the controlled variables (measured outputs) to adjust the values of the manipulated variables-the objective is to keep the controlled variables at desired levels.

Feed forward control configuration uses direct measurement of the disturbances to adjust the values of the manipulated variables – the objective is to keep the values of the controlled outputs at desired level. [2]

2.4.5 Elements of a Control Loop

Usually a control loop consists of the following four basic elements:

1) The process or the plant itself. It may be simple (single stage) or compound (multi-stage).

2) The measuring or feedback element.

3) The controlling element or controller.

4) The regulating element, actuator or final control element.

In general, these four components constitute most control systems; however, more complex systems exist in which more components are used. e.g. some processes require a cascade control system in which two controllers and two measuring elements, or even more are used.

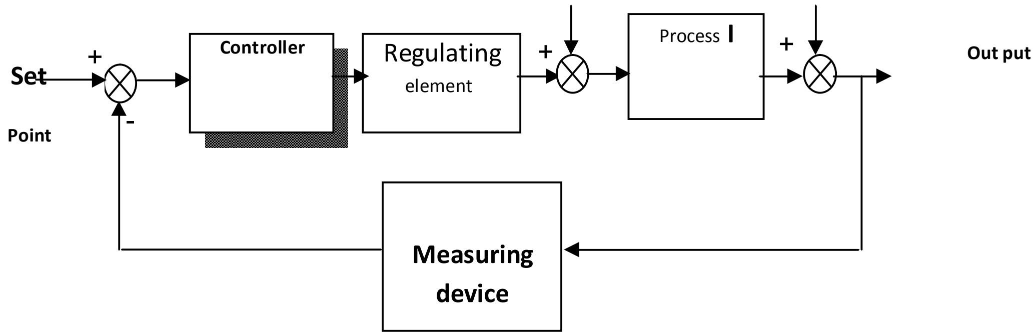


Figure (2.4): Feed Back Control System. [4]

2.5 Controllers

The controller is a device which monitors and affects the operational

conditions of a given dynamical system. The operational conditions are typically referred to as output variables of the system which can be affected by adjusting certain input variables. It receives the information from the measuring device and decides what action should be taken.

Controller tuning inevitably involves a tradeoff between performance and robustness. The performance goals of excellent set-point tracking and disturbance rejection should be balanced against the robustness goal of stable operation over a wide range of conditions. Before starting the tuning, it is general to make various reason and criteria for selecting which controller type will be adequate for which application. In control tuning, feedback control was used. Feedback control is that the controlled variables are measured and the measurement is used to adjust the manipulated variables and the disturbance variable is not measured. [4]

Controller is used to make tuning in process control. The selection made on the basis of the general characteristics of the different feedback controllers are the most practical.

Types of continuous controllers

2.5.1 Proportional controller (P- controller)

In this type of controller, the controller output (control action) is proportional to the error in the measured variable.

The error is defined as the difference between the current value (measured) and the desired value (set point). If the error is large, then the control action is large. [3]

Mathematically:

$C(t) \propto K_c * e(t) \rightarrow c(t) = K_c * e(t) + C_S$ ………… (2.1)

Where:

$e(t) \equiv$ the error

$K_c \equiv$ the controller's proportional gain

$C_S \equiv$ the steady state control action (necessary to maintain the variable at the steady state when there is no error)

$C(t) = K_c * e(t)$ ………………………………… (2.2)

The transfer function is

$$G(s) = c(s) / E(s) = k_c \quad \ldots\ldots\ldots\ldots\ldots \ldots\ldots \quad (2.3)$$

Proportion action repeats the input signal and produces a continuous action.

The gain K_c will be positive if an increase in the input variable requires an increase in the output variable (direct-acting control), and it will be negative if an increase in the input variable requires a decrease in the output variable (reverse-acting control).

Proportional action decreases the rising time making the response faster, but it causes instability by the overshooting (offset). [3]

2.5.2 Integral controller (I- controller)

Also, known as reset controller, this type of controller, the controller

output (control action) is proportional to the integral of the error in the measured variable. [3]

The transfer function is:

$$G(s) = c(s) \setminus E(s) = k_c (1 \setminus \tau_i s) \quad \ldots\ldots\ldots\ldots\ldots\ldots \quad (2.4)$$

Where:

K_c = the controller's proportional gain.

τ_i =integral time.

Integral action has a higher overshoot than the proportional due to its slowly starting behavior, but no steady state error (no offset). [3]

2.5.3 Derivative Controller (D-Controller)

In this type of controller, the controller output (control action) is proportional to the rate of change in error (error derivative) in the measured variable. [3]

$K_c \equiv$ the controller proportional gain

τ_d $\equiv$ derivative time constant in minutes.

The transfer function is:

$$G(s) = c(s) \backslash E(s) = k_c \tau_d s \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots (2.5)$$

2.5.4 PI- controller

In this type of controller, the controller output (control action) is proportional to the rate of change in error and the integral of the error in the measured variable.

Initially the controller output is the proportional action (integral contribution is zero) after a period of it mints the contribution of the integral action starts.

The integral action repeats the response of the proportional action causes the output to changing continuously as long as the error is existing.

Reset time is the time needed by the controller to repeat the initial proportional action change output. [3]

The transfer function:

$$G(s) = c(s) \backslash E(s) = k_c (1 + 1 \backslash \tau_i s) \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots \text{(2.6)}$$

2.5.5 PD- Controller

In this type of controller, the controller output (control action) is proportional to summation of the error and rate of change in error in the measured variable. [3]

The transfer function:

$$G(s) = c(s) \backslash E(s) = k_c (1 + \tau_d s) \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots \text{(2.7)}$$

Derivative time is the time taken by the proportional action to reproduce the initial step of the derivative action.

2.5.6 PID- controller

In this type of controller, the controller output (control action) is proportional to summation of the error and rate of change in error and the integral of the error in the measured variable.

The integral action will provide the automatic reset to eliminate offset following a load change. The derivative action will improve the system stability, which results in reducing the peak deviation as well as providing a faster recovery. Increasing the integral action will eliminate the offset in shorter time, but as it decrease the stability it leads to longer recovery time. Hence some compromise is necessary between the rate of recovery and the offset and the overall recovery time. [4]

$$G(s) = c(s) \backslash E(s) = k_c (1 + 1 \backslash \tau_i s + \tau_d s) \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots \text{(2.8)}$$

2.6 Controller tuning

The controller is the active element that is receives the information from the measurements and takes appropriate control actions to adjust the value of the manipulated variables taking the best response of the process using different controller laws.

In order to be able to use a controller, it must first be tuned to the system. This tuning synchronizes the controller with the controlled variable, thus allowing the process to be kept at its desired operating condition.

2.6.1 Ziegler-Nicholas (Z-N) tuning method

Ziegler-Nicholas is the one of tuning method techniques. It goes through the following steps:

Step1: Set up the system with proportional control only, i.e. set (τ_d) at its minimum value and (τ_i) at its maximum value.

Step 2: make a set point step test and observe the response.

Step 3: Evaluate the period of the constant oscillation; this period is called the ultimate period P_u.

Step 4: Calculate the parameters according to the following formulas:

Table 2.1 Ziegler-Nicholas Closed Loop Relevant Controller

Parameters. [2]

Controller type	Gain K_c	Integral time τ_i	Derivative time τ_d
P	0.5 k_u	-	-
PI	0.45 K_u	$P_u/1.2$	-
PID	0.6 K_u	$P_u/2$	$P_u/8$

2.7 Stability

Stability is the state or quality of being stable.

In mathematics, a condition in which slight disturbances in a system does not produce a significant disrupting in effect on that system. The transient response of a feedback control system is a primary interest and must be investigated. A very important characteristic of transient performance of a system is its stability. A control system is satisfactory it should meet the following requirements:

1- Stability by showing stable behavior.

2-Good control quality or good dynamics which deals with transient response. [2]

2.8.1 Concept of stability

A stable system is one that will remain at rest unless excited by an external source and will return to rest if all excitations are removed.

Dynamically the system stability is determined by its response to inputs or disturbance. A system is considered stable if its response is bounded, or if every bounded input produces a bounded output.

If some excitation causes the outputs to increase continuously or to oscillate with growing amplitude, it is unstable system.

2.8.2 Stability techniques

The system stability can be tested by considering its response to a finite input signal. Several methods have been developed to deduce the system stability from its characteristic equation.

The direct method criterion are short cut methods for assessment the stability of a system by providing information from the S-domain without finding out the actual response of a system in the time domain.

Methods techniques for stability

2.8.3 Routh-Hurwitz criteria

Routh-Hurwitz criteria are an algebraic procedure for determining whether a polynomial has any zeros in the right half plane. It involves examining the signs and magnitudes of the coefficients of the characteristic equation without actually having to determine the roots. It does not indicate the relative degree of stability or instability. The characteristic equation of an nth order system is:

$P_{(S)} = a_0 s^n + a_1 s^{n-1} + \ldots a_{n-1} s + a_n = 0$ ……………………….. (2.9)

If a_0 is negative, multiply both sides of the equation above by -1.

- If any of the coefficient a_0, a_2,………$a_{n-1} a_n$ is negative, there at least one root of the characteristic equation has a positive real part, and the corresponding system value is unstable.
- If the all coefficients are positive we use the array below

Table 2.2 Routh-Hurwitz coefficients

Row 1	a_0	a_2	a_4…
Row 2	A_1	A_3	A_5…
Row 3	A_1	A_2	A_3…
Row 4	B_1	B_2	B_3
……	……	……	…..
Row n+1	W_1	W_2	….

Where:

$A_1 = (a_1.a_2 - a_0.a_3)/a_1$ ……………………….. (2.200)

$A2 = (a_1.a_4 - a_0.a_5)/\ a_1$ ………………………… (2.21)

$B1 = (A_1.a_3\text{-}a_1.A_2)\ A_1$ ………………………… (2.22)

$B2 = A_1.a_5 - a_1.A_3)/\ A_1$… ………………………… (2.23)

Looking at the first column of the array a_0 a_1A_1 B_1 C_1………..w1

a) If any of these elements is negative, there is at least one root to the right of the imaginary axis and the system is unstable.
b) The number of sign changes in the first column is equal to the number
c) Of roots lays at the right side of the imaginary axis.

Therefore, a system is stable if the first column's elements at the array are positive. [10]

The graphical method

To investigate the behavior of roots of the characteristic equation, typically, graphical methods are good enough for evaluating process stability. There are several techniques graphical useful in the stability analysis.

2.8.4 The root locus

A root loci plot is simply a plot of the s zero values and the s poles on a graph with real and imaginary ordinates. The root locus is a curve of the location of the poles of a transfer function at the gain Kc is varied from zero to infinity. This method is very powerful graphical technique for investigating the effects of the variation of a system parameter on the locations of the closed loop poles. A system is stable if all of its poles are the left-hand side of the s-plane for any value of Kc (for continuous systems) or inside the unit circle of the z-plane (for discrete systems). [2]

The root locus of a system provides information about the stability of a close loop system and also informs about its general dynamic response characteristics as Kc changes, the general graph gives a conclusion about the dynamic behavior of the system. [3]

2.8.5 Nyquist plot

Nyquist plot are used to analyze system properties including gain margin, phase margin, stability, and for assessing the stability of a system with feedback.

It is represented by a graph in polar coordinates in which the gain and phase of frequency response are plotted, the gain is plotted as the radius vector and the phase shift is plotted in degrees clockwise from the right hand abscissa. There is a separate point for each frequency; the frequencies are indicated next to a few of the point, or an arrow is used to show the direction of increasing frequency. [2]

2.8.6 Bode plot

A graphical method consists of plotting two curves to present the response of open loop at variance frequencies. It is used to determine the stability limits of the closed loop system, the critical frequency of the system, and predict the optimum controller setting. The response presented at a log-log plot for the amplitude ratios, accompanied by the Simi log plot for the phase angle. [2]

2.9 Types of control systems:

There are several types of control systems:

2.9.1 Feed forward control system:

Feed forward control is strategy used to compensate for disturbances in a system before they affect the control variable. A feed forward control systems measures a disturbance variable, predict its effect on the process, and applied corrective action. Show figure 2.5.

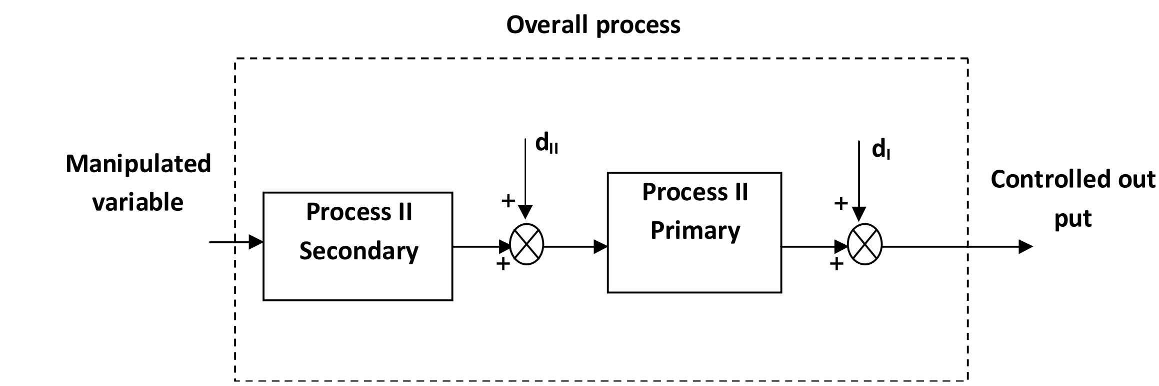


Figure 2.5 open loop control system.

2.9.2 Feedback control system

The feedback control system measures the value of the output using the measuring device, which sends the signal through the transmitter to the controller. The controller compares this value with the desired value (set point) and supplies the deviation signal to the final control element, which in turn changes the value of the manipulated variable. Show figure 2.6.

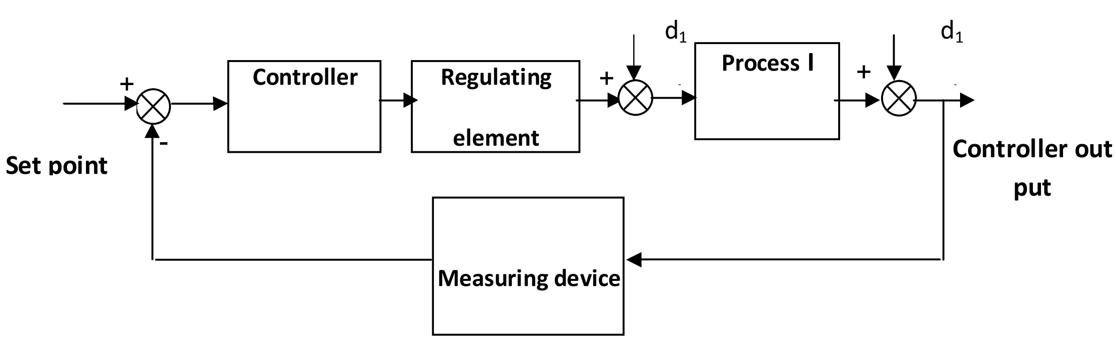


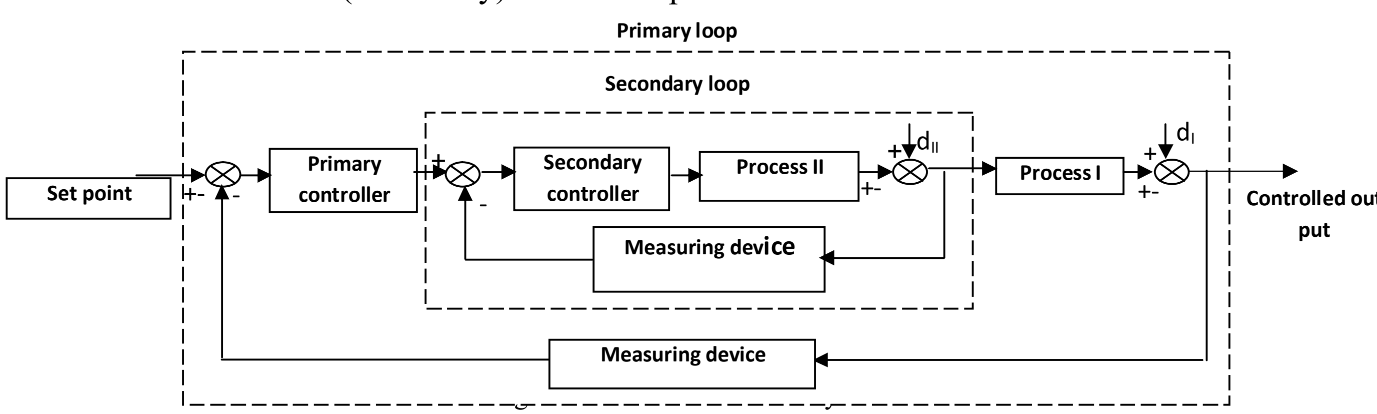
Figure 2.6: Feed Back Control System

2.8.3 Cascade control system:

A cascade control system is a multiple-loop system where the primary variable is controlled by adjusting the set point of a related secondary variable controller. The secondary variable then affects the primary variable through the process.

Cascade control is widely used within the process industries. Conventional cascade scheme have two distinct features:

- There are two nested feedback control loops. There is a secondary control loop located inside a primary a primary control loop.
- The primary loop controller is used to calculate the set point for the inner (secondary) control loop.



There are other control systems with multiple loops than cascade control system:

- Override control system.
- Adaptive control system.
- Ratio control system.
- Inferential control system. [13]

2.10 Computer – aided process control

The earliest proposal to use computer for process control application was made by Brown and Campbell in their paper, in 1950. [8]

Computers have now found extensive use in process and manufacturing industries.

It has brought not only new possibilities but also new challenges to control engineers.

After technological development of digital computer system its use for process control application has tremendously creased.

The basic objective of computer aided process control is to identify the information flow and to manipulate the material and energy flow of given process in a desired optimal way.

Elements of computer- aided process control system:

Basic functions of computer –aided process control system are:

1. Measurements and data acquisition.

2. Data conversion with scaling and checking.
3. Data accumulation and formatting.
4. Visual display.
5. Comparing with limits and alarm raising
6. Recording and monitoring of events, sequence and trends.
7. Data logging and computation.
8. Control actions. [7]

A block diagram of computer aided process control system is shown in figure (2.9)

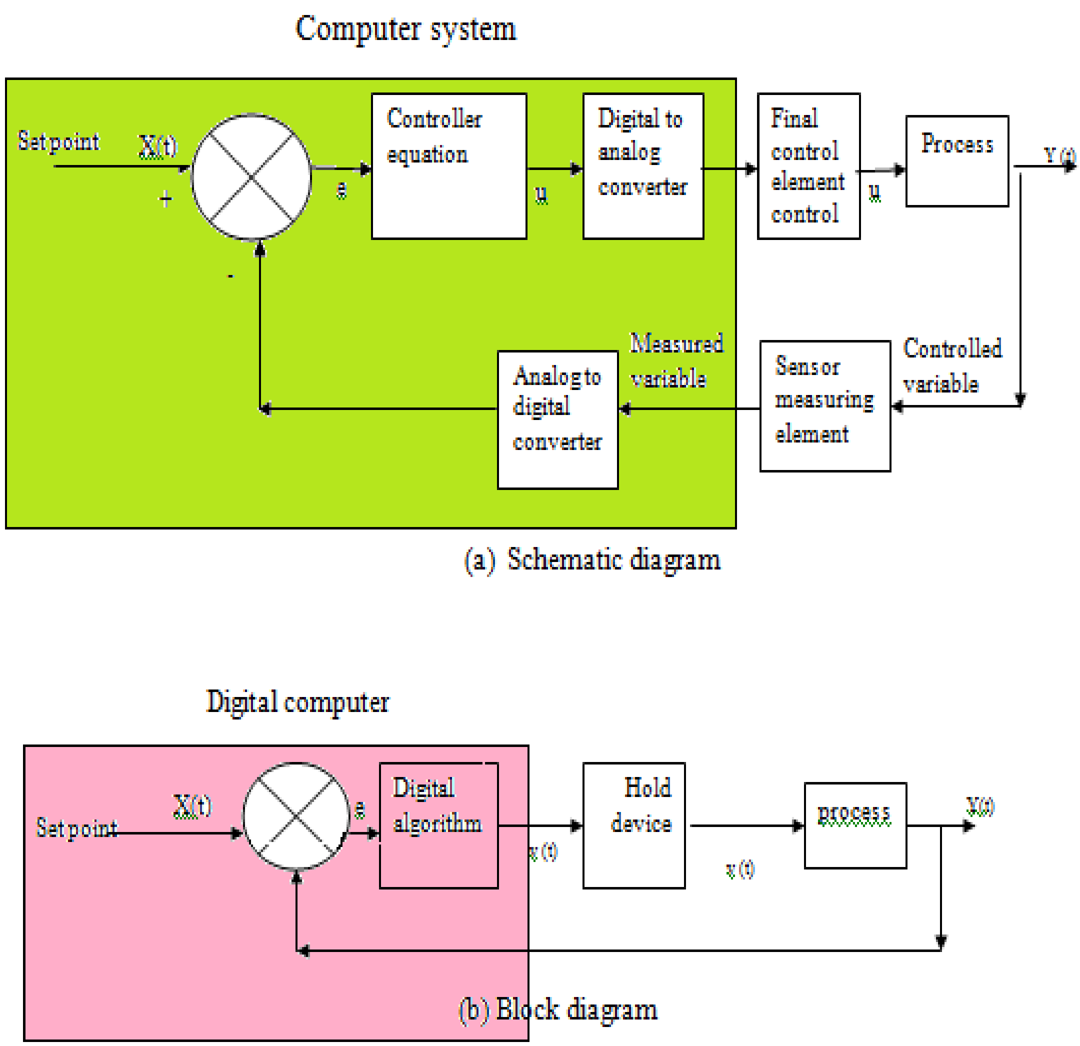


Figure (2.8) A block diagram of computer aided process control system in

As shown in figure (2.9) (a), the controlled variable (output of the process) is measured as before in continuous electrical signal (analog) form, and converted into a discrete – time signal using an analog – to – digital converter (ADC). The value of the discrete signal thus produced is then compared with the discrete of the set- point (desired value) inside the digital computer to produce an error signal (e) an executed with yields a discrete controller output. The discrete signal is then converted into a continuous electrical using a device called digital – to- analog converter (DAC) and the signal is then fed to the final control element. This control strategy is repeated at some predetermined frequency so as to achieve the closed – loop computer control of the process. [7]

2.10.1 Classification of computer – aided process control system

Computer – controller industrial can be classified under one or more of the following categories of operation

(1) Batch or sequential control.
(2) Continuous control.
(3) Supervisory control.
(4) Direct digital control. [7]

2.12 The Case Study

Case study is distillation column in CDU at Khartoum refinery to:

1- Examine; a Control of furnace temp, Control of interface level at reflux drum, Control of level at stripper one, Control of level at bottoms of the column and Control of feed rate (crude oil).
2- Apply advance process control, utilizing technology varying from simple multivariable control to the most advance control system.
3- Various tuning and stability techniques will be used at design with MATLAB software.

2.12.1 Information's about CDU at Khartoum refinery

CDU at Khartoum refinery designed by the East Exploration Design & Research Institute of Petroleum & Natural Gas General Company (CNPC) of the People's Republic of China, the atmospheric distillation equipment of Sudan Khartoum Refinery Co., Ltd. is mainly used to process the mixed crude oil of Heglig, Unity, and Toma etc. of Sudan, with its crude oil processing capability of 2.5 million tons per year. It is fuel type equipment, and its construction was undertaken by China Petroleum Construction Unit, started in August of 1999, finished in January of 2000, as well as succeeded together with its putting into production for one time in February of 2000.

- The crude oil is used in CDU unit is Greater Nile Blend.
- The main products of the CDU unit:
  1- Dry gases.
  2- Naphtha.
  3- Kerosene.
  4- Light gas oil.
  5- Heavy gas oil.
  6- Residue.

- The all controls used in CDU are feedback control system with FOXBORO software.
- The controls signals type are pneumatic.
- The feed process in CDU 315 to 320 ton/h.

- The percentages of products as bellow:

Table 2.3: percentages of CDU unit products

The product	Percentage wt%
Dry gas	**0.2%**
Naphtha	6.45%
Kerosene	3.95%
Light Diesel	14.67%
Heavy Diesel	5.20%
Long Residue	69.38%
Losses	0.15%

- These percentages are stable during last year's.
- To fractionatecrude oil, oil must be heated to 360 ^{0}c by furnace after remove salts from it. [1]
- The CDU unit configuration shown in figure (2.9)

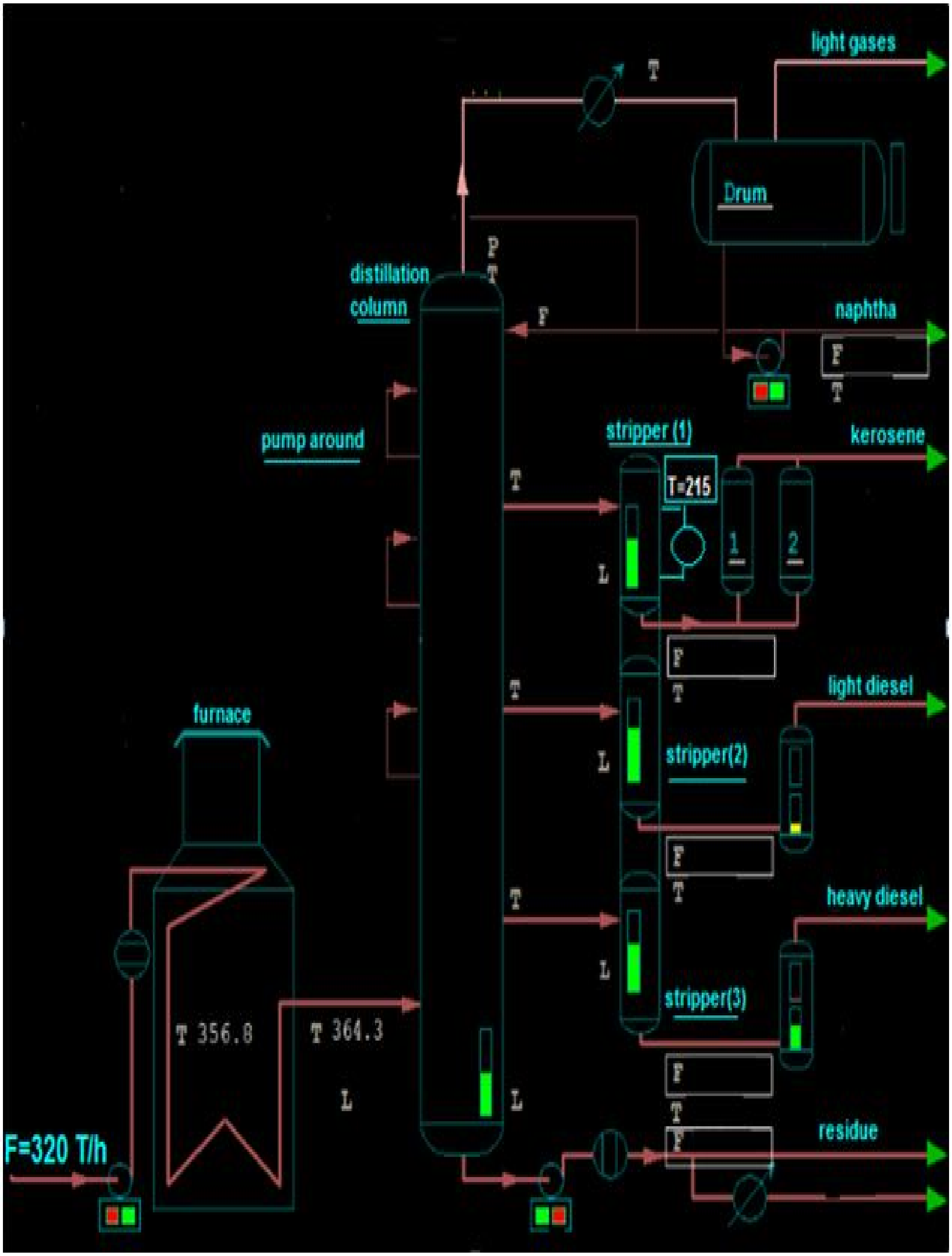


Figure (2.9) configuration of CDU unit

CHAPTER THREE

METHODOLOGY

Chapter three

Methodology

3.1 System Identification

System Identification is the art and science of building mathematical models from measured input-output data. To examine the measured data, and create new data sets from the original one by various preprocessing. Then the models will be estimated using the obtained data. The models are estimated within certain classes of candidate descriptions (model structures), typically by choosing that model that gives the best output fit to the measured data. Quite a few models are normally estimated, and their properties are scrutinized and analyzed in different model views.

In control engineering, the field of system identification uses statistical methods to build mathematical models of dynamical systems from measured data. System identification also includes the optimal design of experiments for efficiently generating informative data for fitting such models. [22]

Ident is a graphical user interface to the System Identification Toolbox that helps you with these tasks. The ident window governs all data handling, model estimation, and model analysis. The ident window also keeps a detailed record of all data sets and models. Data sets and models are represented as icons in the tables of black boxes. [21]

3.1.1 White Box

One could build a so-called white-box model based on first principles,

e.g. a model for a physical process from the Newton equations, but in many cases such models will be complex and possibly even impossible to obtain in reasonable time due to the complex nature of many systems and processes.

A much more common approach is therefore to start from measurements of the behavior of the system and the external influences (inputs to the system) and try to determine a mathematical relation between them without going into the details of what is actually happening inside the system. This approach is called system identification.

Two types of models are common in the field of system identification:

3.1.2 Grey box model

Although the peculiarities of what is going on inside the system are not entirely known, a certain model based on both insight into the system and experimental data is constructed. This model does however still have a number of unknown free parameters which can be estimated using system identification. [21]

3.1.3 Black box model

No prior model is available. Most system identification algorithms are of this type. In the context of non-linear model identification Jin et al.

describe grey box modeling as assuming a model structure a priori and then estimating the model parameters. This model structure can be specialized or more general so that it is applicable to a larger range of systems or devices. The parameter estimation is the tricky part and Jin et al. point out that the search for a good fit to experimental data tends to lead to an increasingly complex model. They then define a black-box model as a model which is very general and thus containing little a priori information on the problem at hand and at the same time being combined with an efficient method for parameter estimation. But as Nielsen and Madsen point out, the choice of parameter estimation can itself be problem-dependent. [21]

In science and engineering, a black box is a device, system or object which can be viewed solely in terms of its input, output and transfer characteristics without any knowledge of its internal workings, that is, its implementation is "opaque" (black). Almost anything might be referred to as a black box: a transistor, an algorithm, or the human mind.

Figure (3.1): Black box model. [22]

3.1.4 Different modeling approaches

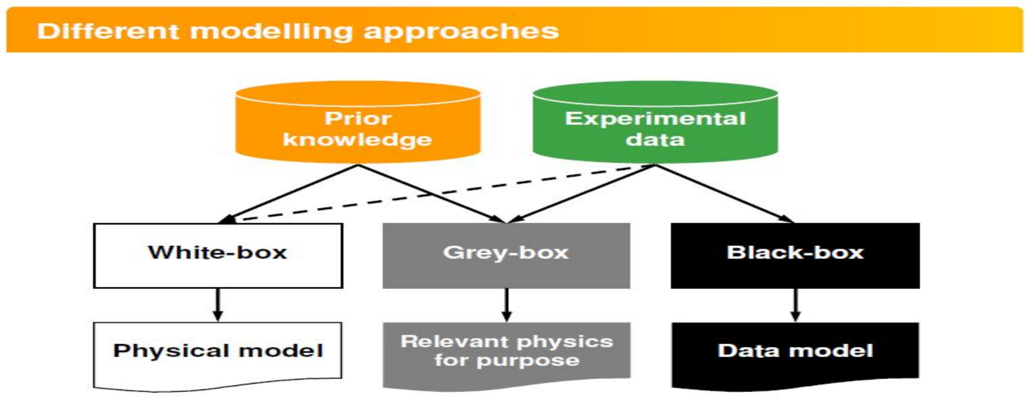


Figure (3.2): different modeling approaches. [22]

3.2 Estimation of the Model

Estimating Models the pop-up menu Estimate provides various methods for estimating models based on the Working Data set. The models will be inserted into the Models board. The pop-up has the following choices:

3.2.1 Parametric models

This allows you to generate dynamic linear models with different structures, orders, and delays. A model structure characterizes the relationship between input and output data and between unknown noise sources and the output data. Supported model structures include ARX, ARMAX, output error (OE), Box-Jenkins (BJ), State-Space, and others. [21]

3.2.2 Process models

This allows you to simple continuous time dynamic linear models characterized by Static gain, time constants and time delays. [21]

3.2.3 Spectral model

Estimates the system's frequency response from the data using either Fourier transform techniques or the Blackman approach. [18]

3.2.4 Correlation model

Estimates the system's transient response (impulse response) by correlating filtered versions of the input-output data. [21]

3.2.5 Quick start

Performs correlation analysis and spectral analysis. Computes a default ARX model with a delay heuristically determined based on the estimated impulse responses of the system as well as a default order state-space model. [21]

3.3 Working and Validation Data

All new data sets and all models are formed from the Working data. This is the data set shown at the center or the ident window. Drag and drop any data set to this position to make it the working data.

All dialogs that depend on the working data will automatically be adjusted to the new set. The two model views model output and model residuals are results of applying the model to a certain data set. This data set is the validation data. It is good practice to let the validation data be different from the working data, which the model was estimated from. The Quick start item of the preprocess pop-up menu gives a default split of a data set into working and validation data. [21]

- MATLAB is the soft ware that will be used at this study.

3.5 MATLAB Soft ware

(Matrix Laboratory) is a programming language for technical computing from The Math Works. Used for a wide variety of scientific and engineering calculations, especially for automatic control. [6]
MATLAB is a high-performance language for technical computing. It integrates computation, visualization, and programming in an easy-to-use environment where problems and solutions are expressed in familiar mathematical notation. Typical uses include:

- Math and computation.
- Algorithm development.
- Modeling, simulation, and prototyping.
- Data analysis, exploration, and visualization.
- Scientific and engineering graphics.
- Application development, including Graphical User Interface building.

MATLAB is an interactive system whose basic data element is an array that does not require dimensioning. This allows you to solve many technical computing problems, especially those with matrix and vector

formulations, in a fraction of the time it would take to write a program in a scalar non interactive language such as C or FORTRAN.
The name MATLAB stands for matrix laboratory. MATLAB was originally written to provide easy access to matrix software developed by the LINPACK and EISPACK projects, which together represent the state-of-the-art in software for matrix computation.

MATLAB has evolved over a period of years with input from many users. In university environments, it is the standard instructional tool for introductory and advanced courses in mathematics, engineering, and science. In industry, MATLAB is the tool of choice for high-productivity research, development, and analysis.

MATLAB features a family of application-specific solutions called toolboxes. Very important to most users of MATLAB, toolboxes allow you to learn and apply specialized technology. Toolboxes are comprehensive collections of MATLAB functions (M-files) that extend the MATLAB environment to solve particular classes of problems. Areas in which toolboxes are available include signal processing, control systems, neural networks, fuzzy logic, wavelets, simulation, and many others.

- **The main parts of MATLAB system**

- The MATLAB language.

This is a high-level matrix/array language with control flow statements,

functions, data structures, input/output, and object-oriented programming features. It allows both "programming in the small" to rapidly create quick and dirty throw-away programs, and "programming in the large" to create complete large and complex application programs.

- The MATLAB working environment.

This is the set of tools and facilities that you work with as the MATLAB user or programmer. It includes facilities for managing the variables in your workspace and importing and exporting data. It also includes tools for developing, managing, debugging, and profiling M-files, MATLAB's applications.

- Handle Graphics.

This is the MATLAB graphics system. It includes high-level commands for two-dimensional and three-dimensional data visualization, image processing, animation, and presentation graphics. It also includes low-level commands that allow you to fully customize the appearance of graphics as well as to build complete Graphical User Interfaces on your MATLAB applications.

- The MATLAB mathematical function library.

This is a vast collection of computational algorithms ranging from elementary functions like sum, sine, cosine, and complex arithmetic, to more sophisticated functions like matrix inverse, matrix eigenvalues, Bessel functions, and fast Fourier transforms.

- The MATLAB Application Program Interface (API).

This is a library that allows you to write C and Fortran programs that interact with MATLAB. It include facilities for calling routines from MATLAB (dynamic linking), calling MATLAB as a computational engine, and for reading and writing MAT-files.

- Black box model is suitable model for this case study. [6]
- The ARMAX is stable model, and very useful model for control purpose. ARMAX supported in the multiple input, multiple output case. ARMAX corresponds to estimating the matrix. [21]

The System Identification Toolbox provides a graphical user interface (GUI). The GUI covers most of the toolbox's functions and gives easy access to all variables that are created during a session. [6]

- **System identification transfer function steps**

It is started by typing: "ident" in the MATLAB command window.

1- Start up the System Id toolbox; use the command "ident". A GUI (graphical user interface) for the toolbox.

2- To import data into the identification tool, select "data", and then "import".

3- Specify the variable containing your time series as an output variable (which it is – it is NOT an input).Note that you have to enter "[]" in the input entry. Adding this tells the tool that: there is no input variable (if you don't do this, you will be asked to confirm that there is no input variable – you can click (yes). Once you have filled in the entries, click on "import".

   Your workspace will now look as follows:

4- Important things to know – notice the "working data" and "validation data" boxes on the GUI. You can specify the dataset to use for estimating your model, or validating your model, by dragging dataset icons from the “data views” panel on the left to the “working data” and “validation data” boxes. Working data is the dataset used to estimate the model. Validation data is data that you use to judge how well the model works – it may be the same as the estimation data, or if you have another dataset collected under the same circumstances, you can use it as a benchmark to test your model by predicted the new data values and comparing them. If there is only one dataset, both “working data” and “validation data” are set to the one dataset. Select “remove means”, and you will now get a new dataset to work with.

5- Drag the mean-centered dataset to the Working Data and Validation Data boxes. Common mistake – not dragging the mean-centered data to the Validation Data box – if you don’t do this when you are using mean-centered data to estimate the model, there will be a constant offset, and your residual autocorrelation diagnostics will be awful.
6- Estimate models. First, go to the “estimate” menu window, and click on the arrow icon.
7- Click on the arrow associated with the “structure” window to choose the type of mode .Select “parametric models "and choose the type of model.
8- amx2221 model must be move to work space. Amx2221 model will appear at command window, define the amx2221 model on command window, and then type the order to get transfer function for the all control system must be designed**.**

3.6 KRC-CDU Case Study

To fractionate crude oil, oil must be heated to 360 ^{0}C by furnace after remove salts from it. The main products of CDU unit are LPG, naphtha, kerosene, diesel and residue.

Control is necessary in CDU unit to avoid any problems that can make the process failed. All controls used in CDU are feedback control system with FOXBORO software. [1]

Figure 3.3 CDU unit.

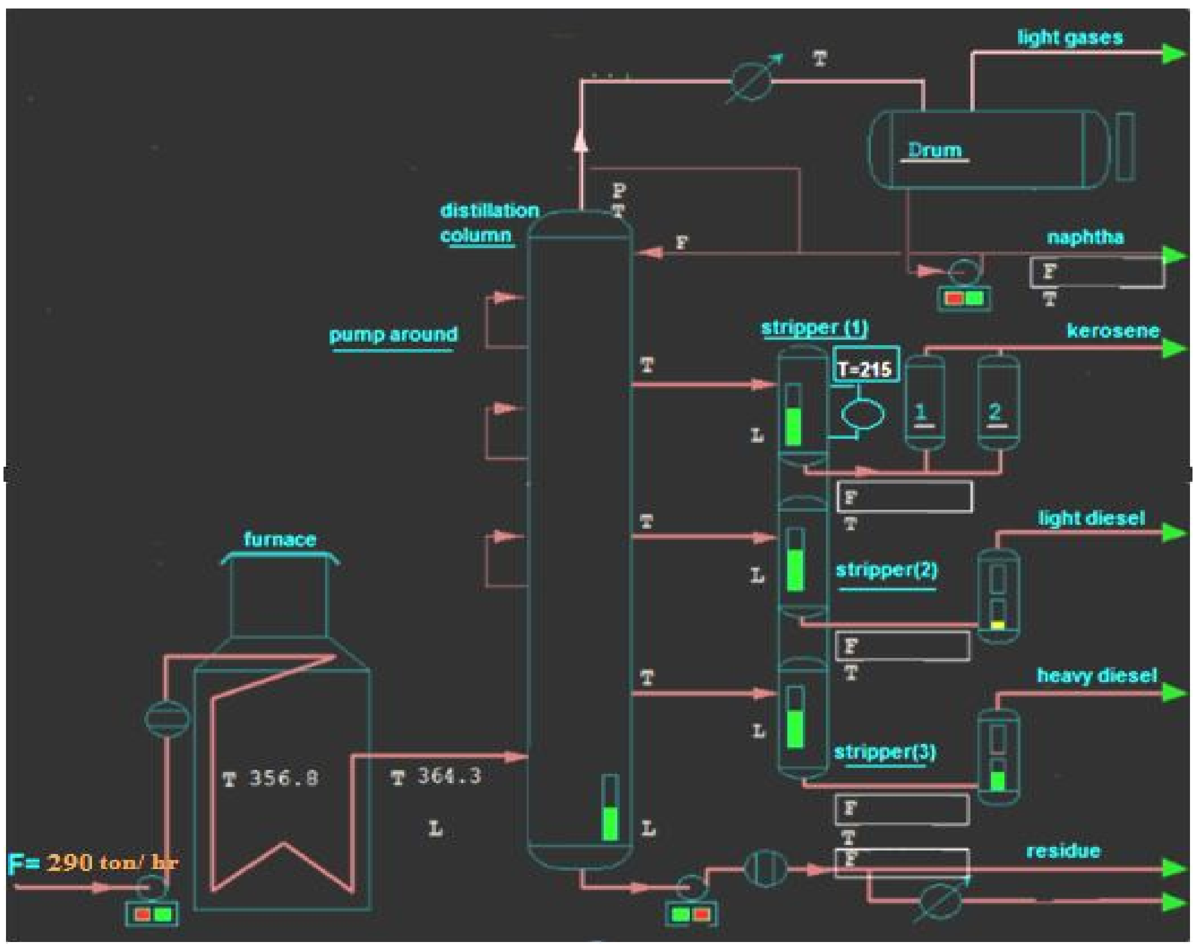


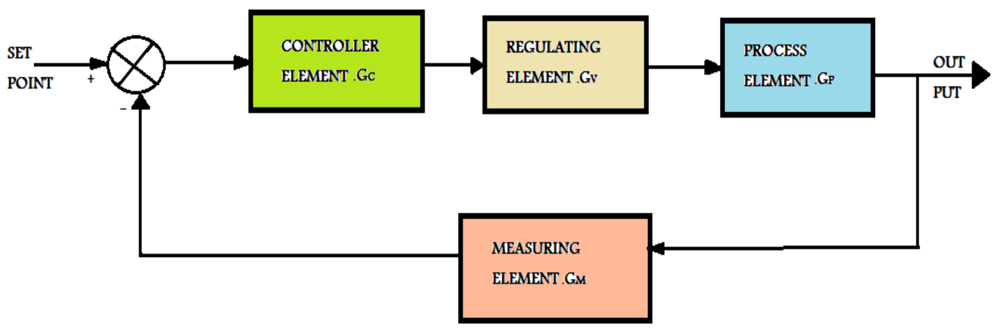
Figure 3.3: CDU configuration. [1]

3.7 Feedback control system

The feedback control action measures the value of the output using the measuring device, which sends the signal through the transmitter to the controller.

The controller compares this value with the desired value (set point) and supplies the deviation signal to the final control element, which in turn changes the value of the manipulated variable. [4]

See figure 3.3a, b:



Fi

gure 3.4a: Feed Back Control System. [4]

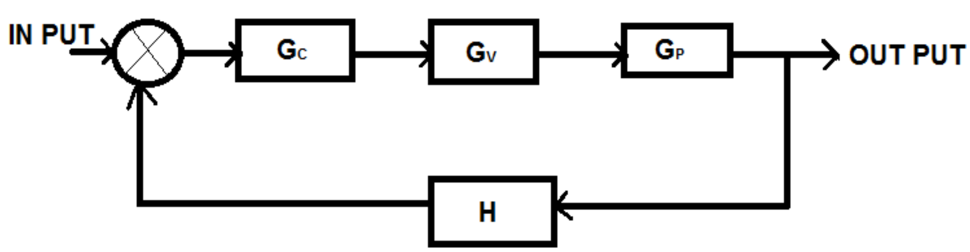


Figure 3.4b: Feed Back Control System. [4]

3.8 Procedure

1- The systems transfer function can be identified.

2- The closed loop characteristic equation of the system will be investigated.

3- The KC, Ʈd and Ʈi parameters can be adjusted from Z-N table.

4- The offset for the systems will investigate by calculation .

5- The system performance using P,PI and PID, upon a unit step change in set point can be drawn.

6- Z-N method, Root Locus method and decay ratio method will be used .

7- The controller performance in each one will be compared

8- The controller that gives the best performance can be select.

CHAPTER FOUR

RESULTS AND DISCUSSIONS

Chapter four

Results and discussion

4.1 The control systems

The control systems that will be examined for a Crude Distillation Unit (CDU) unit are:

1- Control of furnace temp.

2- Control of level at kerosene stripper.

3- Control of pressure at top of the column.

4- Control of level at bottom of the column.

All control systems appearance in a model below in figure**(4.1)**

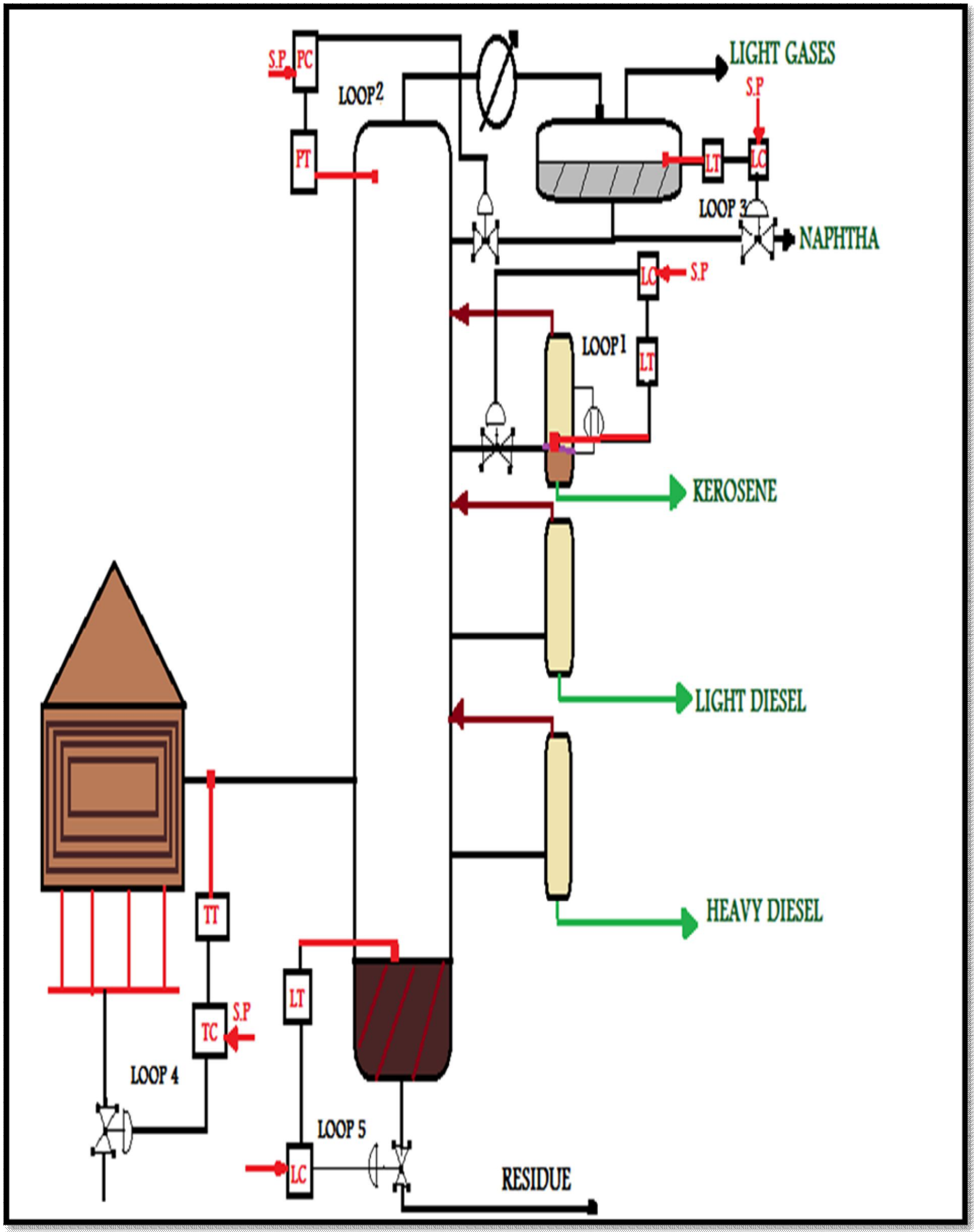


Figure 4.1 control systems strategy

4.1 Transfer functions identification

4.1.1 Control of furnace temp

The furnace temp at (CDU) unit at Khartoum Refinery is the primary control loop and flow rate of fuel oil is the secondary control loop. The MATLAB tool box system identification by the Graphical User Interface will used to determine the model and the transfer functions of the feedback control loops. Different tuning methods are used for investigating the system stability.

- **Transfer function identification:**

The data obtained for the furnace of CDU unit at Khartoum refinery are:

Table (4.1) Flow rate of fuel oil vs Temp of oil

Flow rate of fuel oil (ton/hr)	Temp of oil ^{0}C
Manipulated variable (input)	Controlled variable (output)
1880	363.7
1885	363.8
1897	364.2
1899	364.4
1900	364.4
1901	364.5
2098	364.6
2099	364.7
2100	365.0
2110	365.1

Apply steps in appendix (3.A) by entering the data to MATLAB software to get the transfer function.

The transfer function is:

$$\mathbf{G_{(s)}} = \mathbf{G_{P(s)}} = \frac{0.0004\,S^2 - 0.0042\,S + 0.0059}{S^3 + 0.825\,S^2 + 10.04\,S + 0.028} \quad \text{.........(4.1)}$$

4.1.2Control of level at bottom of the column

Level at **the bottom of distillation column at** (CDU) unit at Khartoum Refinery is the primary control loop and **flow rate of long residue** is the secondary control loop. The MATLAB tool box system identification by the Graphical User Interface will used to determine the model and the transfer functions of the feedback control loops. Different tuning methods are used for investigating the system stability**.**

- **Transfer function identification**

The data obtained from the fractionators of CDU unit at Khartoum refinery are:

Table (4.2): Long residue vs Level at bottom

Long residue ton/hr	Level at bottom m %
Manipulated variable(input)	Controlled variable (output)
183.1	54.1
187.6	54.0

188.0	53.3
210.4	52.1
223.8	51.2
223.0	51.7
183.1	54.0
187.6	53.6
188.0	53.6
210.4	52.2

Apply steps in appendix (3.A) by entering the data to MATLAB software to get the transfer function. The transfer function is:

$$G_{(S)} = G_{P(S)} = \frac{0.0852\ S + 0.0288}{S^2 - 0.840\ S\ \ + 0.110} ......(4.2)$$

4.1.3 Control of pressure at top of the column

The top pressure of the distillation column at (CDU) unit at Khartoum Refinery is the primary control loop and the reflux flow rate is the secondary control loop. The MATLAB tool box system identification by the Graphical User Interface will used to determine the model and the transfer functions of the feedback control loops. Different tuning methods are used for investigating the system stability.

- **Transfer function identification:**

The data obtained from the fractionators of CDU unit at Khartoum refinery are:

Table (4.3): Reflux flow rate at drum VS Pressure at top of distillation column

Reflux flow rate at drum ton/hr	Pressure at top of distillation column MPa
Manipulated variable (input)	Controlled variable (output)
14.00	0.059
14.10	0.059
14.20	0.059
14.20	0.060
14.30	0.061
15.10	0.068
15.00	0.067
15.20	0.068
17.90	0.073
18.90	0.074

Apply previous steps by entering the data to MATLAB software to get the transfer function. The transfer function is:

$$G_{(S)} = G_{P(S)} = \frac{0.0001S^2 - 0.0008S + 0.0017}{S^3 + 1.291S^2 + 10.290S + 0.0137} \text{.......(4.3)}$$

4.1.4 **Control of level at stripper (kerosine stripper)**

Level at the stripper number one at (CDU) unit at Khartoum Refinery is the primary control loop Temp of kerosene enters to stripper is the secondary control loop. The MATLAB tool box system identification by the Graphical User Interface will used to determine the model and the transfer functions of the feedback control loops. Different tuning methods are used for investigating the system stability.

- **Transfer function transfer function identification:**

The data obtained from stripper one of CDU unit at Khartoum refinery is:

Table (3.4): Temp of kerosene enters to stripper one verses kerosene Level at stripper

Temp of kerosene enters to stripper one ^{0}C	kerosene Level at stripper m %
Manipulated variable (input)	Controlled variable (output)
166.7	51.4
172.7	59.6
172.7	56.3
167.0	53.0
166.6	51.4
166.5	51.3
172.7	57.6
167.4	52.4
174.8	56.8
175.1	67.6
175.0	67.4
174.9	56.9

Apply previous steps by entering the data to MATLAB software to get the transfer function.

The transfer function is:

$$G(S) = GP(S) = \frac{-0.3645S + 0.6573}{S^2 + 1.642S + 1.982} \quad \text{......(4.4)}$$

4.2 Controller tuning:

The controller is the active element that is receives the information from the measurements and takes appropriate control actions to adjust the value of the manipulated variables taking the best response of the process using different controller laws.

In order to be able to use a controller, it must first be tuned to the system. This tuning synchronizes the controller with the controlled variable, thus allowing the process to be kept at its desired operating condition. [1]

4.2.1 Controller tuning methods:

There are several methods to controller tuning. In this research the methods will used:

1- Ziegler-Nicholas (Z-N) tuning method

2- Root locus tuning method.

3-Decay ratio tuning method

After test the controller tuning methods the best one will be choose. After that, the three controller's action (P, PI, and PID) will be studied to use the best one controller.

4.2.2 Ziegler-Nicholas (Z-N) tuning method with using MATLAB soft ware

Ziegler-Nicholas is the one of tuning method techniques. It goes through the following steps:

Step1: Set up the system with proportional control only, i.e. set (τ_d) at its minimum value and (τ_i) at its maximum value.

Step 2: make a set point step test and observe the response.

Step 3: Evaluate the period of the constant oscillation; this period is called the ultimate period $P_{u.}$

Step 4: Calculate the parameters according to the following formulas:

Table 4.1: Ziegler-Nicholas Closed Loop Relevant Controller Parameters.

Controller type	**Gain K_c**	**Integral time τ_i**	**Derivative time τ_d**
P	0.5 k_u	-	-
PI	0.45 K_u	$P_u/1.2$	-
PID	0.6 K_u	$P_u/2$	$P_u/8$

4.3 Control of level at kerosene stripper

The process transfer function at stripper one is:

$$G(S) = GP(S) = \frac{-0.3645S + 0.6573}{S^2 + 1.642S + 1.982} \quad ......(4.4)$$

4.3.1 P-controller

Using Simulink Library Browser to draw feedback control system with p-controller

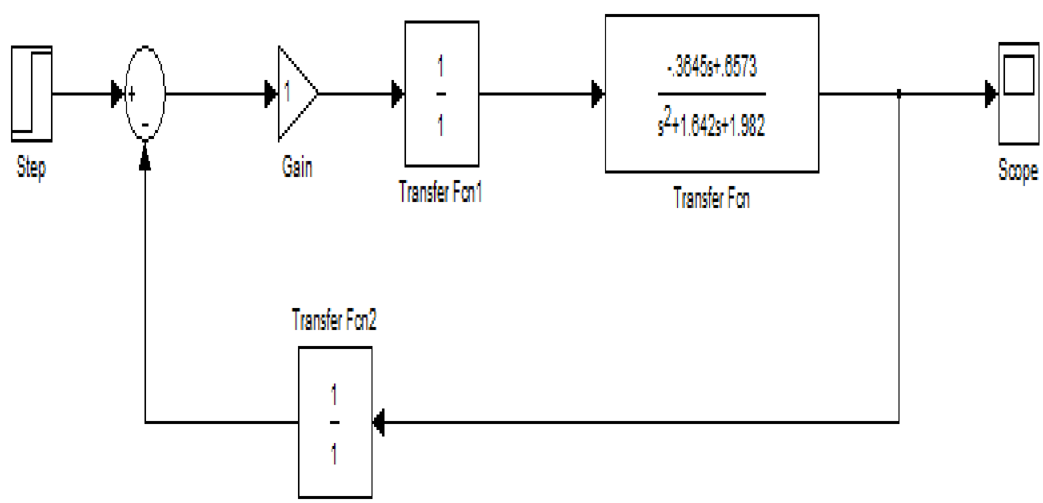


Figure (4.1) Block diagram for kerosene stripper with P- controller adjustable MATLAB parameters.

4.3.2 Ziegler Nicholas tuning method

The trials to estimate the ultimate gain when the controller is proportional shown in figures below.

- The first trial

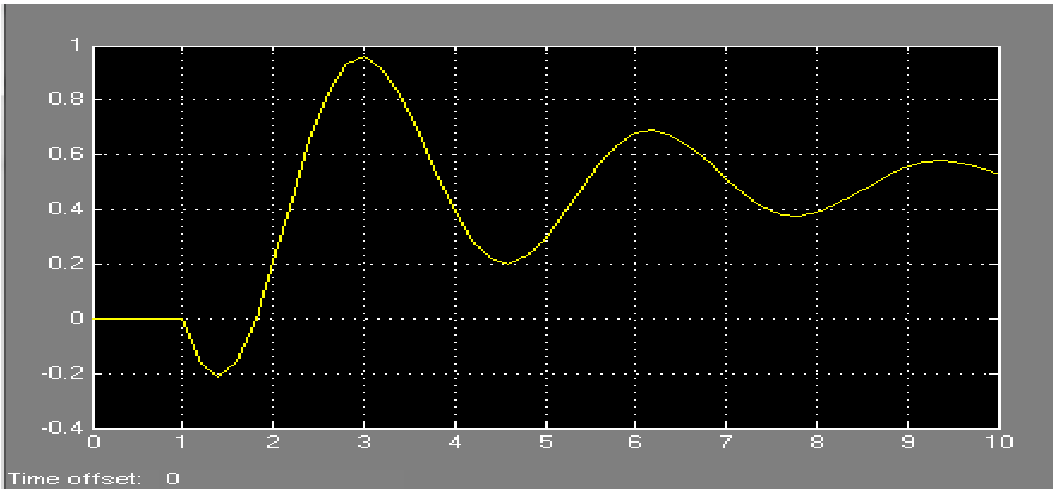


Figure (4.2a) the first trial

- The second trial

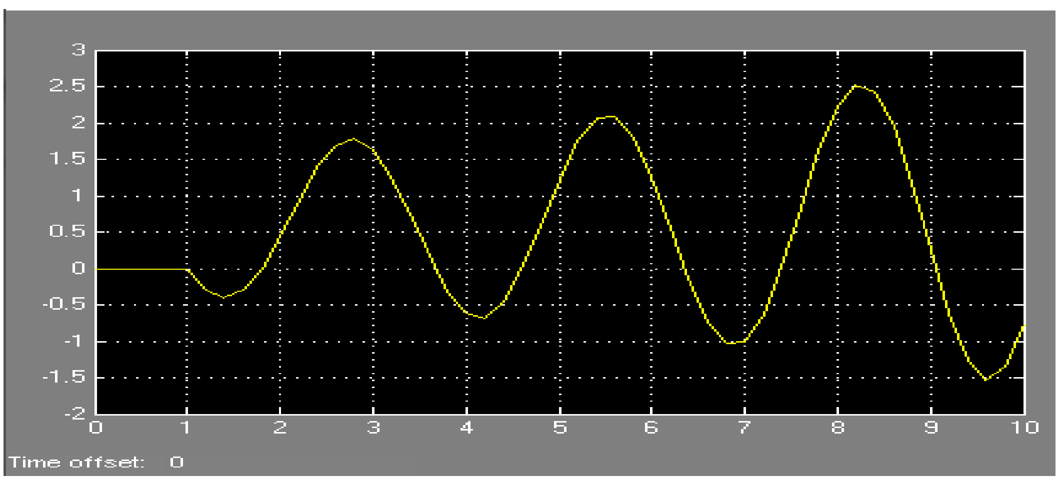


Figure (4.2b) the second trial

- Third trial estimated the ultimate gain

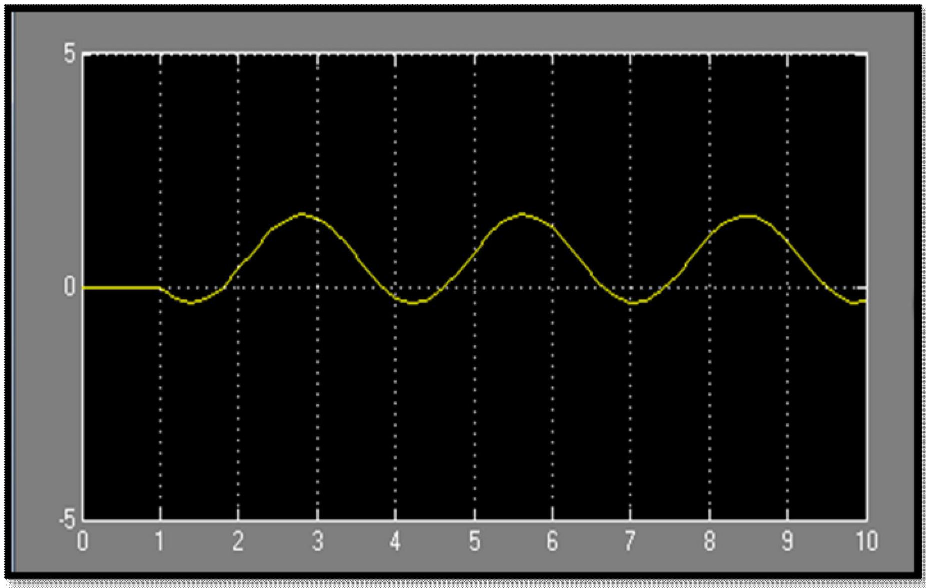


Figure (4.2c): Z-N tuning plot for P-controller at critical stable.

- From figure (4.2c)

  Ku = 4.51

From Ziegler table at P– controller

Ku = 0.5Kc

∴ Kc = 0.5(4.51) = 2.25

∴ The controller transfer function

$G_{C(S)} = 2.25$ ……………………………………….. (4.4)

The overall transfer Function:

The block diagram at P-controller shown below:

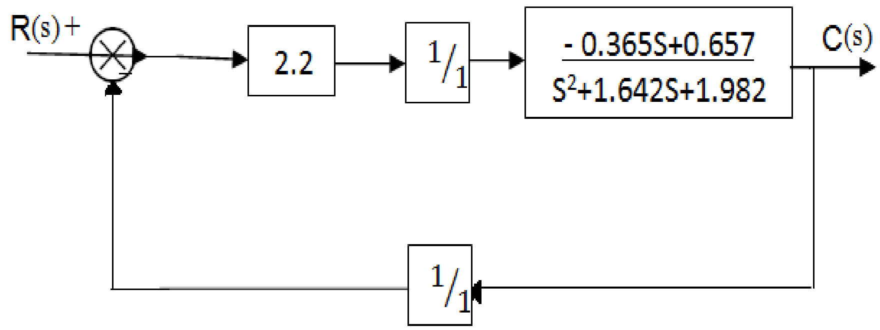


Figure (4.2) the overall block diagram for stripper one

$$OATF = G(s) = \frac{2.25\left[\frac{-0.365S+0.657}{S^2+1.642S+1.982}\right]}{1+2.25\left[\frac{-0.365S+0.657}{S^2+1.642S+1.982}\right]}$$

$$G(s) = \left[\frac{-0.821S+1.48}{S^2+0.82S+8.462}\right]$$

- **The offset:**

Off set is the deviation from steady state.

$$G(s) = \frac{C(s)}{R(s)} = \frac{\text{out put}}{\text{in put}}$$

For a unit step change:

$$r(t) = 1 \xrightarrow{\text{laplace}} R(s) = {}^{1}/_{s}$$

The offset = C (∞) – C (ideal)

For a unit step change:

C (ideal) = 1

$$C(\infty) = \lim_{S \to 0} \left(S\,C(s) \right)$$

$$C(\infty) = \frac{1}{s}\left[\frac{-0.821S+1.48}{S^2+0.82S+8.462}\right]$$

$$C(\infty) = \lim_{S \to 0} \left[S \left[\frac{1}{s} \left(\frac{-0.821S+1.48}{S^2+0.82S+3.462} \right) \right] \right]$$

$$C(\infty) = \frac{1.48}{3.462} = 0.43$$

The offset = C (∞) – C (ideal) …………………… (4.7)

the offset = 0.46 – 1

Offset = – 0.57

Table (4.2) adjustable at strippers for P- controller using Z-N tuning method

Ku	Kc	Offset
4.51	2.25	– 0.57

- Unit step change test using P- controller to measure the swiftness of the system response:

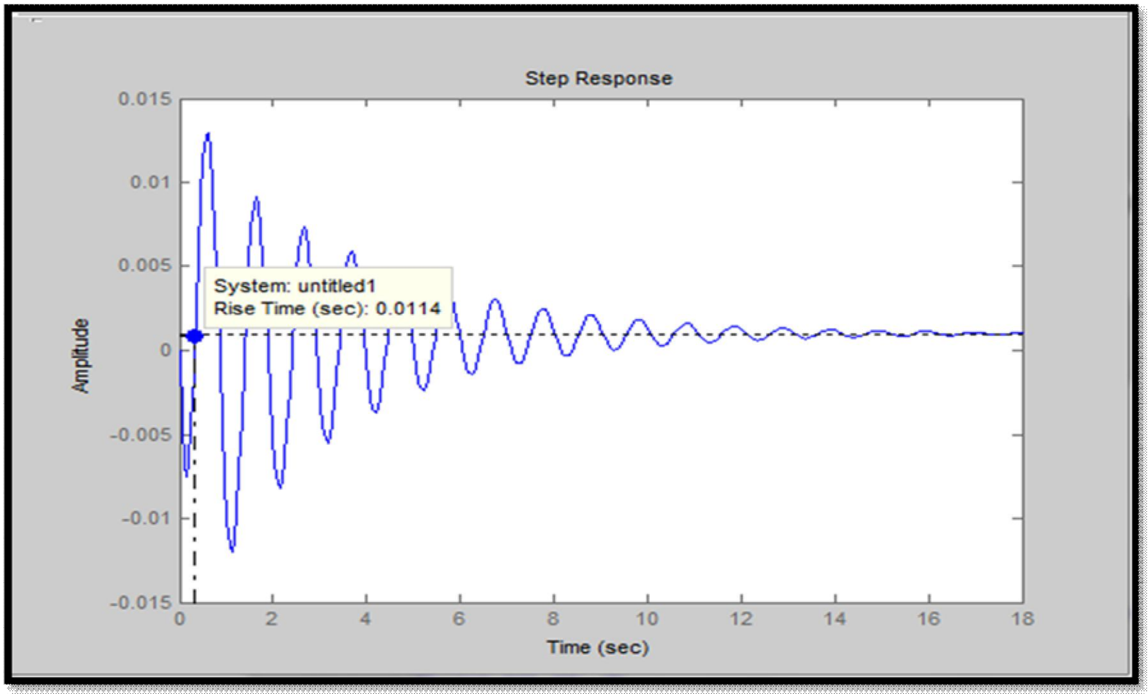


Figure (4.4) rise time at step response for P- controller

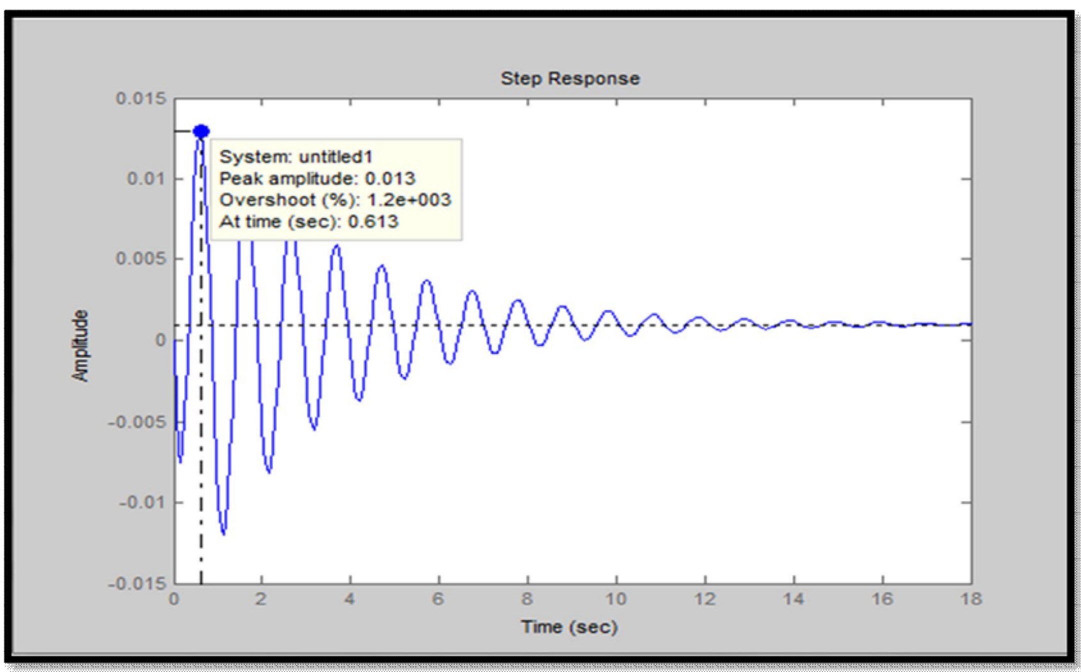


Figure (4.5) overshoot at step response for P- controller

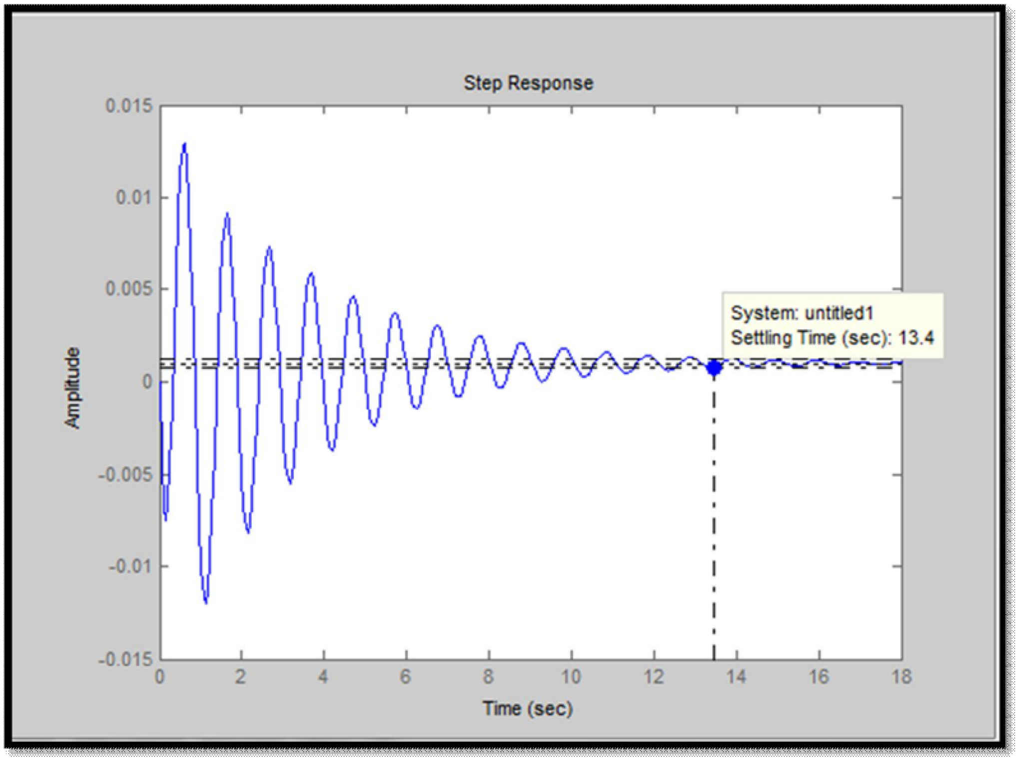


Figure (4.6) setting time at step response for P- controller

4.3.2 PI-controller

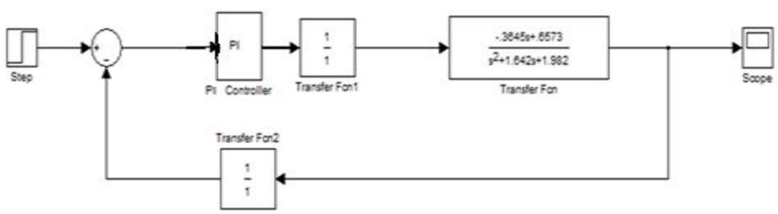


Figure (4.7) Block diagram for stripper one with PI- controller

4.2.2 Ziegler Nicholas tuning method

The trials to estimate the ultimate gain when the controller is proportional integral shown in figures below

- First trial

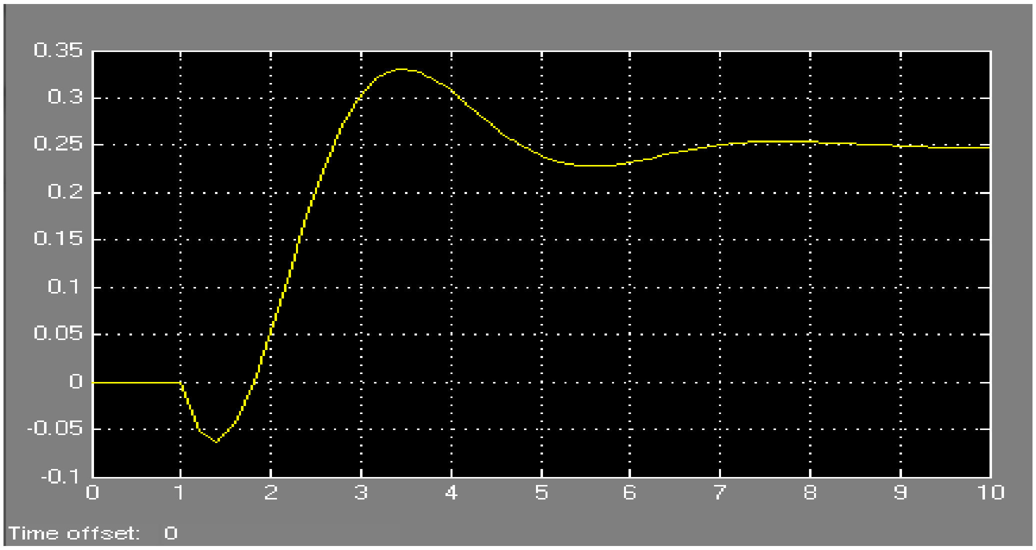


Figure (4.8):first trial

- Second trial

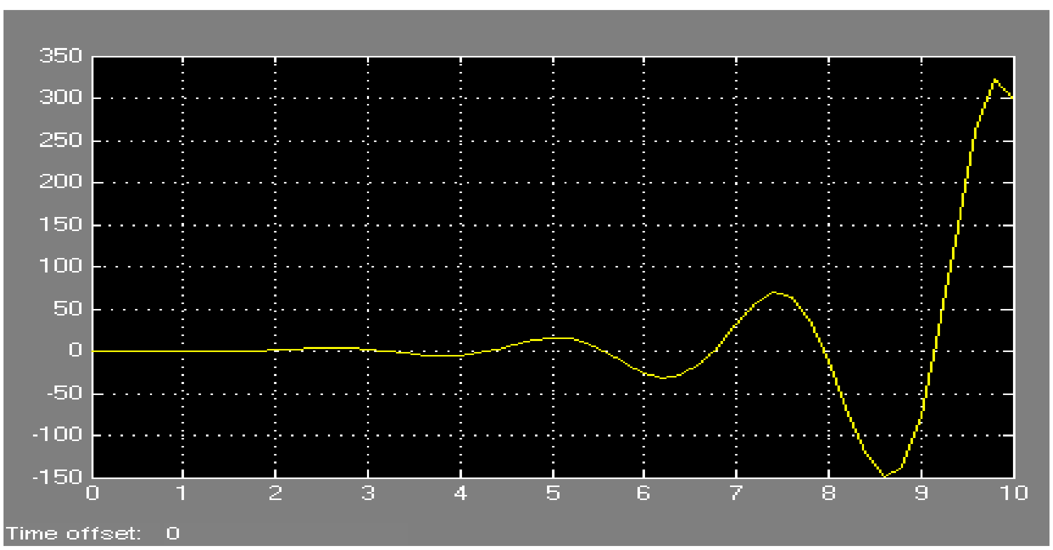


Figure (4.9b): second trial

- Third trial estimated the ultimate gain

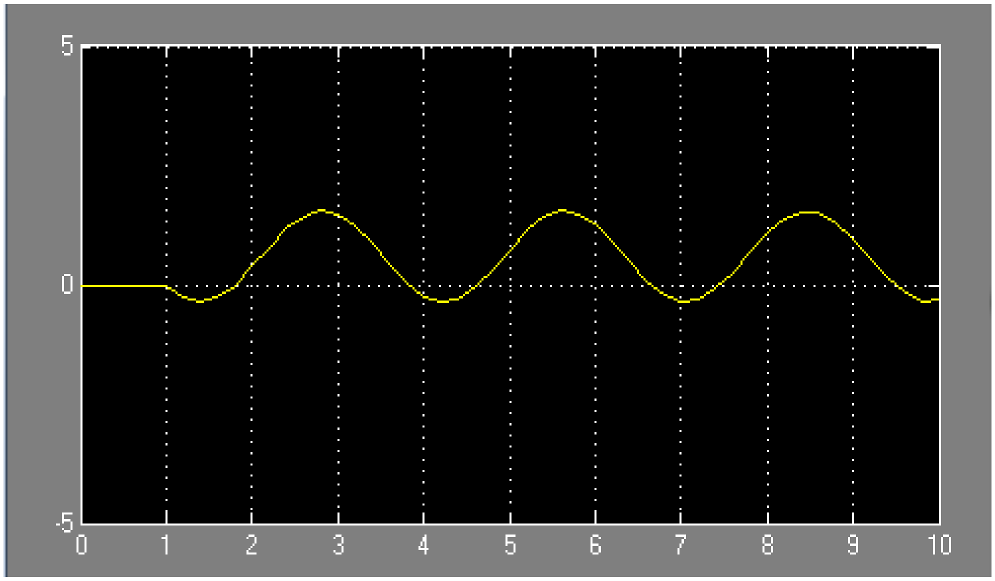


Figure (4.9c) Z-N tuning plot for PI controller

From figure (4.7c):

$K_c = 4.51$

$p_u = 1.81$

From Z-N table for PI – controller

$k_c = 0.45$ ku

τ_i= pu/1.2

∴ Kc = 0.45*2.71 = 2.025

$\therefore$ $\tau i = 1.81/1.2 = 1.508$

$\therefore$ The controller transfer function

$G(c) = k_c \left(1 + \frac{1}{\tau is}\right) = \frac{1}{s}\,(2.025S + 3)$.................(4.8)

The overall transfer Function at PI – controller

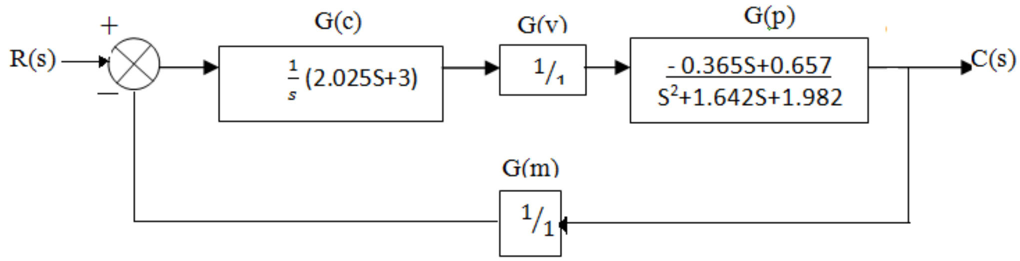


Figure (4.10) the overall block diagram for stripper one

The overall transfer function will be

$$G_{(S)} = \frac{\prod(forward)}{1+\prod(Loop)} \qquad (4.8)$$

$$\prod(forward) = G_c G_v G_p$$

$$\prod(Loop) = G_c G_v G_p G_m$$

$$G(s) = \frac{-5.74S^2 + 0.235S + 1.971}{S^3 + 0.902S^2 + 2.217S + 1.971} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots.(4.9)$$

- **The offset:**

Off set is the deviation from steady state.

The offset at PI controller:

$$\text{OATF} = \frac{G(s)}{R(s)}$$

For a unit step of set point

$$r(t) = 1 \xrightarrow{\text{laplace}} R(s) = {}^{1}/_{s}$$

The offset = C (∞) – C (ideal) …………………………(4.7)

C (ideal) = 1

C(s) = R(s).G(s)

$$C(\infty) = \lim_{S \to 0} [SC(s)]$$

$$C(s) = \frac{1}{s} G(s)$$

$$C(s) = \frac{1}{s}\left[\frac{-0.74S^2+0.235S+1.971}{S^3+0.902S^2+2.217S+1.971}\right]$$

$$C(\infty) = \lim_{S \to 0} S\left[\left[\frac{1}{s}\left(\frac{-0.74S^2-0.235S+1.971}{S^3+0.902S^2+2.217S+1.971}\right)\right]\right]$$

C (∞) = 1.971/1.971= 1

Offset = 1-1=0

Table (4.3) adjustable at strippers for PI- controller using Z-N tuning method

K_U	P_U	K_C	τ_i	Offset
4.51	1.81	2.025	1.508	0.00

- Unit step change test using PI- controller to measure the swiftness of the system response:

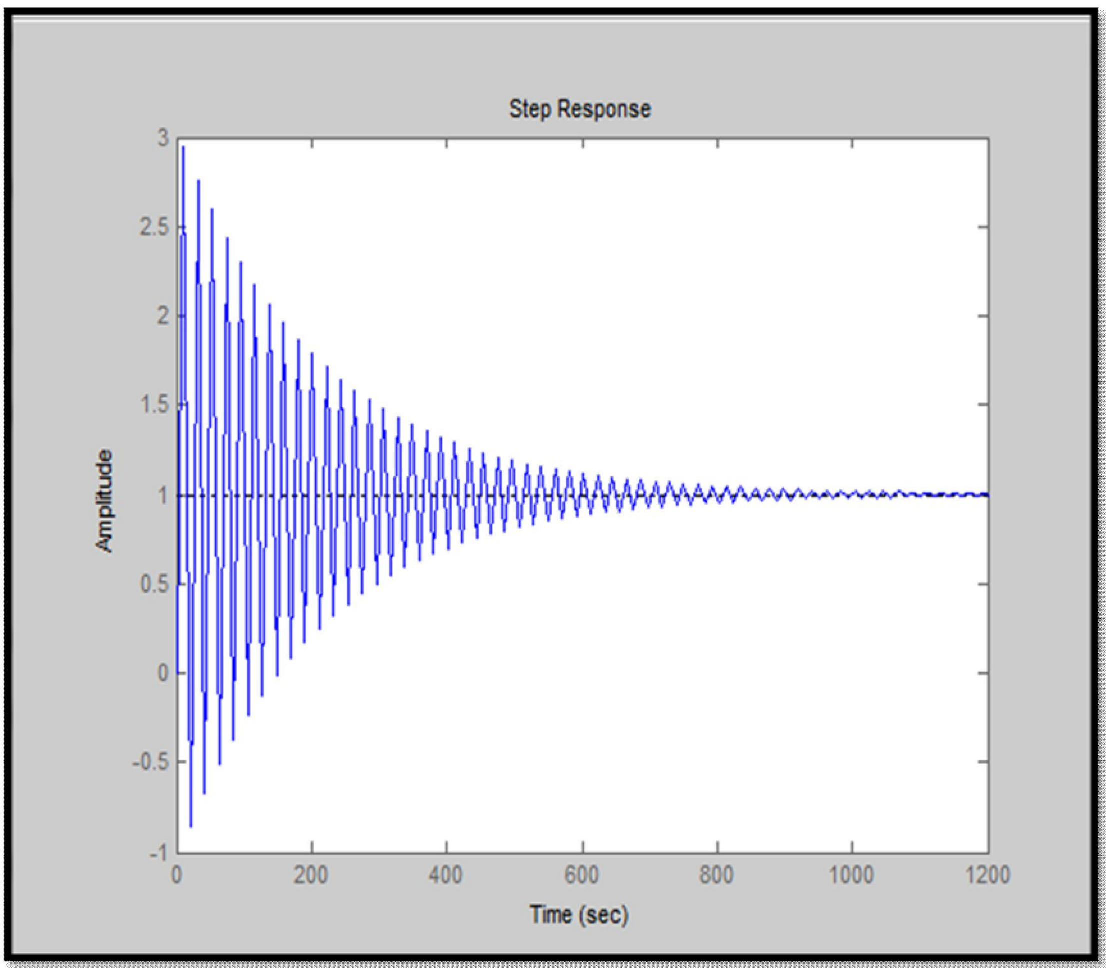


Figure (4.11) step response for PI -controller

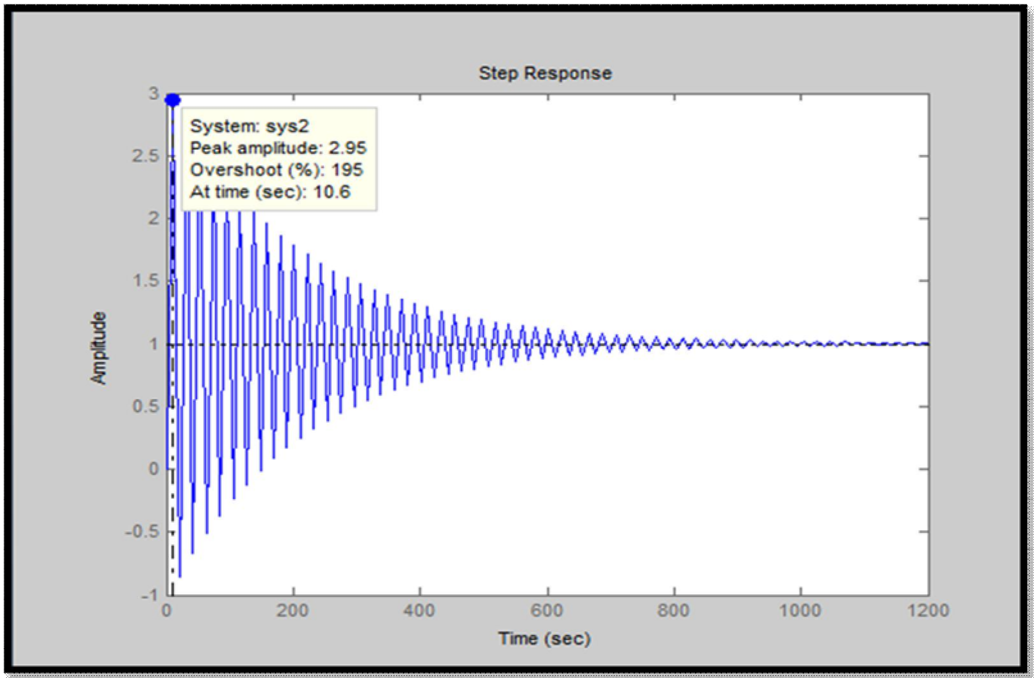


Figure (4.12) overshoot at step response for PI- controller

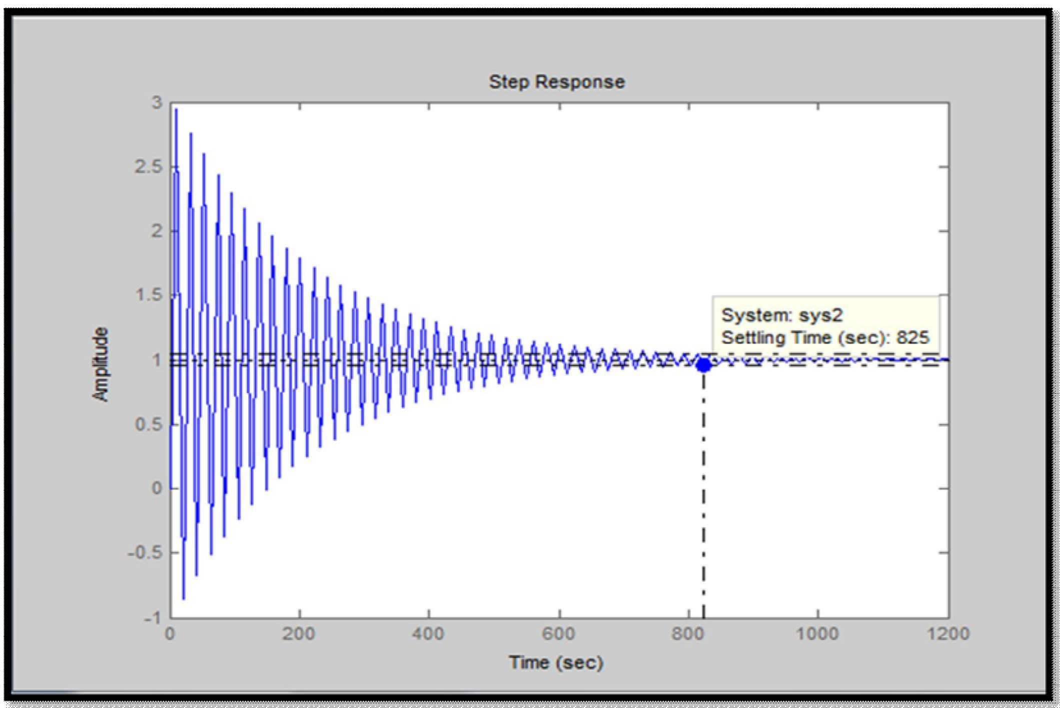


Figure (4.13) setting time at step response for PI -controller

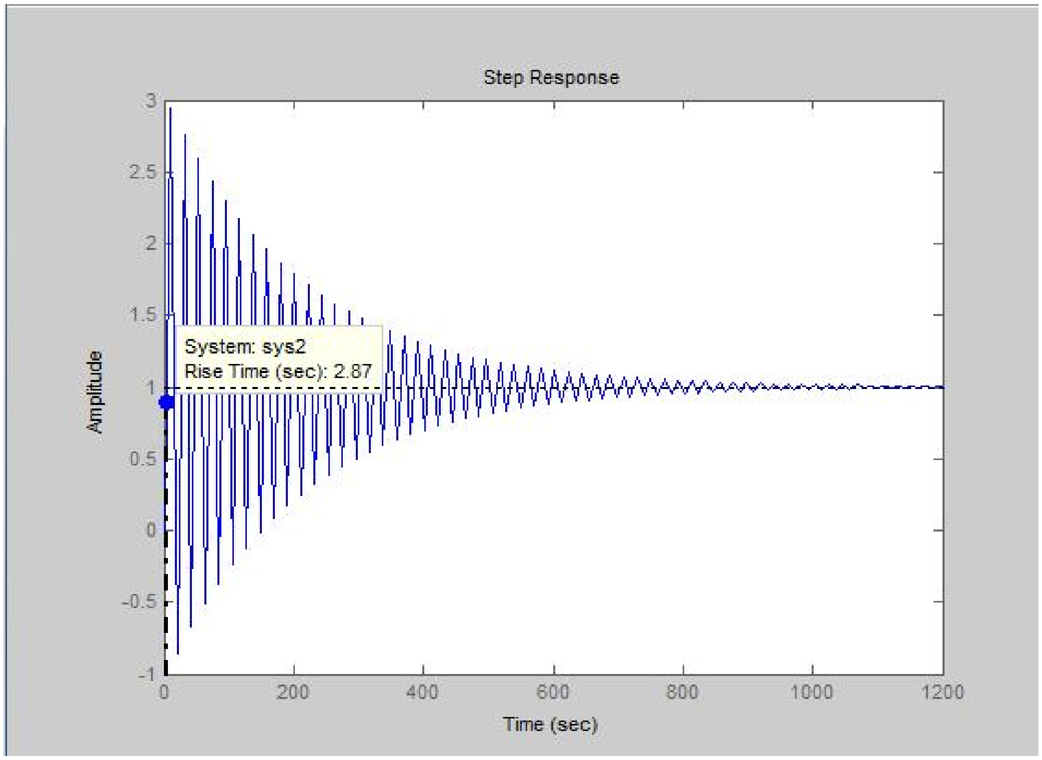


Figure (4.14) rise time at step response for PI- controller

4.3.3 PID controller for stripper one

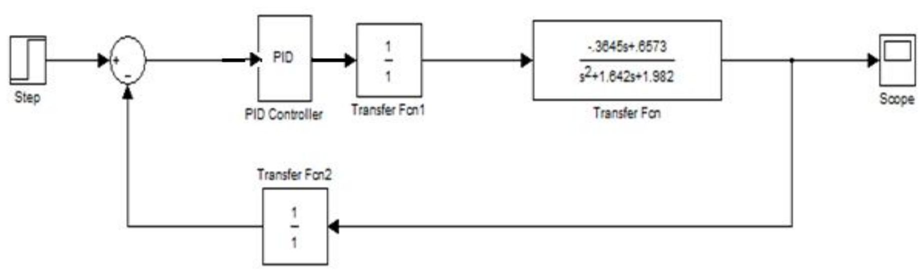


Figure (4.15) block diagram for stripper one at PID controller

4.2.2 Ziegler Nichols tuning method

The trials to estimate the ultimate gain when the controller is proportional integral shown in figures:

- First trial

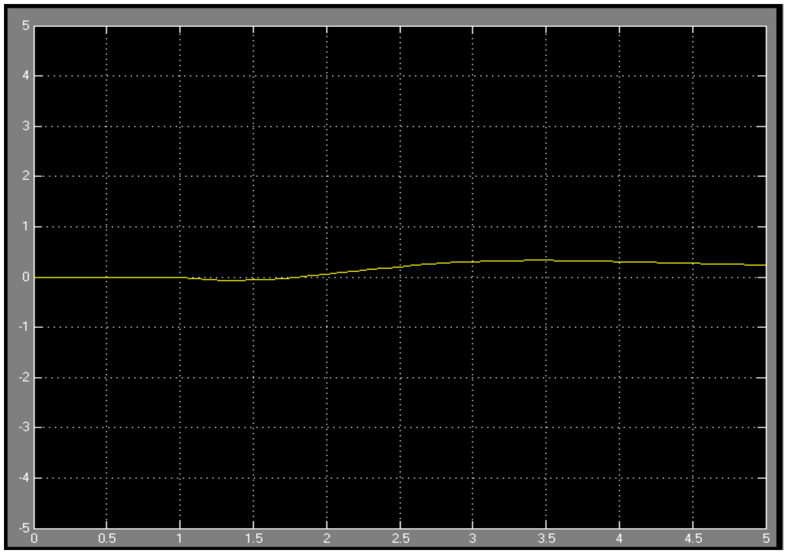


Figure (4.16a): **Z-N** tuning plot for PID controller first trial

Ku = 2, system is stable.

- Second trial

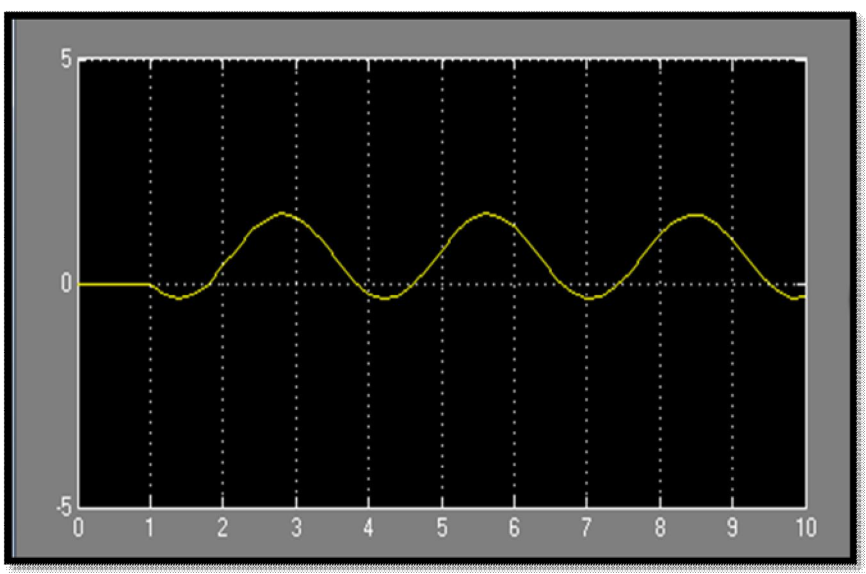


Figure (4.13b) **Z-N** tuning plot for PID controller second trial

Ku = 4.5, system is critical stable.

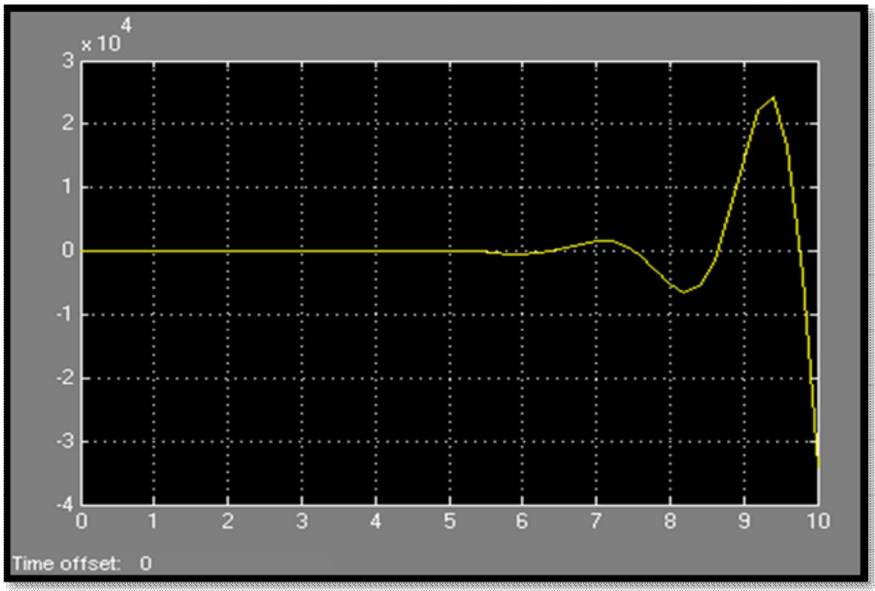


Figure (4.13c): **Z-N** tuning plot for PID controller third trial.

Ku = 11, System is unstable.

From figure (4.13b) system is critical stable:

The ultimate gain

Ku = 4.5

From Z-N setting table for PID controller

Parameters	K_c	Ti	Td
P	$0.5k_u$	–	–
PI	0.45ku	$P_u/1.2$	–
PID	$0.6k_u$	$P_u/_2$	$P_u/8$

$k_u = 4.51$

$P_u = 2.00$

$\therefore K_c = 0.6k_u \Rightarrow k_c = 2.7$

$\therefore \tau_i = p_u/2 \;\Rightarrow \tau_i = 2/2 = 1.00$

$\therefore \tau_d = pu/8 \Rightarrow \tau_d = 2/8 = 0.25$

$\therefore$ the PID controller transfer Function

$$G_c = kc\left(1 + \frac{1}{\tau_i S} + \tau_d S\right)$$

$$Gc(s) = \frac{0.675S^2 + 2.7S + 2.7}{S} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (4.10)$$

The block diagram will be:

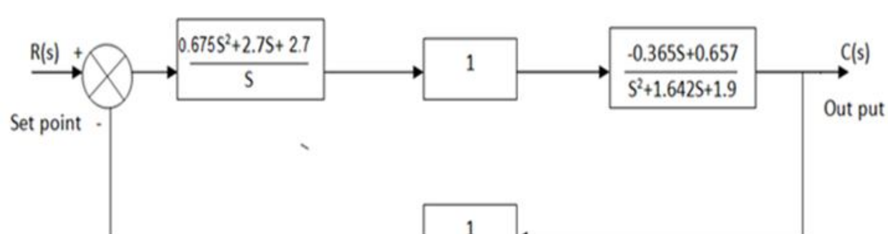
$$Gp(s) = \frac{-0.3654S + 0.657}{S^2 + 1.642S + 1.982} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (*)$$

$$Gc(s) = \frac{0.675S^2 + 2.7S + 2.7}{S} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (4.10)$$

G_p, G_c are the transfer function for process and controller respectively.

G_m=1, (transfer function for measuring element)

G_v=1, (transfer function for valve element)



Figure(4.17): the overall block diagram for stripper one

The overall transfer function will be

$$G_{(S)} = \frac{\Pi(forward)}{1 + \Pi(Loop)} \qquad (4.8)$$

$$\prod(forward) = G_c G_v G_p$$

$$\prod(Loop) = G_c G_v G_p G_m$$

$$G(s) = \frac{-0.245S^3 - 0.546S^2 + 0.78\,S + 1.774}{0.755\,S^3 + 1.094S^2 + 2.685S + 1.774} \qquad \text{.............................. } \mathbf{(4.11)}$$

$\therefore$ the characteristic equation is:

$0.755\,S^3 + 1.094S^2 + 2.685S + 1.774 = 0$ ………................... (4.12)

- **Offset**

Off set is the deviation from steady state.

The offset at PID controller:

The overall transfer function (OATF) is

$$\text{OATF} = \frac{\text{output}}{\text{Input}} = \frac{G(s)}{R(s)} = G(s)$$

R(s) ≡ desired value ≡ input C(s) ≡ output

For step change in r (t) (desired value)

$r(t) = 1 \xrightarrow{\text{laplace}} R(s) = \frac{1}{s}$

$\therefore\ C(s) = \frac{1}{s} G(s)$

The offset = C (∞) – C (ideal) …………………………. (4.7)

For a unit step change:

C (ideal) = 1

$$C(\infty) = \lim_{S \to 0} [SC(s)]$$

C (∞) = lim [SC(s)]

$$C(\infty) = \lim S\left(\frac{1}{s}\left(\frac{0.245S^3 - 0.546S^2 + 0.78\,S + 1.774}{0.755\,S^3 + 1.094S^2 + 2.685S + 1.774}\right)\right)$$

$$C(\infty) = \frac{1.774}{1.774}$$

$$\text{Off set} = 1 - 1$$

$$\text{Offset} = 0$$

Table (4.4) adjustable at strippers for PID- controller using Z-N tuning method

Ku	Pu	Kc	τ_i	τ_d	Offset
4.5	2	2.7	1	0.25	0.00

- Unit step change test using PI- controller to measure the swiftness of the system response:

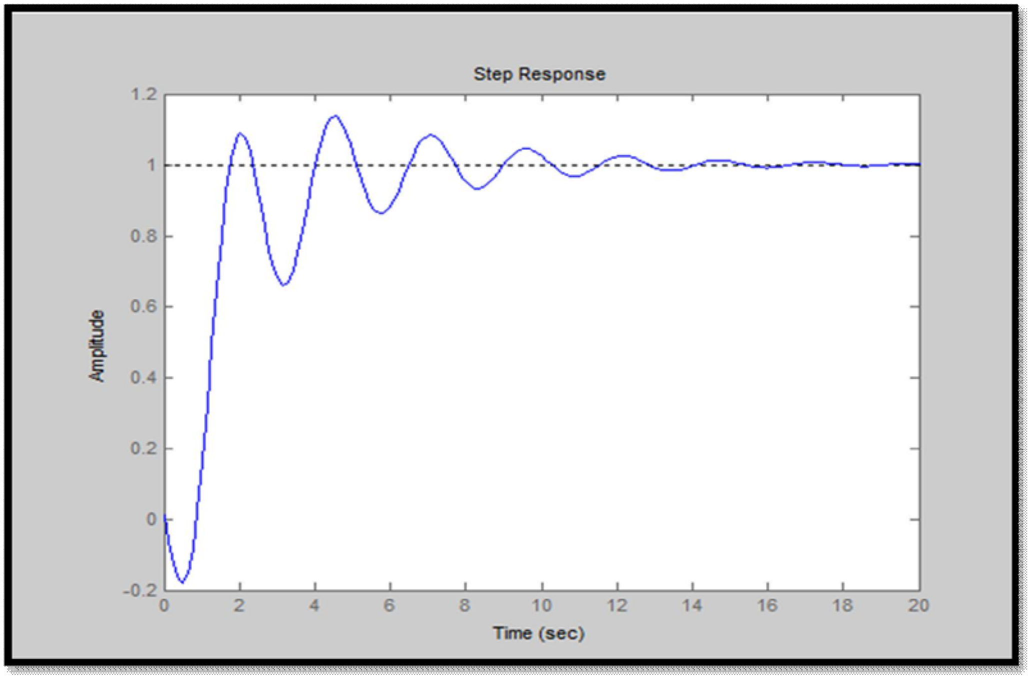


Figure (4.19) step response for PID controller

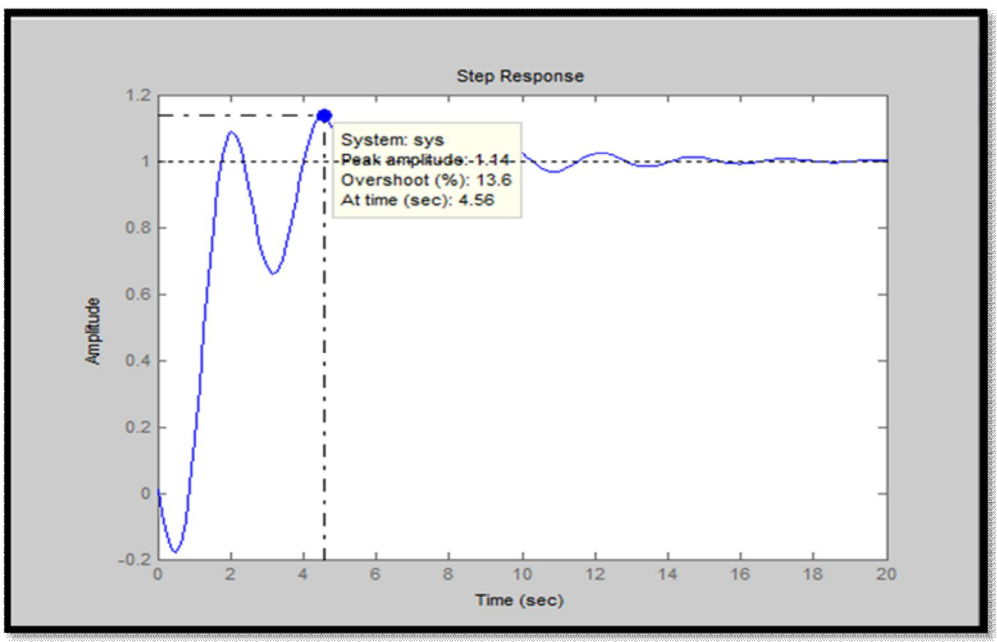


Figure (4.20) overshoot at step response for PID controller

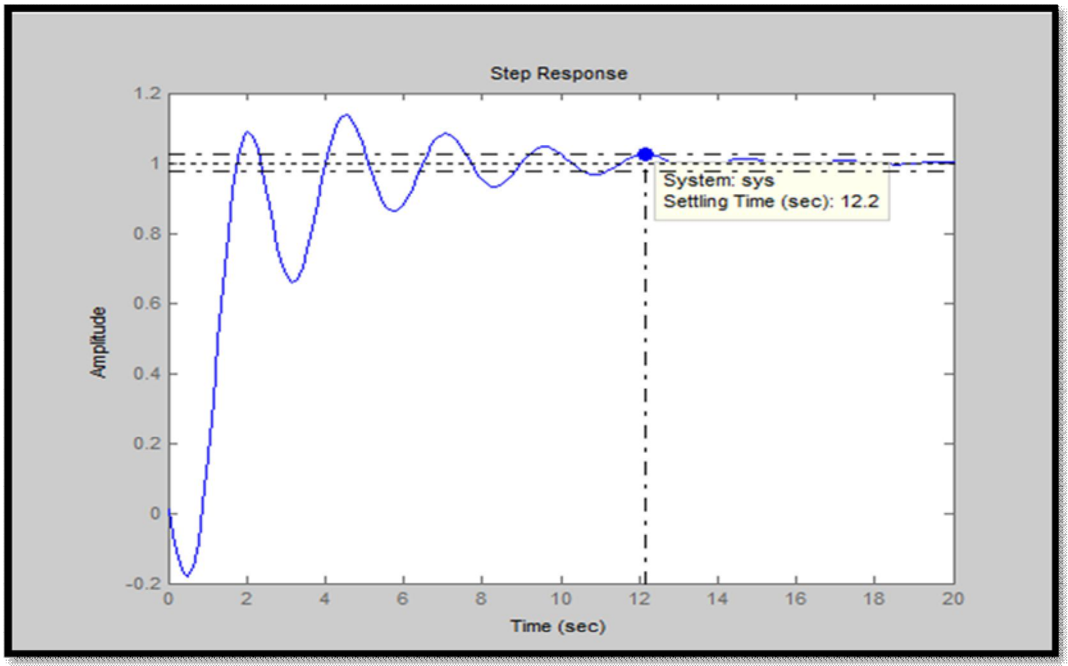


Figure (4.21) setting time at step response for PID controller

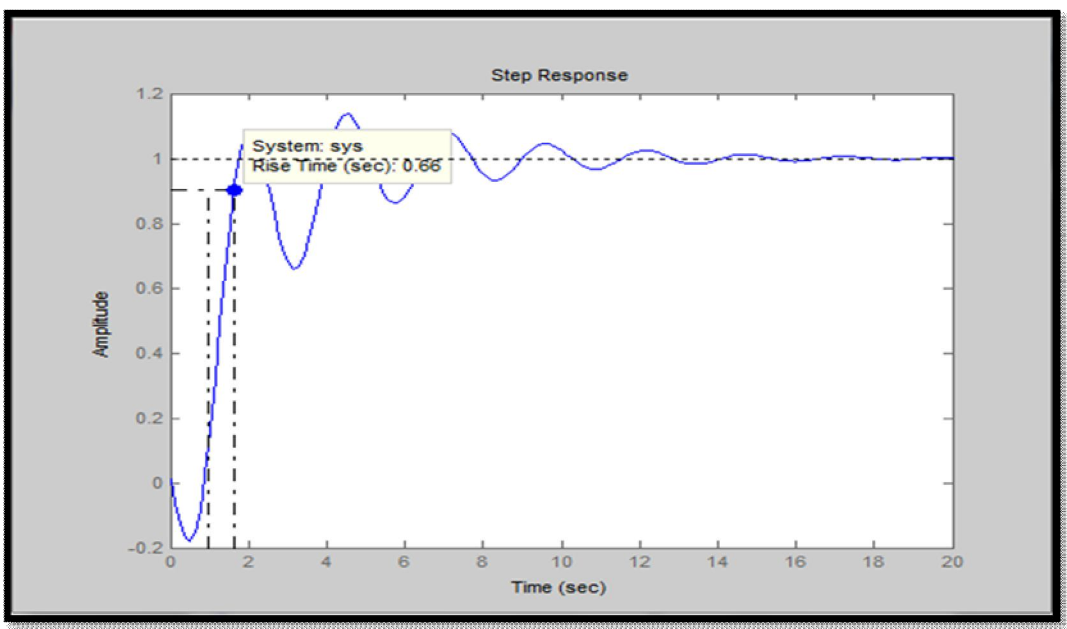


Figure (4.22) rise time at step response for PID controller

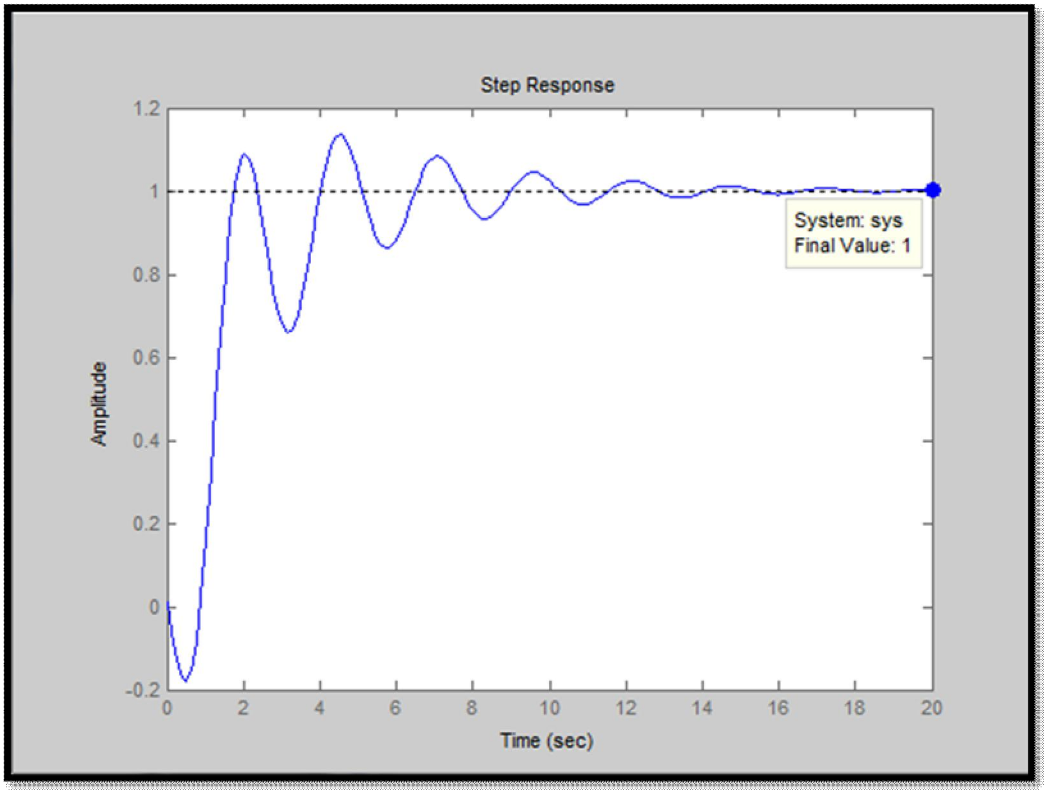


Figure (4.23) final value step response for PID controller

Comparison between PID,PI and P controller to choice the best one.

Table (4.5) comparison between PID, PI, P controller

Controller	Overshoot (%)	Setting time (sec)	Rise time (sec)	offset	Gain k_c
PID	**13.6**	**12.2**	**0.66**	**0.0**	**2.71**
PI	**195**	**825**	**2.87**	**0.0**	**2.025**
P	**66**	**13.4**	**0.011**	**-0.59**	**2.25**

∴ From the previous results the PID controller is the best one controller because PID has high gain with zero offset and low over shoot.

4.4 Root locus tuning method

To determine the ultimate gain and ultimate period from the root locus, this can be realized by the following methods:

Using the close –loop transfer function

$G_c = k_u$

$$Gp(s) = \frac{-0.365S+0.657}{S^2+1.642S+1.982} \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (*)$$

$G_m = 1$

$G_v = 1$

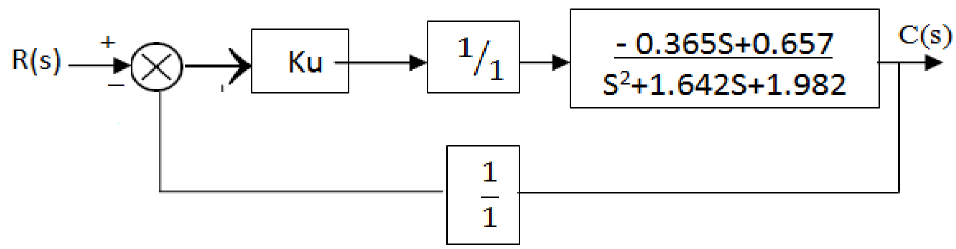


Figure (4.24) the overall block diagram for stripper one

Open loop transfer function is

$$OLTF = ku\ \frac{(-\ 0.365S+0.657)}{(S^2+1.642S+1.952)}$$

Controller tuning using Step change of Root locus method by using MATLAB soft ware:

>>num=[-.365 .657]

den=[1 1.642 1.982] >>

>> SYS= tf(num,den)

Transfer function:

s + 0.6570.365 -

----

s^2 + 1.642 s +1.982

rlocus (sys) >>

The Root locus plot of the feedback transfer function is plotted using MATLAB as shown in figure (4.21) below

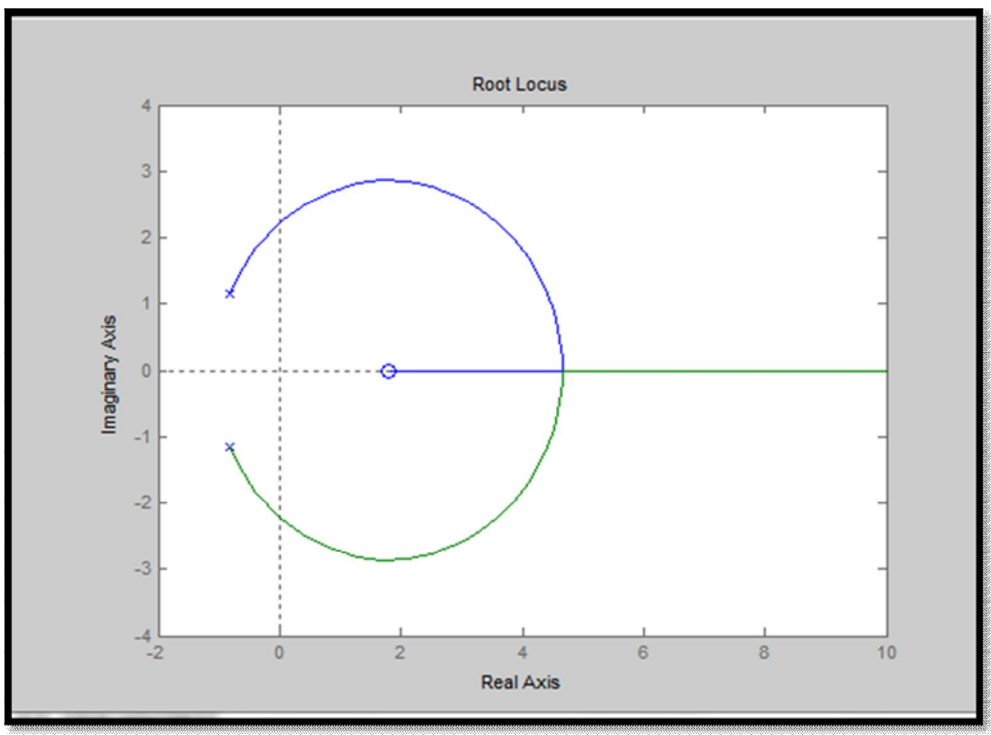


Figure (4.25) root locus plot of feedback system for Stripper one

The root locus plot of the feedback transfer function is plotted using MATLAB as shown in figure (4.22) below

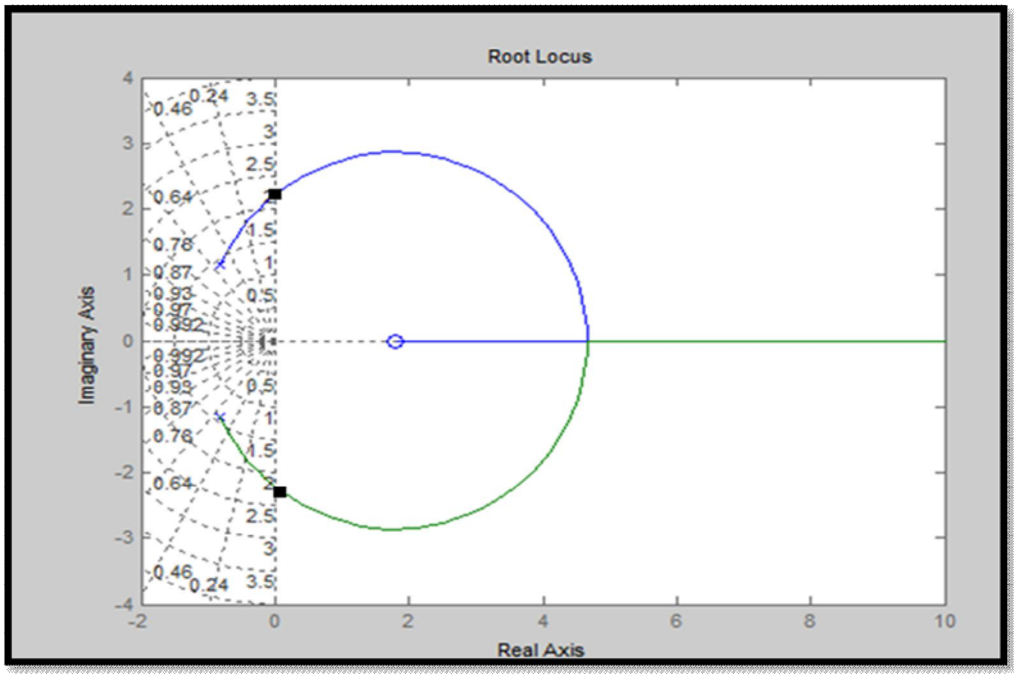


Figure (4.26): root locus grid plot of feedback system for Stripper one

Read from root locus plot figure (4.22)

$W_{co} = 2.2$

$K_u = \ldots\ldots$

$$OLTF = ku\,\frac{(-0.365S+0.657)}{(S^2+1.642S+1.952)}$$

1+OLTF=0

$$1+Ku\,\frac{(-0.365S+0.657)}{(S^2+1.462S+1.982)} = 0$$

- The characteristic equation is:

$$S^2 + (1.642 - 0.365ku)S + (1.982 + 0.657ku) = 0$$

Substitute

S=iw

$(iw)^2 + (1.642\text{-}0.365ku)\ iw + (1.982+0.657ku) = 0$

But

$i^2 = -1$

The equation will be

$- w^2 + (1.642\text{-}0.365ku)iw+(1.982+0.657ku) = 0$

Equating the imaginary part to zero gives

$(1.642 - 0.365\ Ku)\ iw = 0$

$\therefore\ +w^2 = + (1.982+0.657\ Ku)$

But

$\therefore\ w = 2.2$

$\therefore\ Ku = 4.4$

$$Pu = \frac{2\pi}{W_{co}}$$

$$Pu = \frac{2\pi}{2.02}$$

Pu=3.11

From Z-N setting table

parameters	Kc	τ_i	τ_d
PID	0.6ku	$\frac{Pu}{2}$	$\frac{Pu}{8}$

$\therefore\ Kc = 0.6 \times 4.4 = 2.64$

$$\therefore\ \tau_d = \frac{Pu}{8} = \frac{3.11}{8} = 0.39$$

$$\therefore\ \tau_i = \frac{Pu}{2} = \frac{3.11}{2} = 1.56$$

Table (4.6) adjustable at strippers for PID- controller using root locus tuning method

K_U	P_U	K_C	τ_i	τ_d	W_{co}
4.40	3.11	2.64	1.56	0.39	2.02

4.5 Decay ratio tuning method:

The second order system equation is

$\tau^2 S^2 + 2\,\xi\,\tau s+1=0$ ……………………… (4.13)

The characteristic equation for this system is

$S^2 + (1.64 - 0.365k_c)\ S + (1.982+0.657k_u) = 0$

$\therefore\ \tau^2 = 1$

$\therefore\ 2\,\xi\,\tau = 1.64-0.365k_u$

At critical stable

$\xi = 0$

$\therefore\ 1.642-0.365\ k_u = 0$

$\therefore\ K_u = 4.48$

Substitute the value of k_u at equation and put S= iw

$$-(iw)^2 = -(1.982+0.647(4.48))$$

$$\therefore W_{co} = 2.02$$

$$\therefore Pu = \frac{2\pi}{W_{co}} = \frac{2\pi}{2.02} = 3.11$$

From Z_N table:

	kc	τ_i	τ_d
PID	0.6ku	Pu/2	Pu/8

$$\therefore k_c = 4.48 * 0.6 = 2.68$$

$$\therefore \tau_d = \frac{Pu}{8} = \frac{3.11}{8} = 0.39$$

$$\therefore \tau_i = \frac{Pu}{2} = \frac{3.11}{2} = 1.56$$

Table (4.7) adjustable at strippers for PID- controller using decay ration tuning method

K_U	P_U	K_C	τ_i	τ_d	W_{co}
4.48	3.11	2.68	1.56	0.39	2.01

4.6 Comparison between tuning methods

	Kc	Pu	Ku	τ_i	τ_d	offset	Rise time
Z-N tuning method	2.71	2.00	4.5	1.00	0.25	0.00	0.66
Root locus method	2.64	3.11	4.4	1.56	0.39	0.00	1.7

Second order	2.68	3.11	4.48	1.56	0.39	0.00	1.08

As seen from table 4.5, 4.7 and table 4.8, decay ratio and root locus tuning methods both gave very similar results, but Zeigler-Nicholas method showed them the best results: the highest gain with the shortest period.

4.7 Control of Pressure at top of the column

$$G_{(S)} = G_{P(S)} = \frac{0.0001\,S^2 - 0.0008\,S + 0.0017}{S^3 + 1.291\,S^2 + 10.290\,S + 0.0137} \text{.......(**)}$$

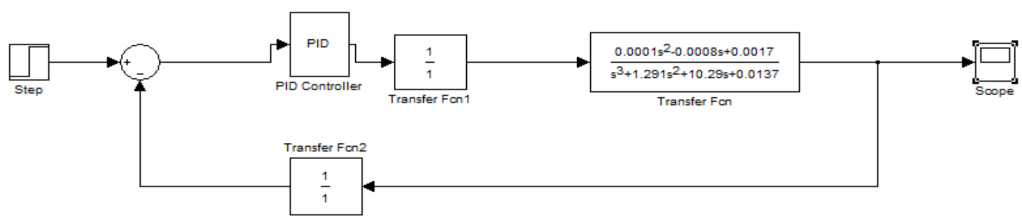


Figure (4.23) block diagram for pressure at top using PID controller

- the trial estimated the ultimate gain is

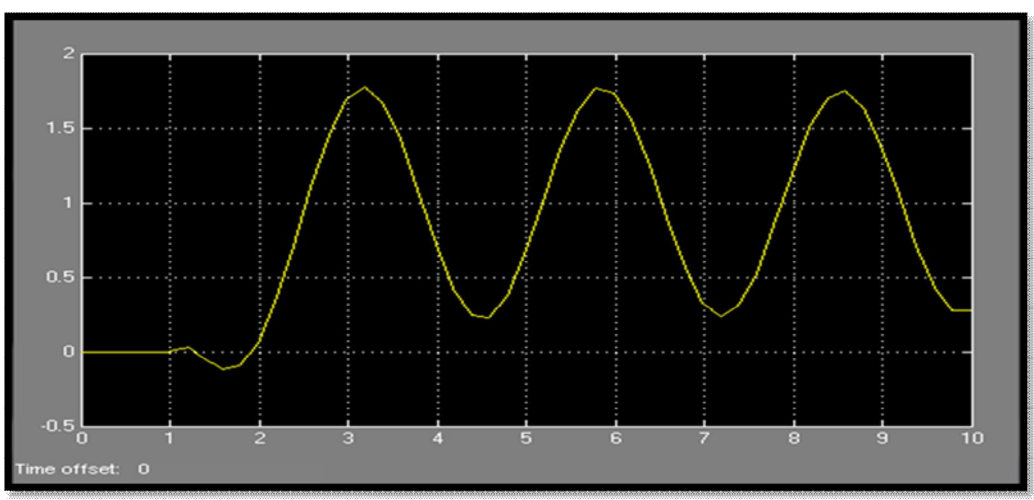


Figure 4.24 **Z-N** tuning plot for PID controller trial

Ku = 6000, system is critical stable.

Ku=6000

Pu=3.00

Table (4.8) adjustable at pressure at top for PID- controller using Z-N tuning method

K_U	P_U	K_C	τ_i	τ_d
6000	3.00	3600	1.5	0.375

$$G_{(S)} = G_{P(S)} = \frac{0.0001\,S^2 - 0.0008\,S + 0.0017}{S^3 + 1.291\,S^2 + 10.290\,S + 0.0137} \text{.......... ......}(**)$$

$$Gc(s) = \frac{2025S^2 + 5400\,S + 3600}{S} \text{......................................}(4.14)$$

- The overall transfer function is

$$G(S) = \frac{0.203S^4 - 1.08S^3 - 0.518S^2 + 6.35S + 6.12}{1.203S^4 + 0.211\,S^3 + 9.773S^2 + 6.31S + 6.12} \text{..........}(4.15)$$

Unit step change test using PID- controller to measure the swiftness of the system response

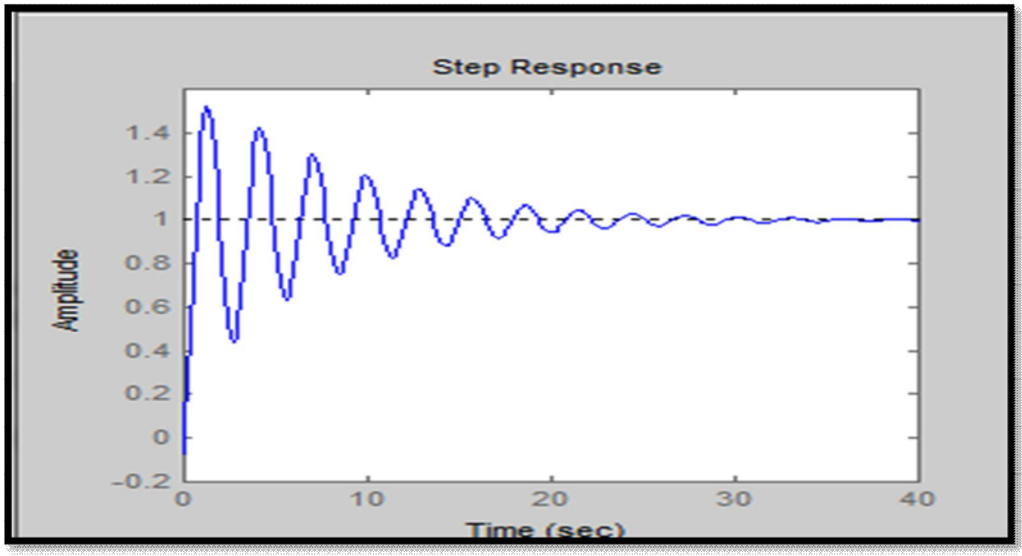


Figure (4.25) step response for PID controller

4.8 Control of level at bottom of the column

$$Gp(s) = \frac{0.08525 + 0.0266}{s^2 - 0.08825s + 0.1023} \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots . (***)$$

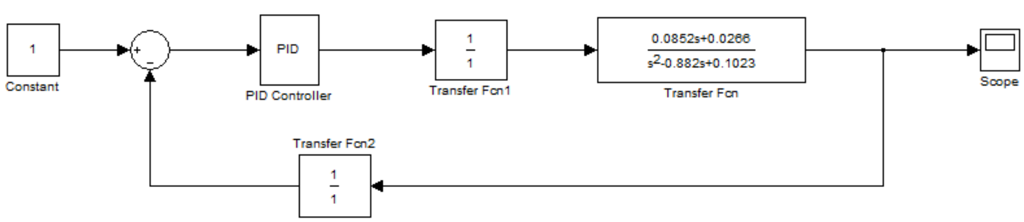


Figure (4.26) block diagram for level at bottom using PID controller

- the trial estimated the ultimate gain is

Ku = 0.02, system is critical stable.

Ku=0.02, Pu=6.00

Table (4.9) adjustable for level at bottom for PID- controller using Z-N tuning method

K_U	P_U	K_C	τ_i	τ_d
8.3	6.00	4.98	3.00	0.75

$$Gp(s) = \frac{0.08525 + 0.0266}{S^2 - 0.08825S + 0.1023} \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots. (***)$$

$$Gc(s) = \frac{2.25S^2 + 14.94\,S + 4.98}{S} \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots. (4.16)$$

- The overall transfer function is

$$G(S) = \frac{0.192S^3 + 1.336S^2 + 0.819S + 0.132}{1.192\,S^3 + 1.248S^2 + 0.921S + 0.132} \ldots\ldots. (4.17)$$

- Unit step change test using PID- controller to measure the swiftness of the system response:

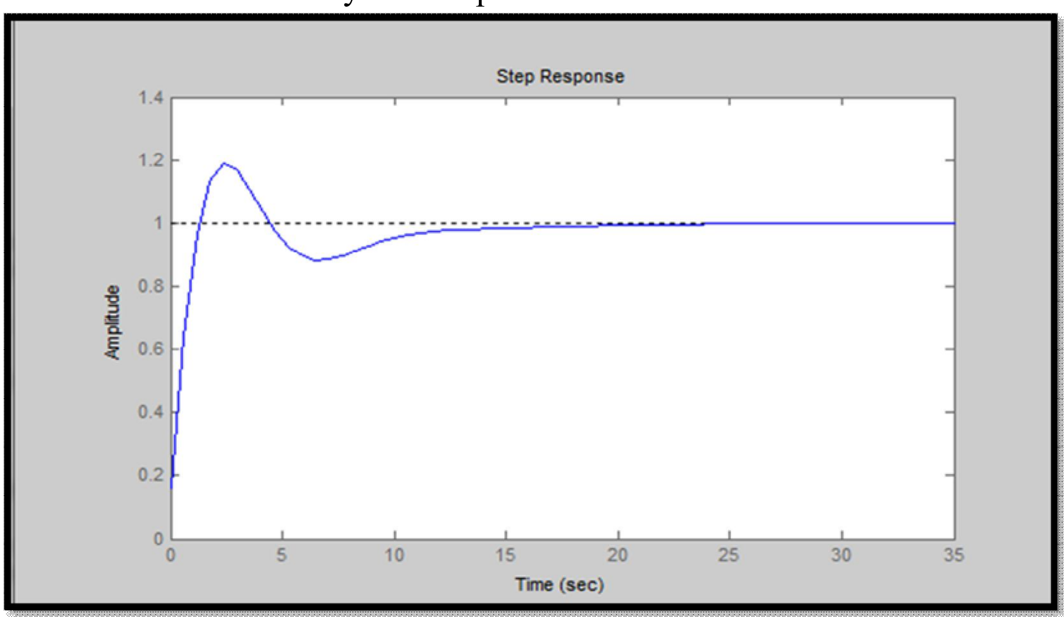


Figure (4.28) step response for PID controller of level at bottom

4.9 Control of furnace temp

$$Gp(s) = \frac{0.0004s^2 - 0.0042S + 0.0059}{s^3 + 0.825s^2 + 10.04s + 0.028} \ldots\ldots\ldots\ldots\ldots\ldots\ldots(****)$$

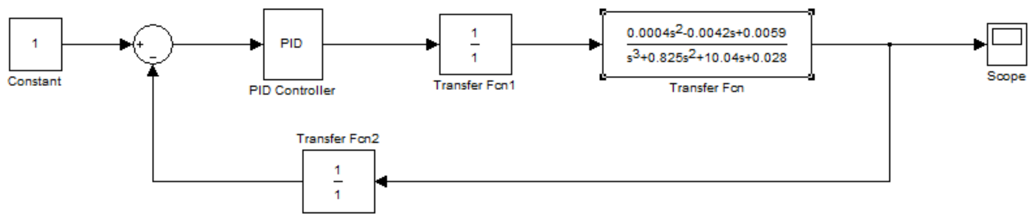


Figure (4.29) block diagram for the furnace using PID controller

- the trial estimated the ultimate gain is

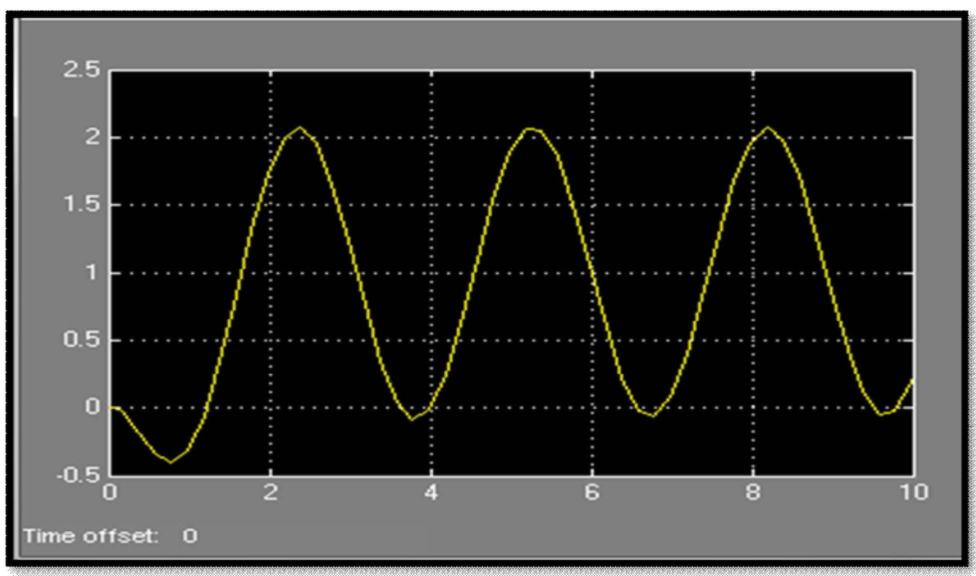


Figure (4.30) **Z-N** tuning plot for PID controller trial

Ku = 1420, system is critical stable.

Ku=1420

Pu=3.00

Table (4.11) adjustable for level at bottom for PID- controller using Z-N tuning method

K_U	P_U	K_C	τ_i	τ_d
1420	3.00	852	1.5	0.375

$$Gp(s) = \frac{0.0004S^2 - 0.0042S + 0.0059}{S^3 + 0.825S^2 + 10.04S + 0.028} \ldots\ldots\ldots\ldots\ldots\ldots (****)$$

$$Gc(s) = \frac{319.5S^2 + 1278\,S + 825}{S} \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (4.18)$$

- The overall transfer function is

$$G(S) = \frac{0.127S^4 - 0.767S^3 - 2.889S^2 + 4.132S + 5.02}{1.128S^4 + 0.058\,S^3 + 7.16S^2 + 4.16S + 5.02} \ldots\ldots (4.19)$$

4.10 The stability criteria

4.10.1 Bode Criterion

A feedback control system is stable according to bode criterion if the amplitude ratio of the corresponding open-loop transfer function is less than one at the crossover frequency.

The crossover frequency is the frequency where the phase lag is equal to (- 180^0). [2]

4.10.2 Nyguist stability criterion

It states that if the open-loop Nyguist plot of feedback system encircles the point (- 1.0) as the frequency w takes any value from (-∞ to + ∞). [2]

4.10.3 Root locus criterion

This criterion of stability does not require calculations of the actual

values of the roots of the characteristic equation, it is only requires that we know if any root is to the right of the imaging axis (i.e., have positive real part) that leads to unstable system. [2]

4.10.1 Bode Criterion:

1- Level at stripper one control stability test using Bode criterion

The bode plot of the feedback transfer function is plotted using MATLAB software

```
>> num=[-0.245 -0.546 0.785 1.477]

num =

   -0.2450   -0.5460    0.7850    1.4770

>> den=[0.755 1.094 2.6855 1.774]

den =

    0.7550    1.0940    2.6855    1.7740
```

```
>> sys=tf(num,den)

Transfer function:
-0.245 s^3 - 0.546 s^2 + 0.785 s + 1.477
----------------------------------------
0.755 s^3 + 1.094 s^2 + 2.686 s + 1.774

>> bode(sys)
```

Results shown in figure (23) below:

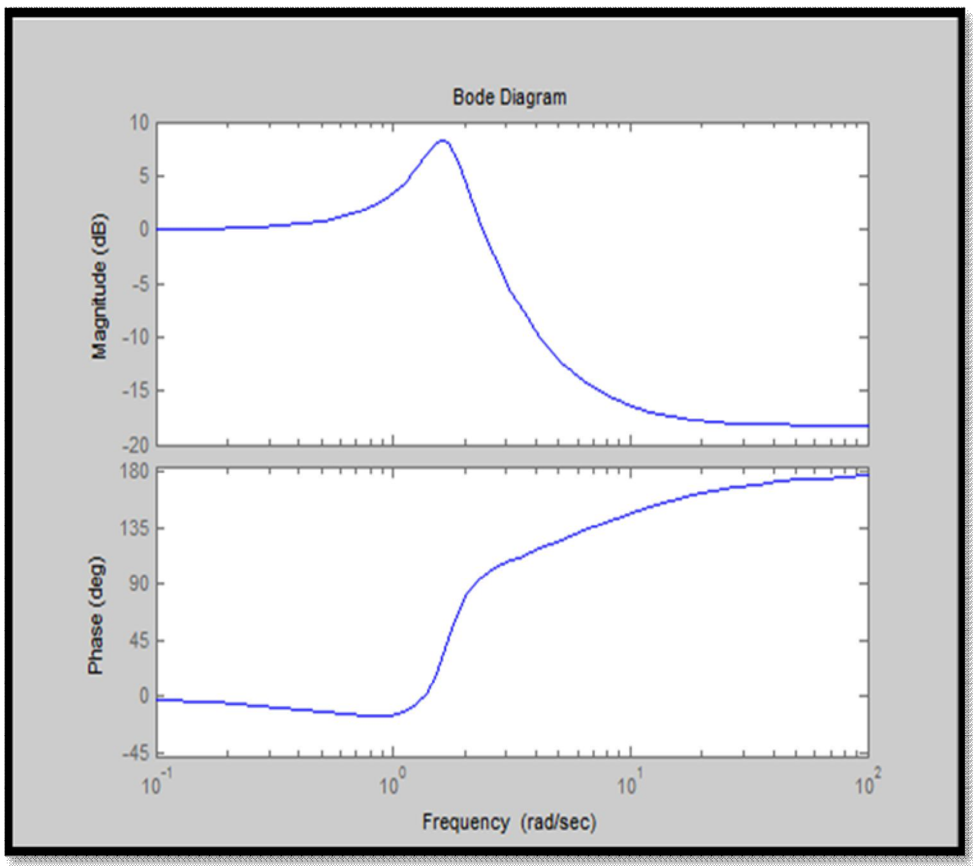


Figure (4.31) Bode plot of feedback control system for stripper one

Since the Bode phase plot does not reach (-180˚), the system is stable according to Bode criterion.

2- Pressure at top of the column control stability test using Bode criterion

The bode plot of the feedback transfer function is plotted using MATLAB software

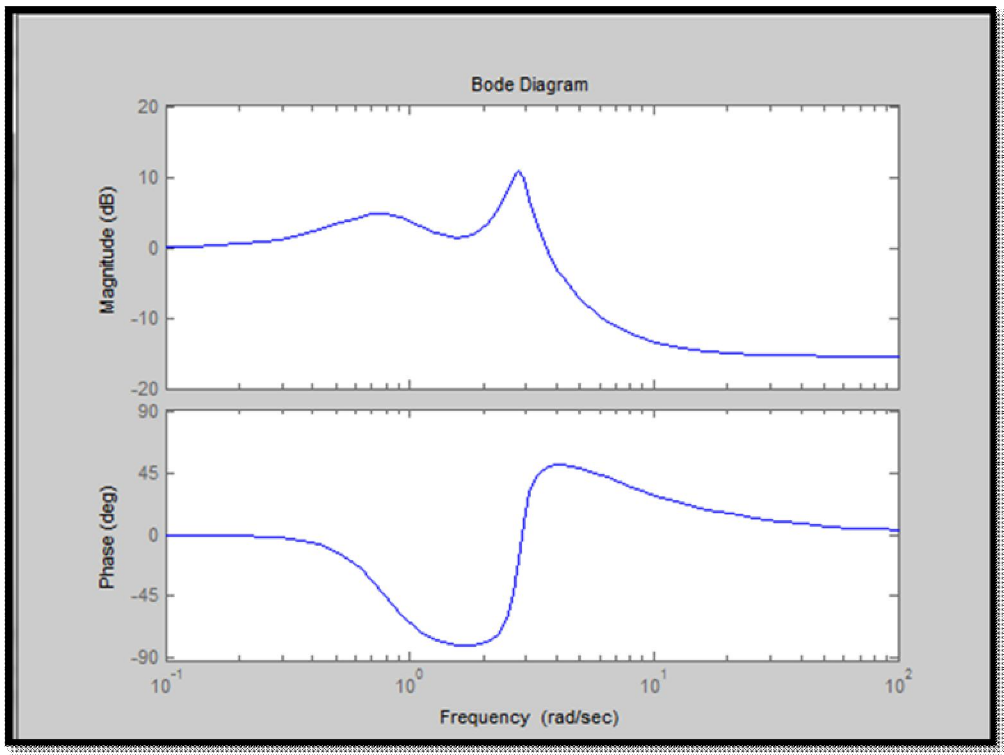


Figure (4.32) Bode plot of feedback control system for pressure at top

Since the Bode phase plot does not reach (-180˚), the system is stable according to Bode criterion.

3- Level at the bottom of the column control stability test using Bode criterion.

The bode plot of the feedback transfer function is plotted using MATLAB software

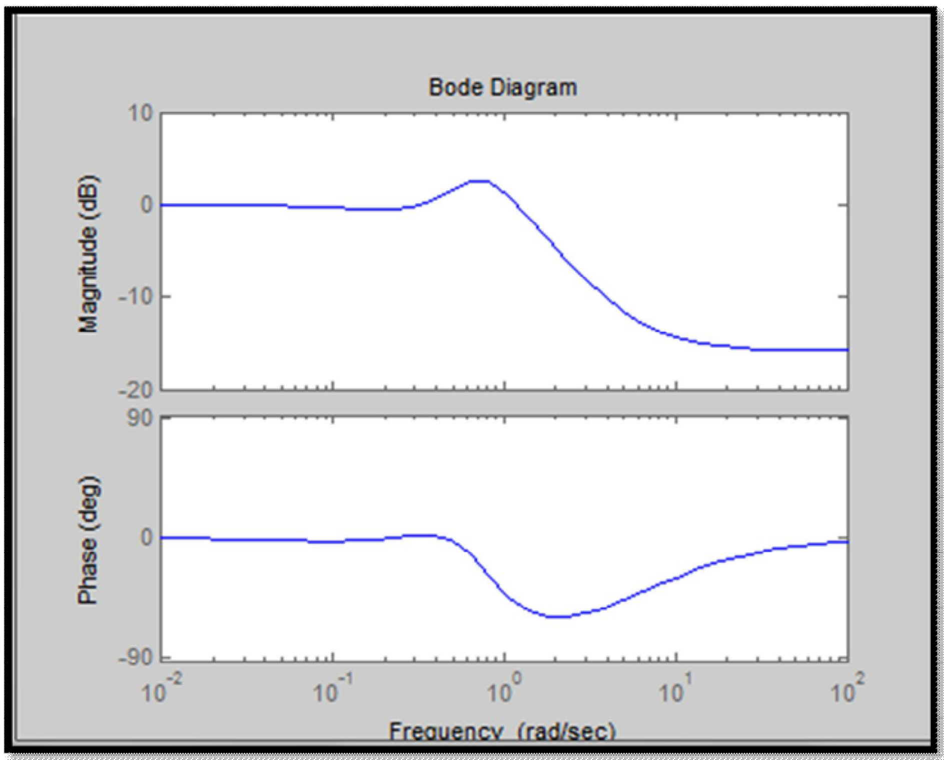


Figure (4.33) Bode plot of feedback control system for the level at bottom

Since the Bode phase plot does not reach (-180˚), the system is stable according to Bode criterion.

4- Furnace temp control stability test using Bode criterion.

The bode plot of the feedback transfer function is plotted using MATLAB software

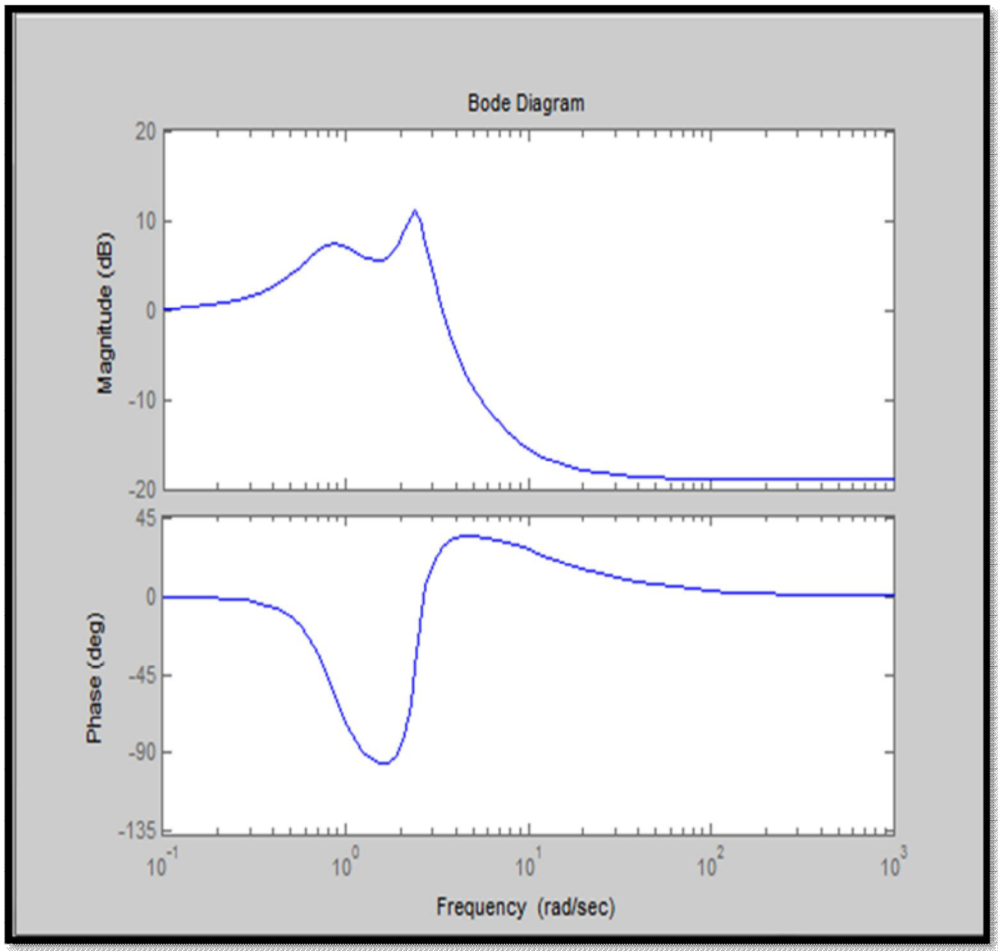


Figure (4.34) Bode plot of feedback control system for furnace

Since the Bode phase plot does not reach (-180˚), the system is stable according to Bode criterion.

4.10.2 Nyguist stability criterion

1- Level at stripper one control stability test using Nyquist criterion

The Nyquist plot of the feedback transfer function is plotted using MATLAB as shown in figure (4.35) below:

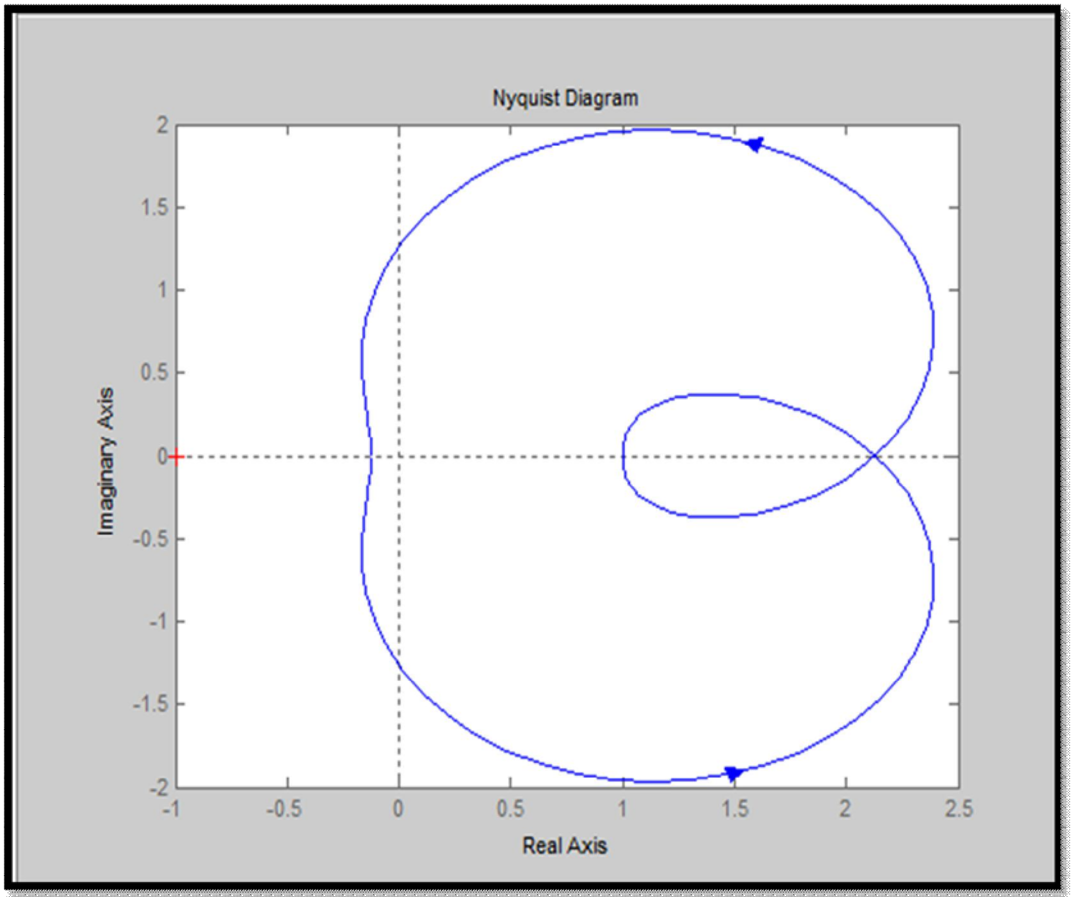


Figure (4.35) Nyquist plot of the feedback control system for

Stripper one

The Nyquist plot of the feedback control system does not encircle the point (-1, 0) therefore, the system is stable according to Nyquist criterion.

2-Pressure at top control stability test using Nyquist criterion

The Nyquist plot of the feedback transfer function is plotted using MATLAB as shown in figure (4.36) below:

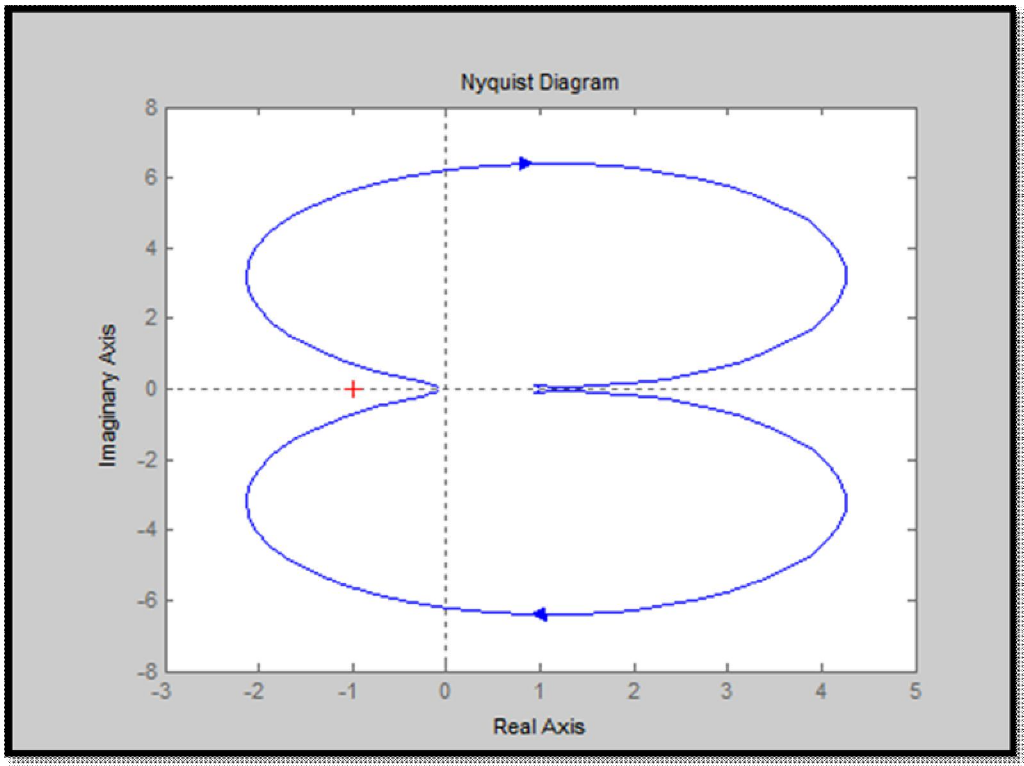


Figure (36) Nyquist plot of the feedback control system for

Pressure at top

The Nyquist plot of the feedback control system does not encircle the point (-1, 0) therefore, the system is stable according to Nyquist criterion.

3-Level at bottom control stability test using Nyquist criterion

The Nyquist plot of the feedback transfer function is plotted using MATLAB as shown in figure (4.37) below:

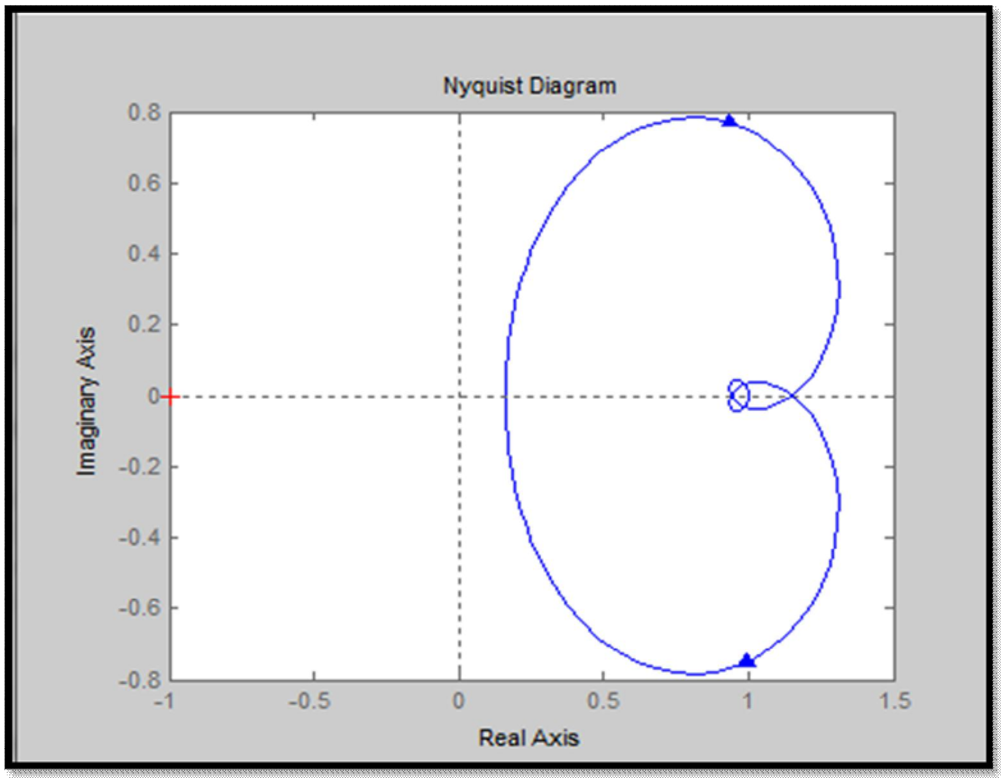


Figure (4.37) Nyquist plot of the feedback control system for

Level at bottom

The Nyquist plot of the feedback control system does not encircle the point (-1, 0) therefore, the system is stable according to Nyquist criterion.

4-Furnace temp control stability test using Nyquist criterion

The Nyquist plot of the feedback transfer function is plotted using MATLAB as shown in figure (4.38) below:

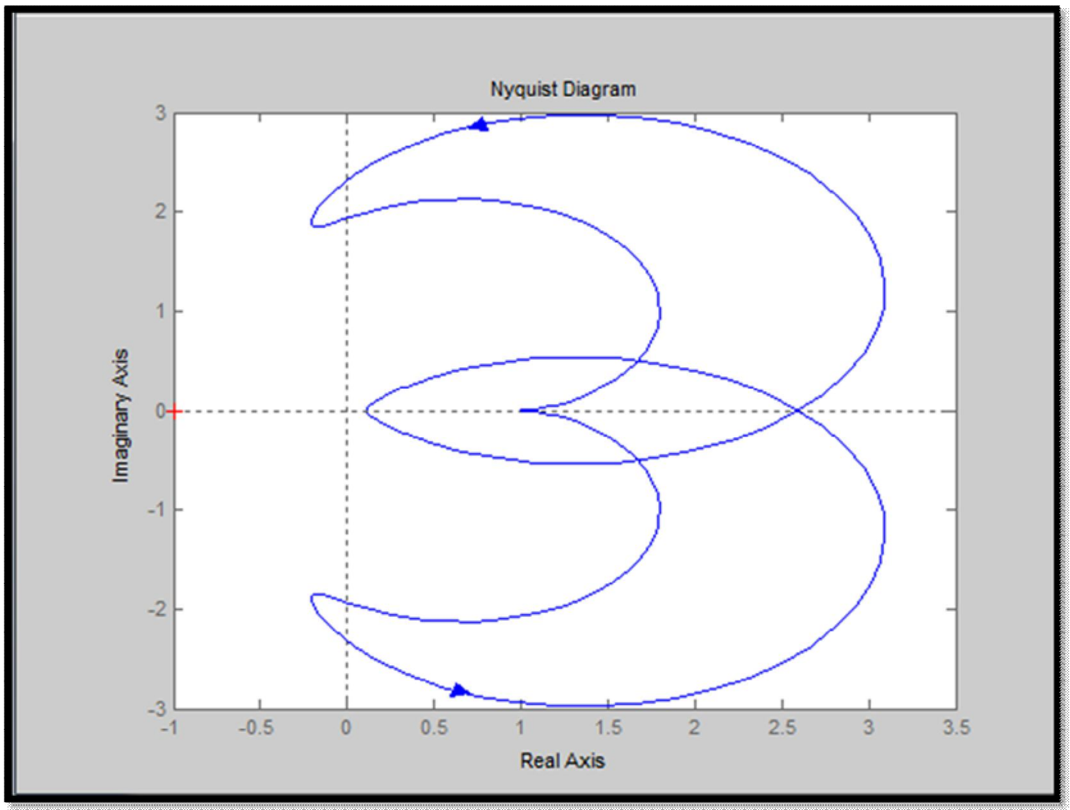


Figure (4.38) Nyquist plot of the feedback control system for

The furnace

The Nyquist plot of the feedback control system does not encircle the point (-1, 0) therefore, the system is stable according to Nyquist criterion.

4.10.3 Root locus criterion

1- Level at stripper one control stability test using root locus criterion

Root locus plot of the feedback transfer function is plotted using MATLAB as shown in figure (4.39) below:

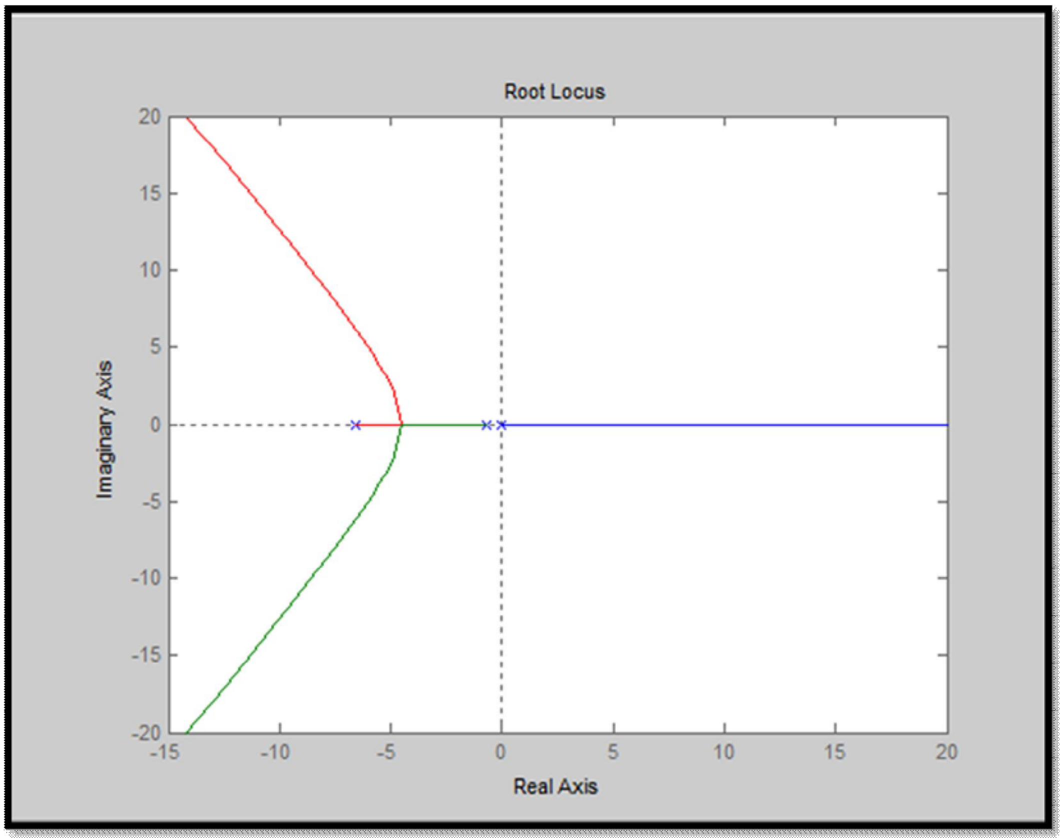


Figure (4.39) Root locus plot of the feedback control system for stripper one

The feedback system cannot go unstable, since the roots of the characteristic equation cannot have positive real parts (*i.e.,* root locus plot does not cross the imaginary axis).

2- Pressure at top control stability test using root locus criterion

Root locus plot of the feedback transfer function is plotted using MATLAB as shown in figure (4.40) below:

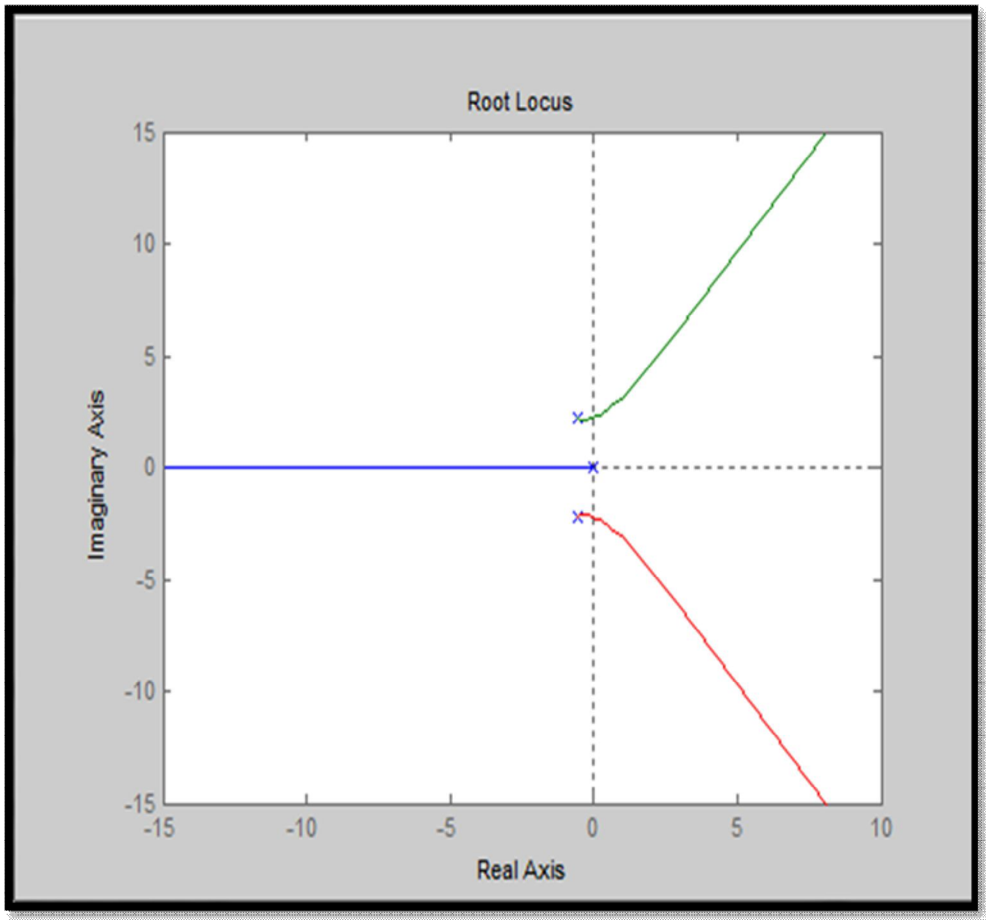


Figure (4.40) Root locus plot of the feedback control system for pressure at top

The feedback system cannot go unstable, since the roots of the characteristic equation cannot have positive real parts (*i.e.,* root locus plot does not cross the imaginary axis).

3- Level at bottom control stability test using root locus criterion

Root locus plot of the feedback transfer function is plotted using MATLAB as shown in figure (4.41) below:

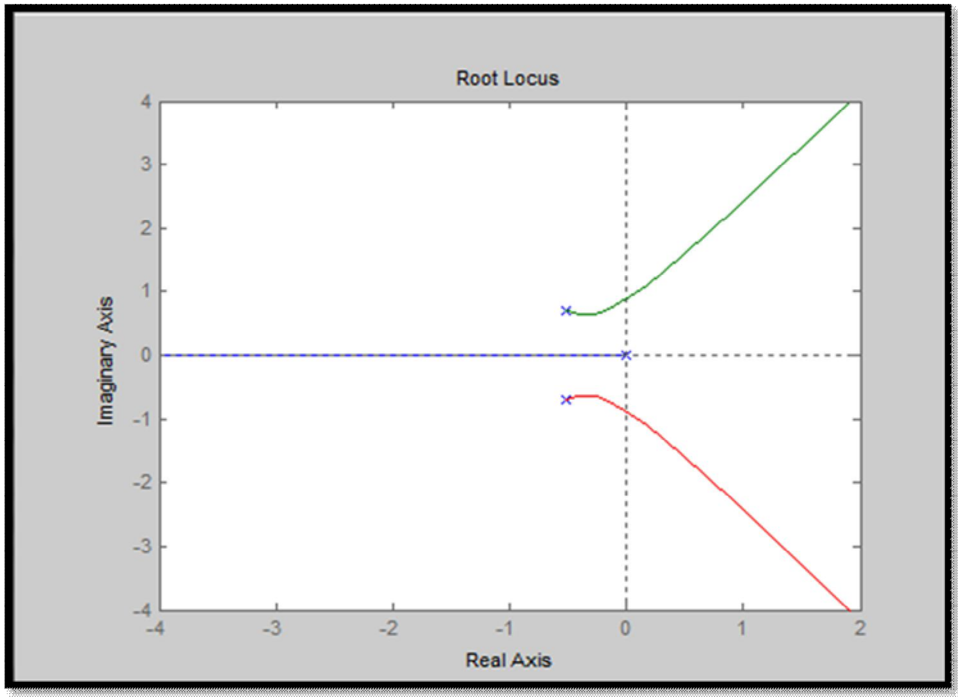


Figure (4.41) Root locus plot of the feedback control system for level at bottom.

The feedback system cannot go unstable, since the roots of the characteristic equation cannot have positive real parts (*i.e.,* root locus plot does not cross the imaginary axis).

4- Furnace temp control stability test using root locus criterion

Root locus plot of the feedback transfer function is plotted using MATLAB as shown in figure (4.42) below:

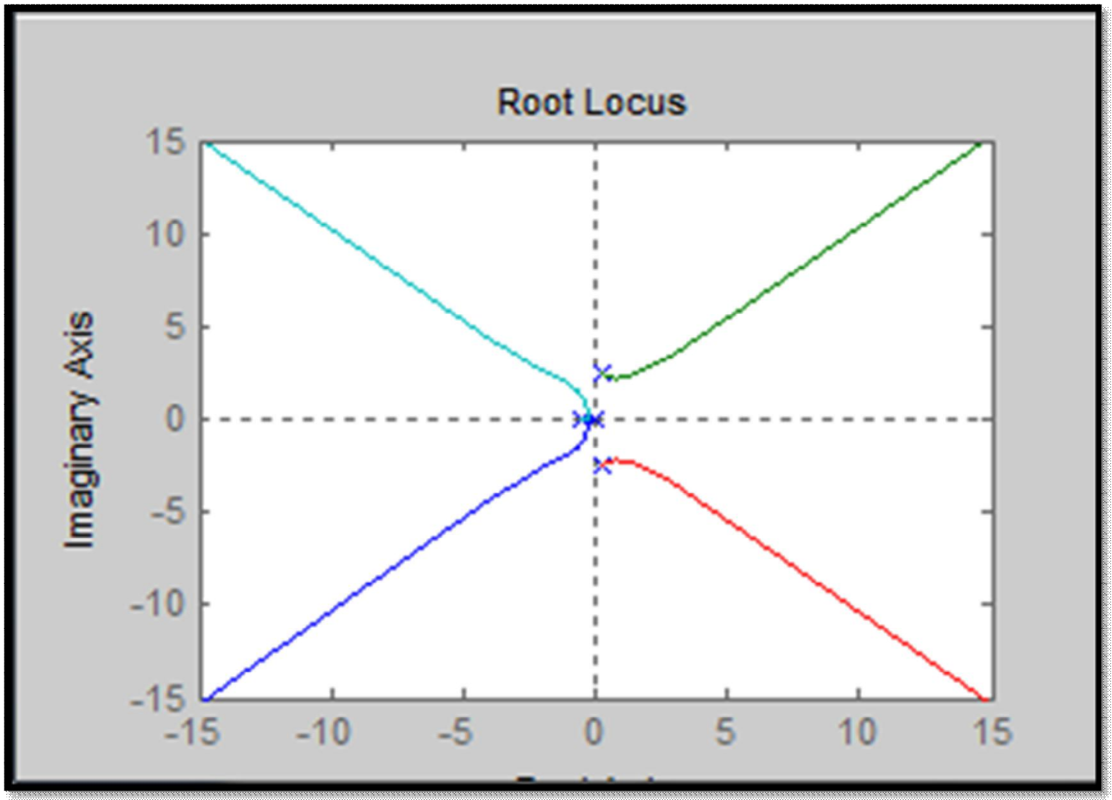


Figure (4.42) Root locus plot of the feedback control system furnace temp

The feedback system can go unstable, since the roots of the characteristic equation can have positive real parts.

Stability Criteria Results

Table 4.12: Stability Criteria Results

CRITERION	**Bobe**	**Nyquist**	**Root locus**
Control of level at stripper one	Stable	Stable	Stable
Control of pressure at top of the column	Stable	Stable	Stable
Control of level at bottom of the column	Stable	Stable	Stable
Control of oil out from furnace temp	Stable	Stable	Un stable

4.11 Digital control

1- Level at stripper one digital control system

The overall transfer function

$$G(s) = \frac{-0.245S^3 - 0.546S^2 + 0.78\,S + 1.774}{0.755\,S^3 + 1.094S^2 + 2.685S + 1.774} \quad \text{..............................} \quad (4.11)$$

- **The step response for this transfer function is as follow:**

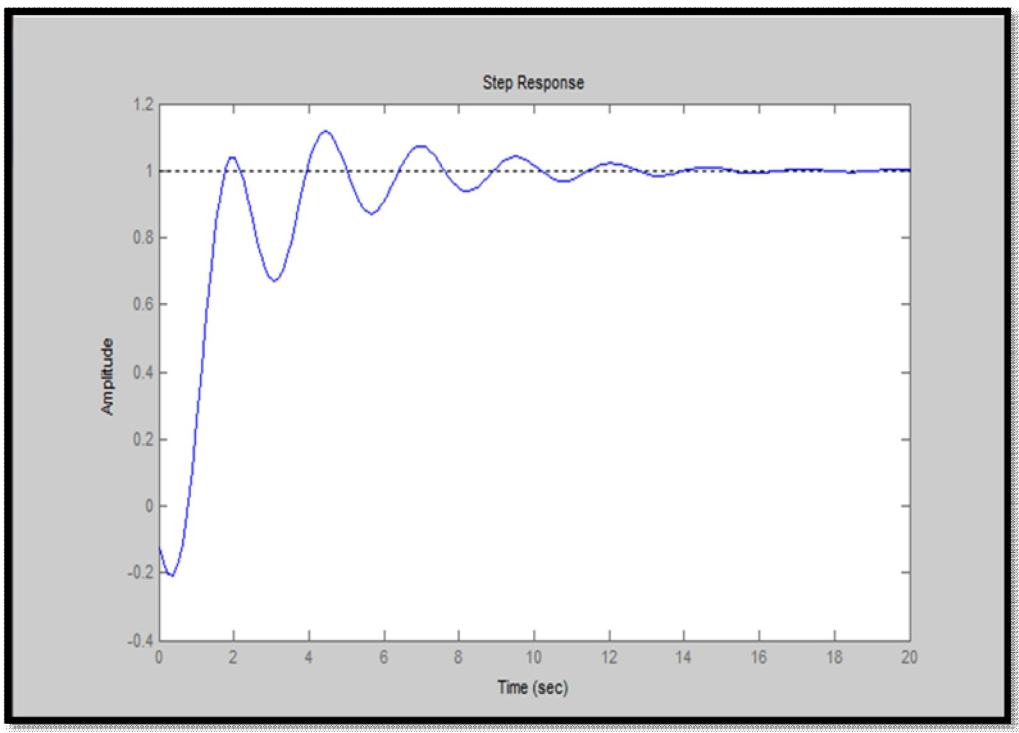


Figure (4.15): continuous response due to step change for overall feedback system

Sampling time for Analog/Digital converter can be estimated as 10% process time constant. Then, the sampling time will be used is 0.1 and the following command have been entered to MATLAB:

```
>> num=[-0.245 -0.546 0.78 1.774]

num =

   -0.2450   -0.5460    0.7800    1.7740

>> den=[0.755 1.094 2.685 1.774]

den =

    0.7550    1.0940    2.6850    1.7740

>> sys=tf(num,den)

Transfer function:
-0.245 s^3 - 0.546 s^2 + 0.78 s + 1.774
---------------------------------------
0.755 s^3 + 1.094 s^2 + 2.685 s + 1.774

>> zsys=c2d(sys.01,'zoh')
??? zsys=c2d(sys.01,'zoh')
                |
Error: Unexpected MATLAB expression.

>> zsys=c2d(sys,0.1,'zoh')

Transfer function:
-0.3245 z^3 + 0.9061 z^2 - 0.8273 z + 0.2479
--------------------------------------------
     z^3 - 2.831 z^2 + 2.698 z - 0.8651

Sampling time: 0.1
```

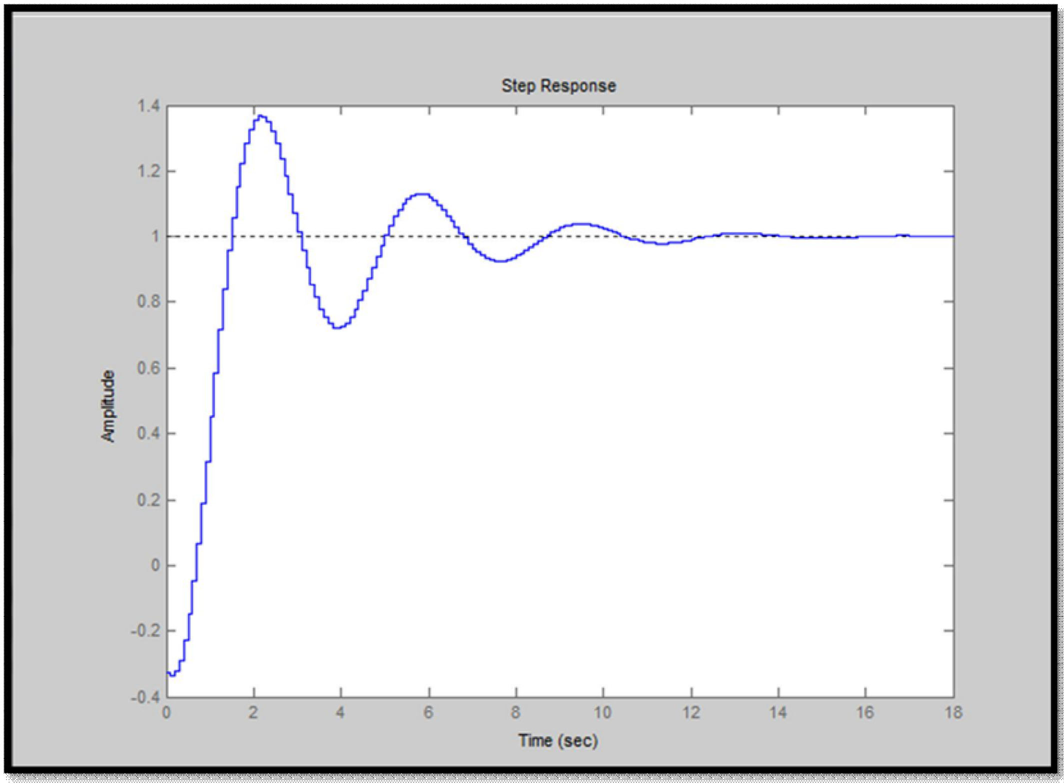


Figure (4.43): discrete response due to step change for overall feedback system for stripper one

2- Pressure at top digital control system

Sampling time for Analog/Digital converter can be estimated as 10% process time constant. Then, the sampling time will be used is 0.1 and the following command have been entered to MATLAB

```
num =

   -0.0670    0.6250    3.3480    3.6160

>> den=[0.933 1.004 4.617 3.616]

den =

    0.9330    1.0040    4.6170    3.6160

>> sys4=tf(num,den)

Transfer function:
-0.067 s^3 + 0.625 s^2 + 3.348 s + 3.616
----------------------------------------
0.933 s^3 + 1.004 s^2 + 4.617 s + 3.616


>> zsys4=c2d(sys4,0.1,'zoh')

Transfer function:
-0.07181 z^3 + 0.2945 z^2 - 0.3361 z + 0.1171
---------------------------------------------
      z^3 - 2.849 z^2 + 2.751 z - 0.898

Sampling time: 0.1
```

- **Step change for the transfer function at z-domain:**

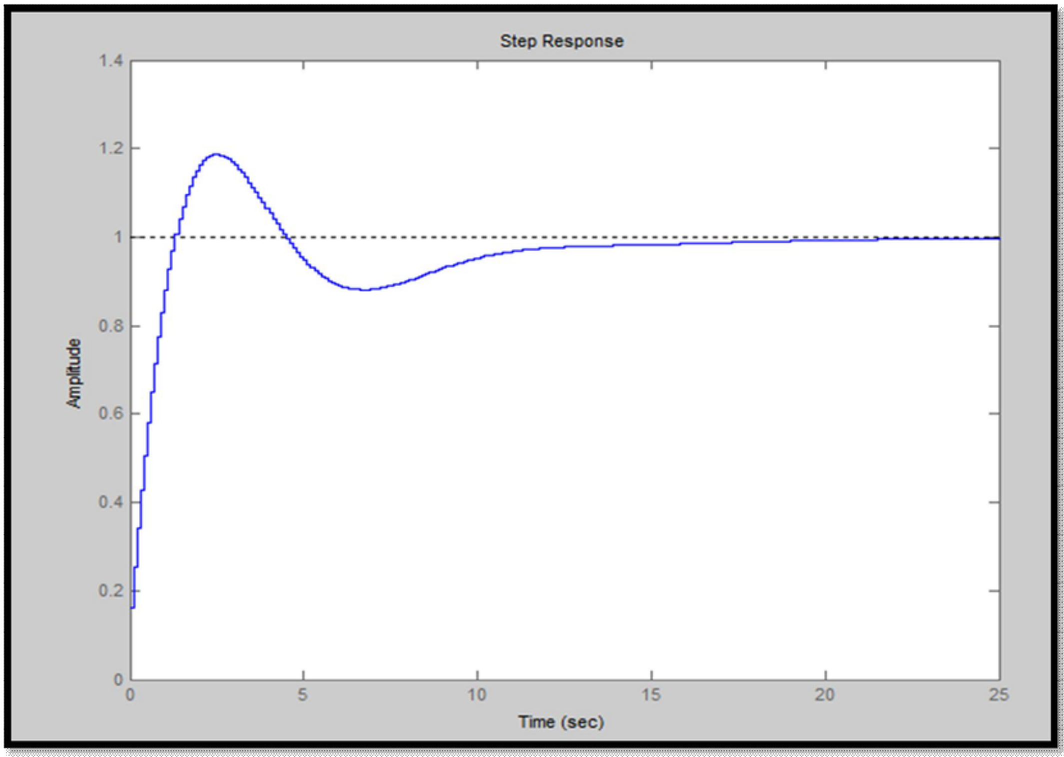


Figure (4.44): discrete response due to step change for overall feedback system for pressure at top.

3- Level at bottom digital control system

Sampling time for Analog/Digital converter can be estimated as 10% process time constant. Then, the sampling time will be used is 0.1 and the following command have been entered to MATLAB

```
>> num=[0.192 1.336 0.819 0.132]

num =

    0.1920    1.3360    0.8190    0.1320

>> den=[1.192 1.248 0.921 .132]

den =

    1.1920    1.2480    0.9210    0.1320

>> sys8=tf(n,den)
??? Undefined function or variable 'n'.

>> sys8=tf(num,den)

Transfer function:
0.192 s^3 + 1.336 s^2 + 0.819 s + 0.132
---------------------------------------
1.192 s^3 + 1.248 s^2 + 0.921 s + 0.132

>> zsys8=c2d(sys8,0.1,'zoh')


Transfer function:
0.1611 z^3 - 0.373 z^2 + 0.2694 z - 0.05739
-------------------------------------------
    z^3 - 2.893 z^2 + 2.794 z - 0.9006


Sampling time: 0.1
```

- **Step change for the transfer function at z-domain:**

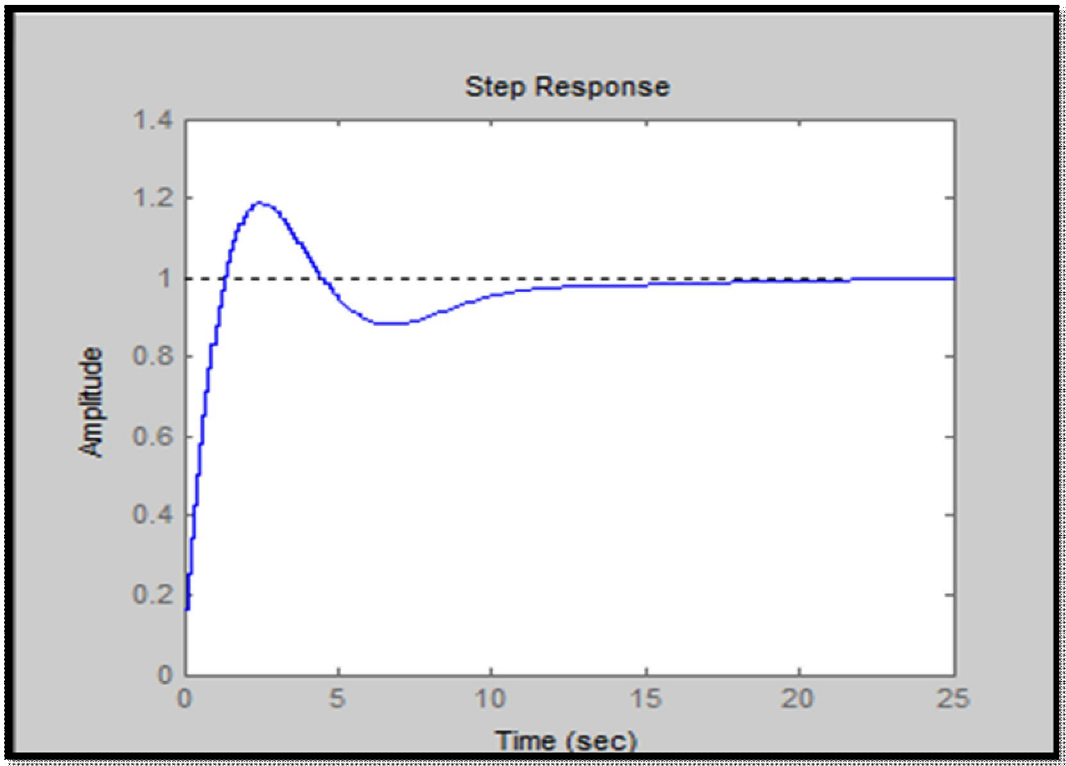


Figure (4.45): discrete response due to step change for overall feedback system for level at bottom.

CHAPTER FIVE

CONCLUSION AND RECOMMENDATION

Chapter five

Conclusion and Recommendations

5.1 Conclusion

The fractionation unit in Khartoum Refinery is considered as one of the important unit for fractionating crude oil. The Graphical User Interphase is used in identification and analysis of the control system. The dynamic performance of the system is investigated, the selection of the best mode of controller is selected (PID-Controller). Different tuning methods were used. These are: Ziegler, Root Locus and decay ratio.

Different stability criterions; Bode, Nyguist and Root locus were applied and all control systems were found to be stable except furnace control system with root locus method was unstable.

Implementing advanced process control and optimization on a refinery plant is quite a difficult task but the results could be remarkable: energy savings maximized throughput and improved yields while increasing the overall profit of the refinery.

The furnace temperature rises significantly and needs to be adjusted and controlled by automatic control.

5.2 Recommendations

Further work has to be done on the following:

1- Design of a digital control system for the furnace to keep the inputs and out puts in their desired variables.
2- Implementation of the system investigated in this thesis.
3- Application of override control is essential for controlling the steam rate at the bottom of the column.

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Appendixes

- Chapter three
- Appendix (3.0)

Type of sensors

Table*: Typical measuring devices for process control

Measuring process variable	Measuring device	Comments
Temperature	Thermocouple Resistance thermometer Filled-system thermometers Radiation pyrometer Oscillating quartz crystal	Most comment for relatively law temperature Used for high temperature
Pressure	Manometers Bourdon-tube element Bellows elements Diaphragm element Strain gages Piezoresistivity elements Piezoelectric elements	With float or displacer Based on the elastic deformation of materials Used to convert pressure to

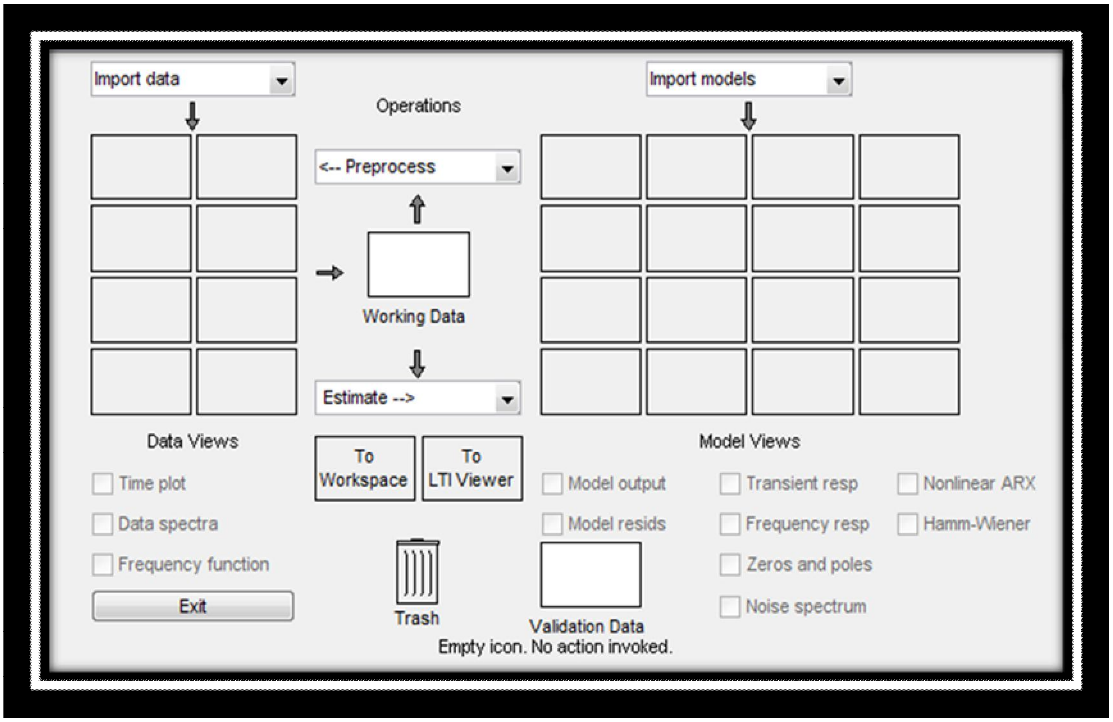
		electrical signal
Flow	Orifice plates Ventura flow nozzle Dahl flow tube Dennison flow nozzle Turbine flow meters Ultrasound Hot wire anemometry	Measuring pressure drop across a flow constriction For high precision
Liquid level	Float-actuated device Displacer device Conductivity measurements Dielectric measurements Liquid head pressure devices Sonic resonance	Coupled with various types of indicators and signal convectors Good for systems with two phases

Composition	Chromatographic analyzer Infrared analyzers Ultraviolet analyzers Visible-radiation analyze Turbidimetry analyzer Paramagnetism analyzer Nephlometery analyzer Potentiometry Conductmetery Oscillometeric analyzers pH meter polo graphic analyzers Coulometer Spectrometers (x-ray, electron, ion, Mossbauer, Raman etc) Different thermal analyzes	Long time required for analysis Convenient for one or two chemicals Not very convenient for process control Expensive for low cost control loops
Concentration	Electro-pnumeter Gas-liquid chromatography	

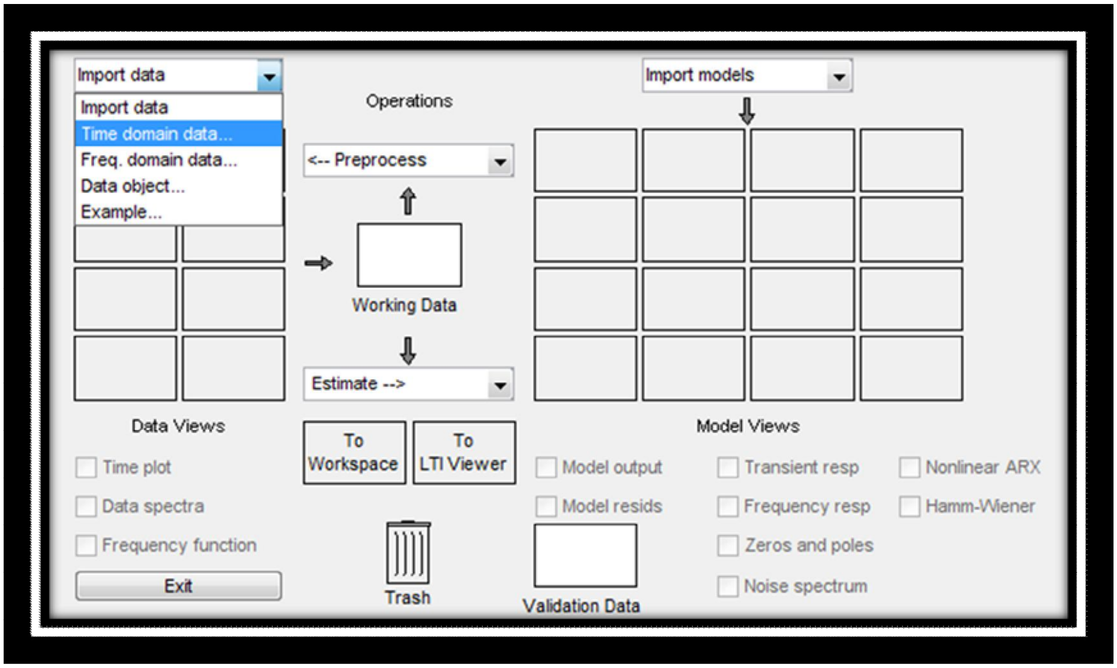
Appendix (3.A)

Steps to estimate transfer functions for CDU unit in KRC.

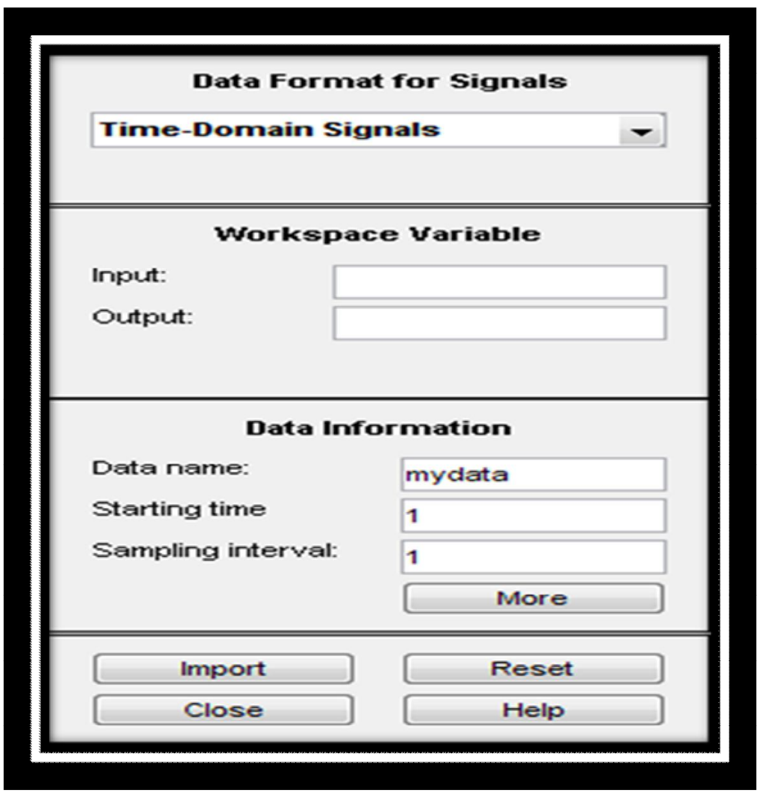
1- Command window
2- ident



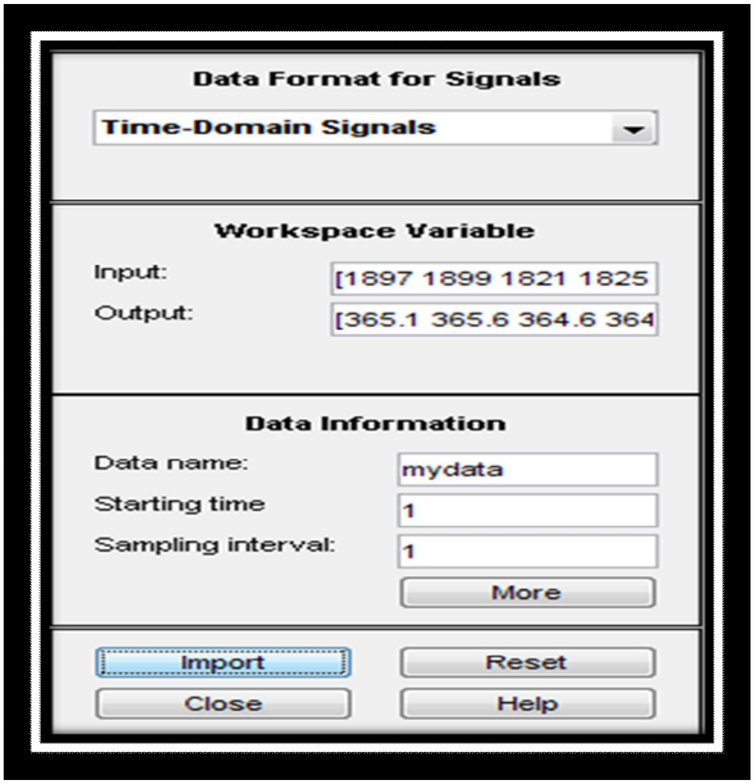
3- import data in table (1) in three steps; which appearing in the following 3 below boxes:



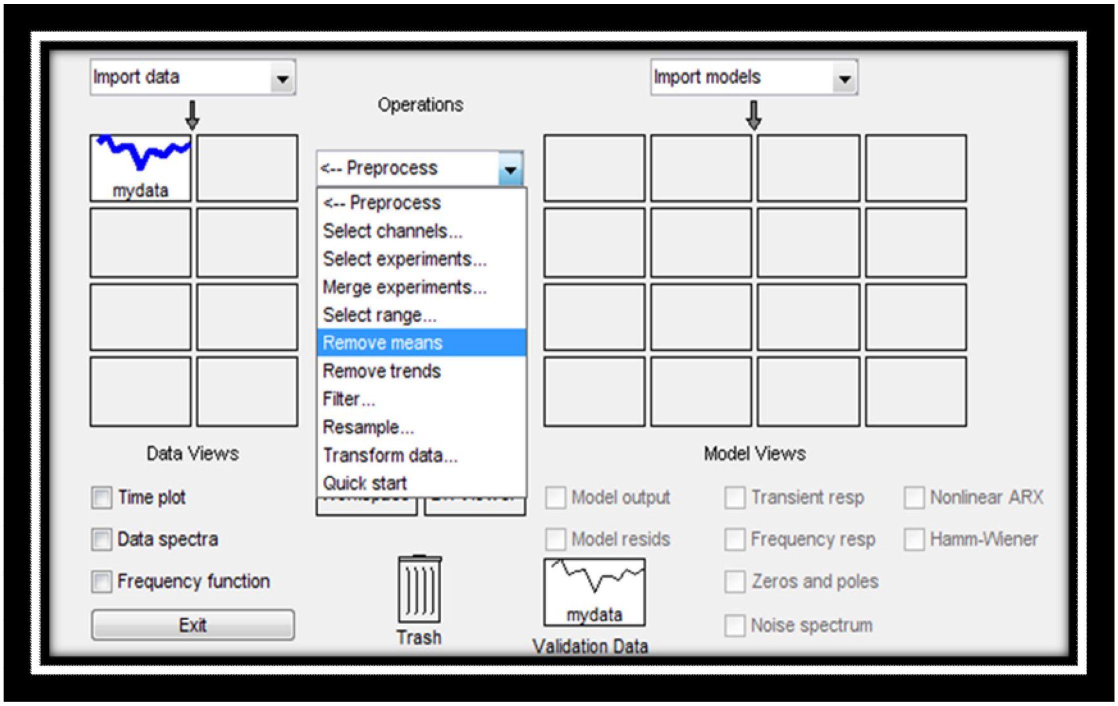
Import data
Import data
Time domain data...
Freq. domain data...
Data object...
Example...
Operations
<-- Preprocess
Working Data
Estimate -->
Import models
Data Views
Time plot
Data spectra
Frequency function
Exit
To Workspace
To LTI Viewer
Trash
Validation Data
Model Views
Model output
Model resids
Transient resp
Frequency resp
Zeros and poles
Noise spectrum
Nonlinear ARX
Hamm-Wiener

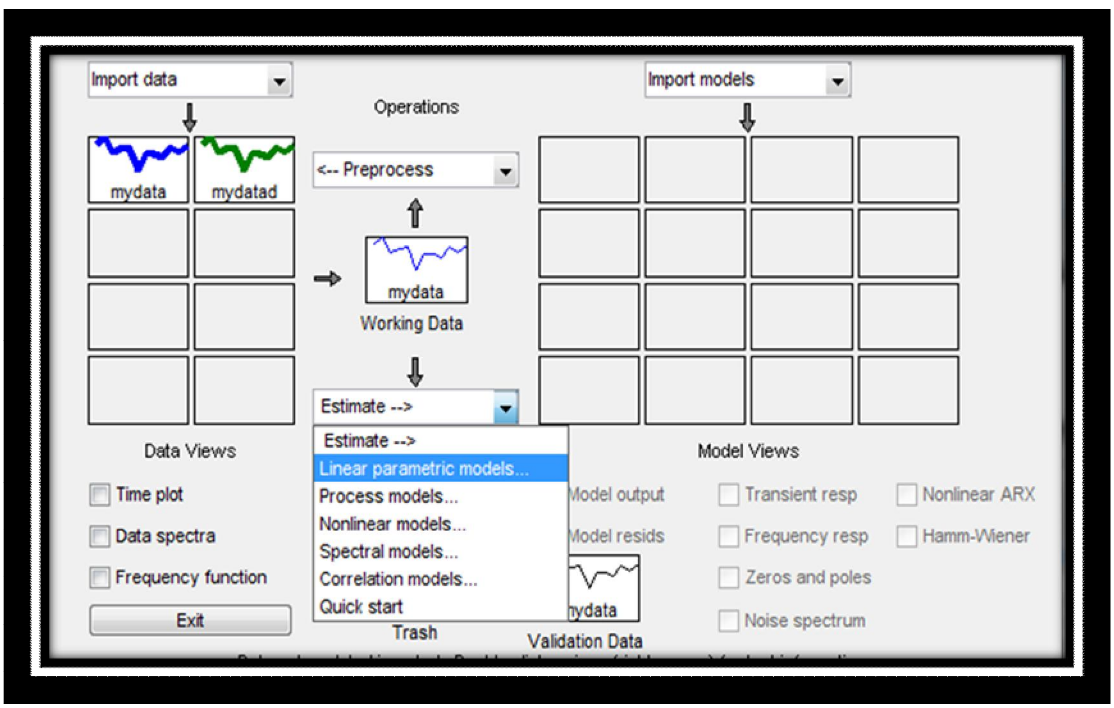


Data Format for Signals
Time-Domain Signals
Workspace Variable
Input:
Output:
Data Information
Data name:
mydata
Starting time
1
Sampling interval:
1
More
Import
Reset
Close
Help

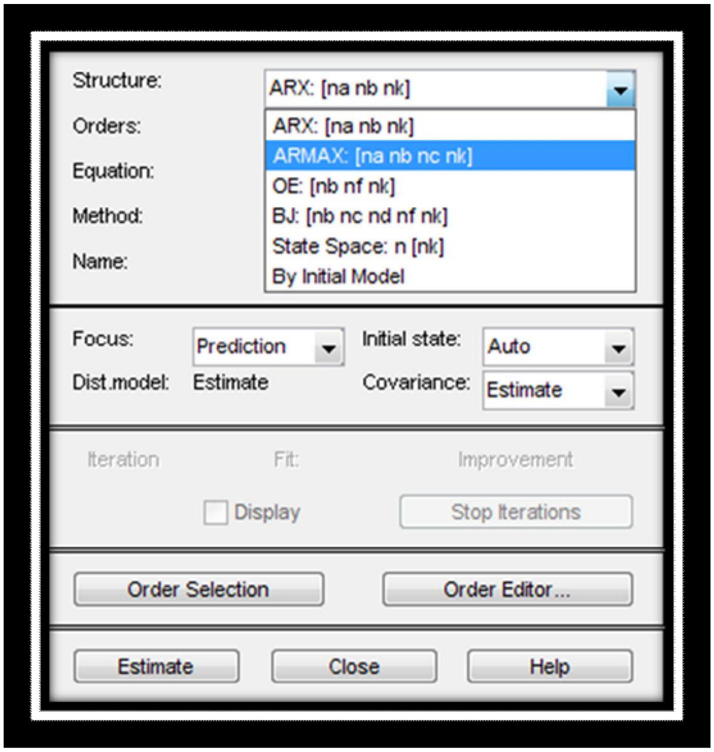


3- Drag the model to working data: choose remove means from operations, then choose linear parametric model from estimate. See below boxes:

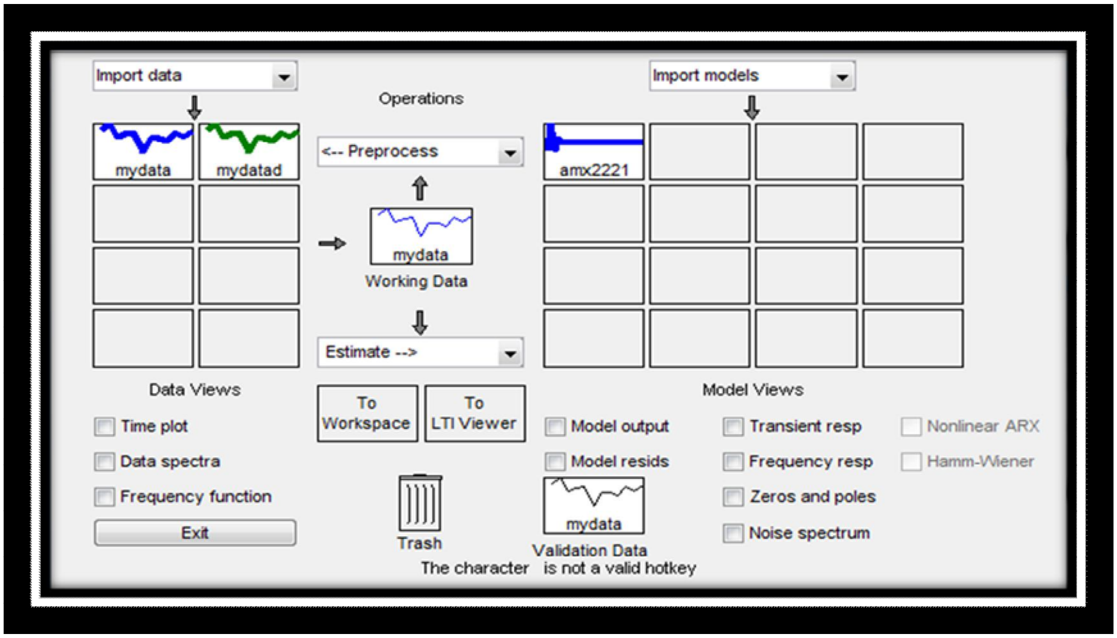




4- Choose ARMAX structure model and press estimate:

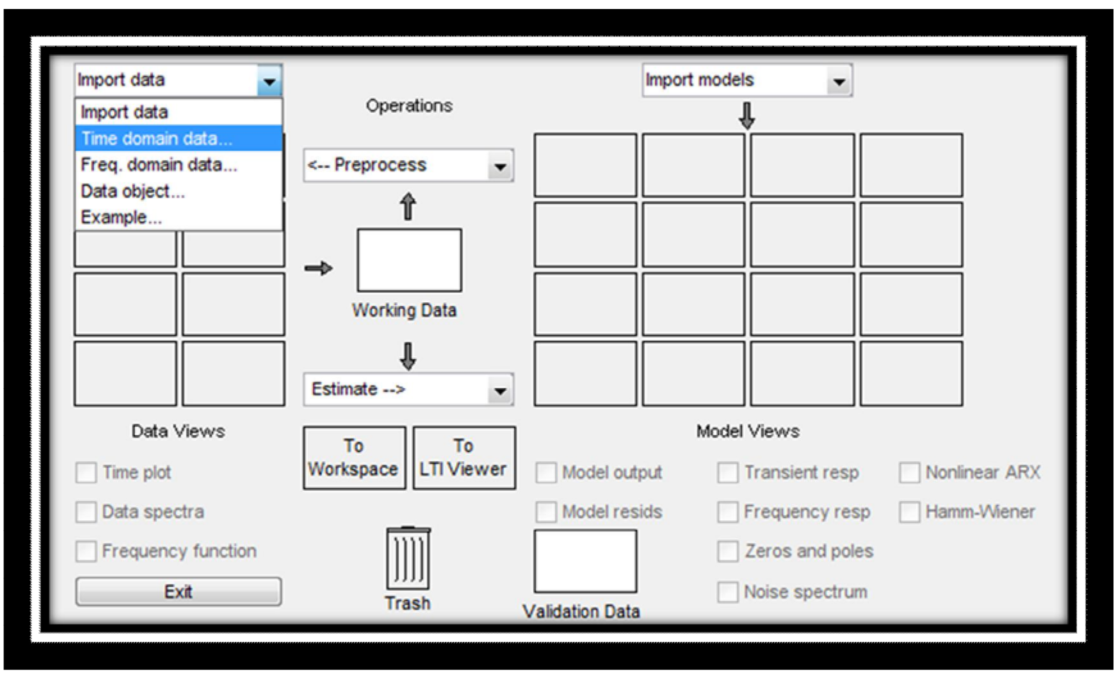


5- The model ARMAX appears on the right hand side of the GUI window, Drag the model to work space:

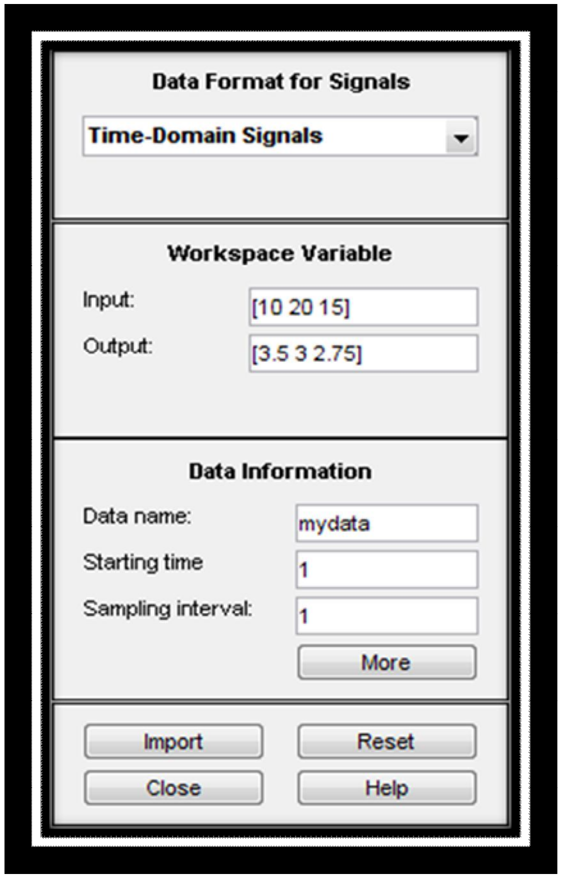


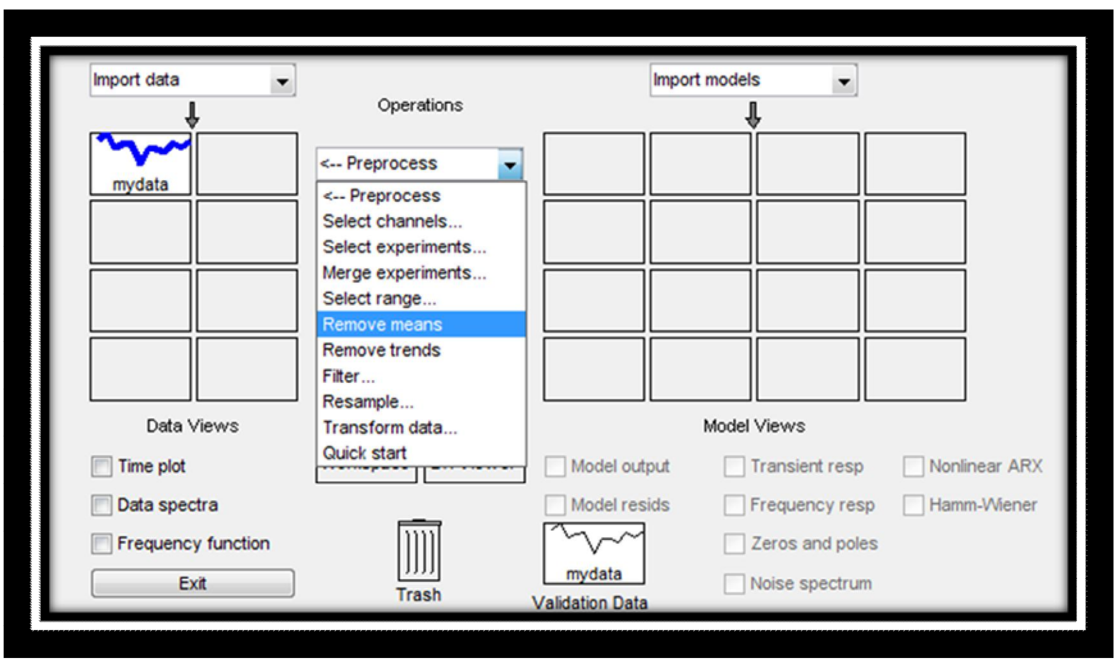
9- Write the model name in command window.

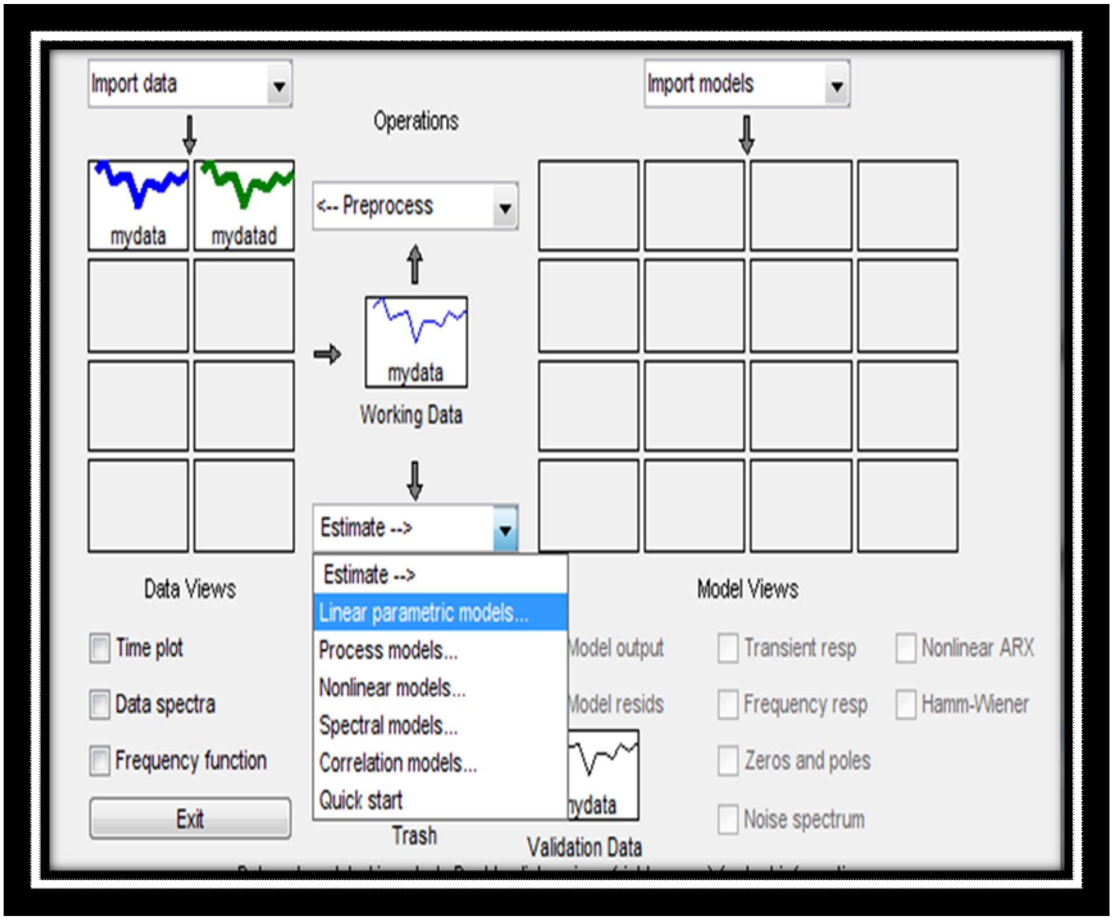
10- Find the transfer function for the model

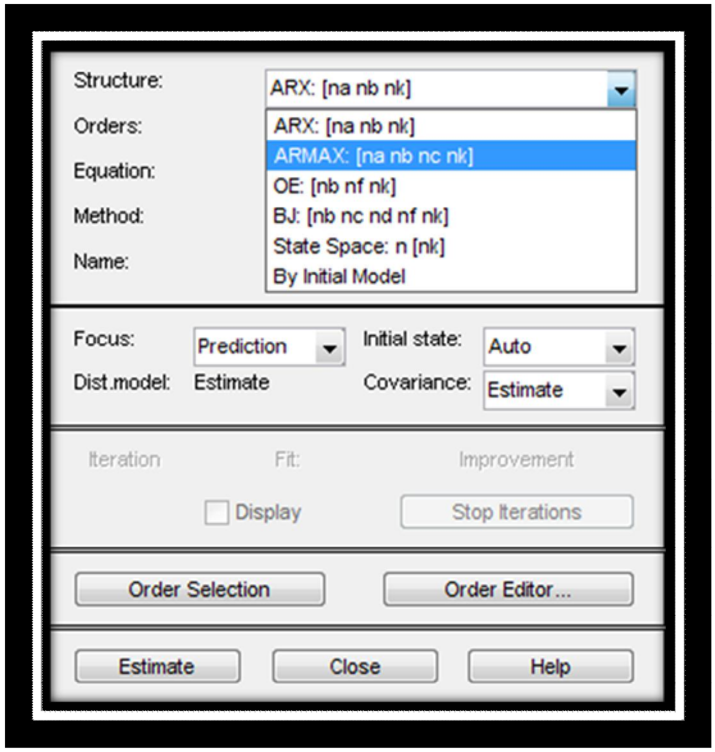


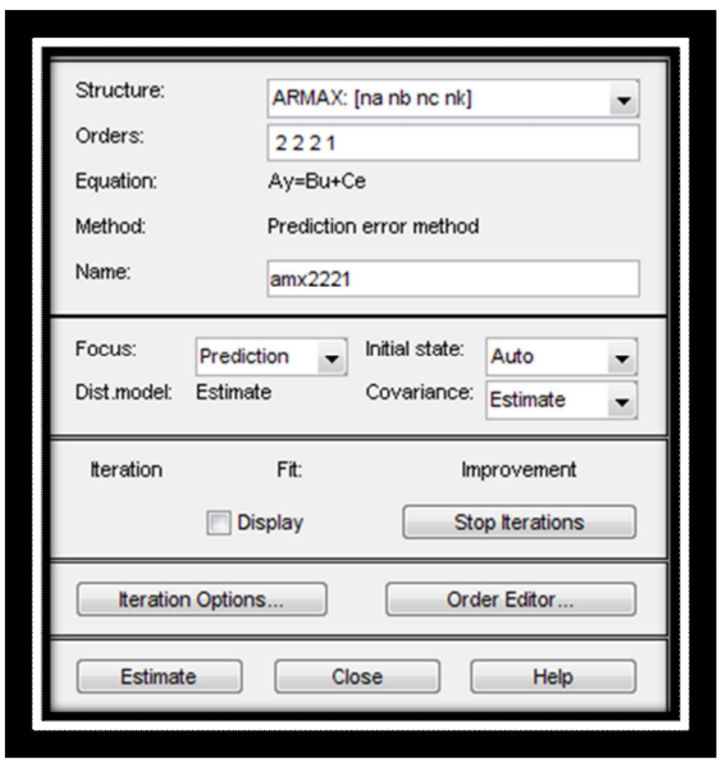

Data Format for Signals
Time-Domain Signals
Workspace Variable
Input:
Output:
Data Information
Data name:
mydata
Starting time
1
Sampling interval:
1
More
Import
Reset
Close
Help

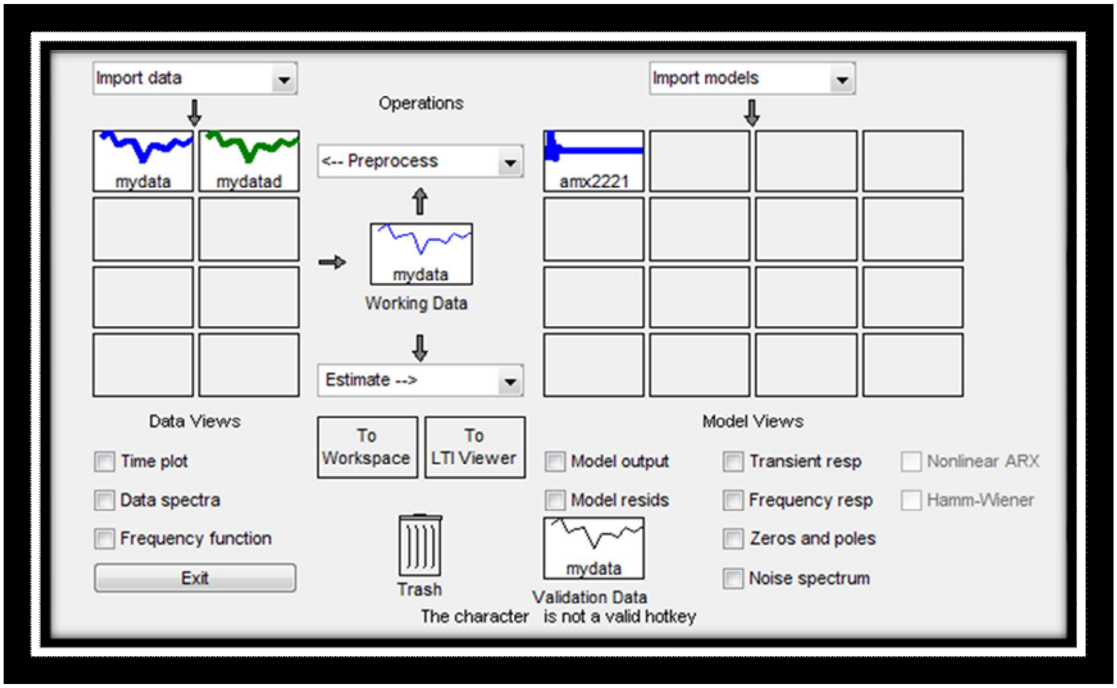

Data Format for Signals
Time-Domain Signals
Workspace Variable
Input:
[10 20 15]
Output:
[3.5 3 2.75]
Data Information
Data name:
mydata
Starting time
1
Sampling interval:
1
More
Import
Reset
Close
Help


Import data
Operations
Import models
mydata
<-- Preprocess
<-- Preprocess
Select channels...
Select experiments...
Merge experiments...
Select range...
Remove means
Remove trends
Filter...
Resample...
Transform data...
Quick start
Data Views
Model Views
Time plot
Data spectra
Frequency function
Exit
Trash
mydata
Validation Data
Model output
Model resids
Transient resp
Frequency resp
Zeros and poles
Noise spectrum
Nonlinear ARX
Hamm-Wiener


Import data
Operations
Import models
mydata
mydatad
<-- Preprocess
mydata
Working Data
Estimate -->
Estimate -->
Linear parametric models...
Process models...
Nonlinear models...
Spectral models...
Correlation models...
Quick start
Data Views
Model Views
Time plot
Data spectra
Frequency function
Exit
Trash
Validation Data
Transient resp
Frequency resp
Zeros and poles
Noise spectrum
Nonlinear ARX
Hamm-Wiener


Structure:
ARX: [na nb nk]
Orders:
Equation:
Method:
Name:
ARX: [na nb nk]
ARMAX: [na nb nc nk]
OE: [nb nf nk]
BJ: [nb nc nd nf nk]
State Space: n [nk]
By Initial Model
Focus:
Prediction
Initial state:
Auto
Dist.model:
Estimate
Covariance:
Estimate
Iteration
Fit:
Improvement
Display
Stop Iterations
Order Selection
Order Editor...
Estimate
Close
Help


Structure:
ARMAX: [na nb nc nk]
Orders:
2 2 2 1
Equation:
Ay=Bu+Ce
Method:
Prediction error method
Name:
amx2221
Focus:
Prediction
Initial state:
Auto
Dist.model:
Estimate
Covariance:
Estimate
Iteration
Fit:
Improvement
Display
Stop Iterations
Iteration Options...
Order Editor...
Estimate
Close
Help


Import data
Operations
Import models
mydata
mydatad
<-- Preprocess
amx2221
mydata
Working Data
Estimate -->
Data Views
Model Views
To Workspace
To LTI Viewer
Time plot
Data spectra
Frequency function
Exit
Model output
Model resids
Transient resp
Frequency resp
Zeros and poles
Noise spectrum
Nonlinear ARX
Hamm-Wiener
Trash
mydata
Validation Data
The character is not a valid hotkey