DESIGNING OF MODEL REFERENCE ADAPTIVE CONTROL SYSTEM FOR CONTINUOUS STIRRED TANKER REACTOR

MSC THESIS

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DESIGNING OF MODEL REFERENCE ADAPTIVE CONTROL SYSTEM FOR CONTINUOUS STIRRED TANKER REACTOR

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We, the undersigned, members of the Board of Examiners of the final open defense by **Mr. Berihu Berhe** Id. No: **PGCon/012/09**, have read and evaluated his/her thesis entitled **"Designing of Model Reference Adaptive Control System for Continuous Stirred Tanker Reactor"**, and examined the candidate. This is, therefore, to certify that the thesis has been accepted in partial fulfillment of the requirements for the degree.

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Declaration

I hereby declare that this MSc thesis entitled **“Designing of Model Reference Adaptive Control System for Continuous Stirred Tanker Reactor”** is my original work and has not been presented for a degree in any other university, and will not be presented by me to any other university for similar or any other degree award, and all sources of material used for this thesis have been duly acknowledged.

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ABSTRACT

Continuous Stirred Tank Reactor (CSTR) is the significant process which plays an extensive role in chemical processing and chemical industries. In this process, like temperature of reaction and concentration of reaction are to be controlled. Since most of chemical process are multi input multi output (MIMO) system, it is important to design decoupling system and to find relative gain array (RGA) for the purpose of avoiding the interaction and to select best pairing respectively among the input output relationship. Because control of MIMO systems are usually much more difficult as compared to the SISO case especially for adaptive control system. This thesis deals with the operation, mathematical modeling and controller design for the jacketed continuous stirred tank reactor. PID and model reference adaptive control (MRAC) Controller are designed to control the reactant mixture temperature and concentration of continuous stirred tank reactor. The transient and steady state performances of the MRAC are compared with PID controller. The simulation has shown that the control performance of model reference adaptive control (MRAC) is better and effective than PID controller.

Keywords: Multivariable, continuous stirred tank reactor, decoupler, PID controller, Model Reference Adaptive Control (MRAC)

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ABBREVIATION

CSTR ……………………………..Continuous Stirred Tank Reactor

MIMO……………………………..Multi input multi output

PID………………………………...proportional integral derivation

STR………………………………..Self-tuning regulators

MRAC …………………………....Model reference adaptive controller

TITO……………………………....two input-two output

RGA ………………………………Relative Gain Array

M_p…………………………………maximum overshoot

T_s……………………………….....settling time

A…………………………………..Area of heat exchange [m2]

C_a..………………………………...Concentration of reactor [mol/m3]

C_{af}………………………………...Concentration of feed stream [mol/m3]

C_p………………………………… Heat capacity [J/kg.K]

F…………………………………..Volumetric flow rate [L/Hr]

T…………………………………...Reactor temperature [‘c]

T_f…………………………………feed temperature [‘c]

U…………………………………..overall heat transfer coefficient[kcal/m2.’c.hr]

V…………………………………..reactor volume [L]

ΔE………………………………....Activation energy [kcal/kmol]

(-ΔH)………………………………heat of reaction [kcal/kmol]

ρ…………………………………..Density [kg/m3]

K_o…………………………………exponential factor

R…………………………………..ideal gas constant

$\boldsymbol{r_A}$…………………………………Rate of reaction [mol/m3.sec]

CHAPTER ONE

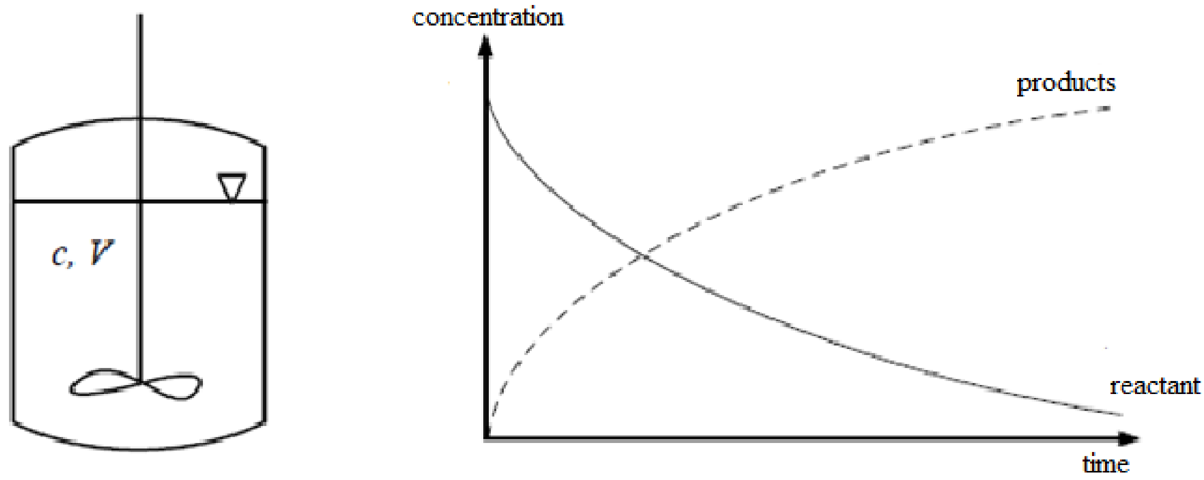
1. INTRODUCTION

1.1. Background

In any manufacturing process, where a chemical change is taking place, a chemical reactor is the heart of the plant. Depending on the mode of operation, reactors are classified as batch-wise or continuous. In batch processing, the reactants of the planet are charged at the beginning of the reaction and the products of the plant are removed at the end of each reaction. In continuous stirred tank reactor (CSTR), reactants of the planet are continuously charged and products are continuously removed [1].

Batch

A batch chemical reactor are a discontinuous reactor. It is essentially a stirred tank that is filled with the reactants before the reaction starts and emptied after it has run to completion (or to the extent that is needed). One of the disadvantages of this type of reactor is that for large production quantities the reaction has to be done multiple times in series. This requires the emptying and refilling of the reactor, often accompanied by cooling it off first and heating it up with the new batch. This large number of steps takes time and attention, and thereby reduces the productivity of the reactor. On the other hand, these reactors have the advantage that if multiple similar but different reactions are needed, often the same equipment can be used, and the additional effort is comparatively small [1].



a) Schematic diagram b) concentration of reactants

Figure1.1 a and b are represented Schematic diagram and the concentration of reactants and products in a batch reactor respectively [1].

Plug flow reactor (PFR)

Another type of continuous chemical reactors is the plug flow reactor (PFR). It is a tubular reactor, meaning that it consists of a long cylindrical pipe through which the reaction mixture is flowing steadily. Typically the assumption is made that the temperature, pressure, and composition do not vary radially with in the pipe, creating a plug that flows through the reactor. As the reactants flow through the PFR, they are consumed, creating a concentration profile along the length of the pipe. While these reactors can have a heating or cooling duty requirement that varies along the reactor, the reactor volume necessary to reach a particular conversion is lower than for a CSTR, while keeping the advantages of a continuous process [2].

Continuous stirred tank reactor (CSTR)

A continuous stirred tank reactor is like a batch reactor in that it consists of a tank and a stirrer, however with the addition of an inlet and an outlet that allow for a constant flow into and out of the reactor. Once the reactor is started up and reaches steady-state, it is usually assumed to have a constant volume as well as constant and homogeneous temperature, pressure, and composition. While continuous processes don't have the variability of batch processes, and during start-up will produce product that does not meet specifications, they have a number of advantages that make them attractive to use. For one, continuous reactors don't have to be cooled off, emptied, cleaned, refilled, and then heated to operating temperature. For another, if a reaction produces heat and the reactor needs to be cooled, the cooling duty for a CSTR is constant, and can be tuned as needed. For a batch reactor the cooling duty needed would vary with the reaction rate, and insufficient cooling can lead to a runaway reaction. Additionally, the product from one reactor is often used in subsequent steps for other reactions. If multiple steps are done in series in batch reactors, and each step takes a different amount of time, the intermediate products need to be stored in buffer tanks. These tanks can be eliminated or greatly reduced in size if each reactor produces a steady stream that can be fed to the next reactor. If a process has to be done in batches, several reactors are often used in parallel, shifted in time to give a continuous stream from the group of reactors [1, 4].

In general, Chemical reaction systems are usually nonlinear dynamical systems. A Continuous Stirred Tank Reactor (CSTR) is one of the most important unit operations in chemical industries which exhibits highly nonlinear behavior and usually has wide operating ranges. Chemical reactions in a reactor are either exothermic or endothermic and require that energy can either be removed or added to the reactor to maintain a constant temperature .The CSTR is normally run at steady state and is usually operated so as to be quite well mixed. As a result of this quality, the CSTR is generally modeled as having no special variations in concentration, temperature or reaction rate throughout the vessel, they are the same at the exit point as they are elsewhere in the tank. Thus the temperature and concentration at the exit are modeled as being the same as those inside the reactor [2, 3].

In common sense, ‘to adapt’ denotes the change in behavior to confirm new circumstances. An adaptive controller is a controller that can modify its behavior in response to the changing dynamics of the process and the character of the disturbances. The core element of all the approaches is that they have the ability to adapt the controller to accommodate changes in the process. This permits the controller to maintain a required level of performance in spite of any noise or fluctuation in the process. An adaptive system has wide application when the plant undergoes transitions or exhibits non-linear behavior and when the structure of the plant is not known. Gain scheduling is one form of adaptive control but it requires knowledge about all the process to be effective. Another alternative is to adapt the controller’s parameters or when a model is available to use the model identification error to tune the controller’s parameters [6].

There are two main reasons why adaptive controller is needed in chemical processes. First, most chemical processes are nonlinear. Therefore, the linearized models that are used to design linear controllers depend on the particular steady state (around which the process is linearized). It is clear that as the desired steady state operation of a process changes, the 'best' values of the controller's Parameters change. This implies the need for controller adaptation. Second, most of the chemical processes are nonstationary.

1.2. Problem statement

Since our environmental temperature varying from time to time due to so many factors, the variation of temperature cause the variation of chemical and physical property in the reaction in terms of concentration, time rate of reaction, temperature of product and etc even through, the environmental temperature is kept constant. The chemical reaction itself is either exothermic or endothermic and require that energy can either be removed or added to the reactor. Generally there is temperature variation either due to external factor due to environmental temperature or internal factor the chemical reaction itself.

In addition, the most of the chemical processes are nonlinear. Then the linearized models are needs to design linear controllers depend on the particular steady state (around which the process is linearized). It is clear that, as the desired steady-state operation of a process changes; the best values of the controller parameter change. This implies the need for adaptive controller and most of the chemical processes are dynamic. This changes lead again to deterioration in the performance of a controller, which was designed using some nominal values for the process parameters. Thus, it requires adaptation of the controller parameters.

Depending from the above problem, this thesis try to solve the problems of nonlinearity and time variant system of the chemical processes for the purpose of maintaining the required level of temperature and concentration at the output of the system using adaptive control system.

1.3. Objectives

1.3.1. General objective

The main objective of this thesis is to design the model reference adaptive control for the system using MIT rule for temperature and concentration of continuous stirred tanker reactor.

1.3.2. Specific objective

- To develop modified mathematical model of CSTRs.
- To optimize objective function (related to profit) using as manipulative variables.
- To design the PID controller for CSTR.
- To simulate both controllers and select the effective one in terms of performance characteristics.
- To design the Decoupler which reduces the interaction between input and output variables.

1.4. Scope

Model reference adaptive control for continuous stirred tanker reactor as multivariable control and nonlinear system will be done in this thesis, the PID is tuned using auto tuning mechanism and the MRAC adaptation mechanism is analyzed by MIT rule method. The designed system is finally simulated using MATLAB for reference inputs of the controlled variables.

1.5. Methodology

Methodology is the technique or method we apply to perform our specific objectives.

- Literature reviewing relevant theses which have inputs for the thesis.
- To develop the mathematical model for the continuous stirred tanker reactor (CSTR).
- Designing of relative gain array (RGA) for the continuous stirred tanker reactor (CSTR) to identify the best pair for input output relationship.
- Developing decoupling system for the continuous stirred tanker reactor (CSTR) in order to create SISO system.
- Designing of PID controller for the continuous stirred tanker reactor (CSTR) system using MATLAB auto tuning method.

- Designing model reference adaptive control system for continuous stirred tanker reactor (CSTR) using MIT rule in MATLAB/Simulink.
- Simulating the open-loop, PID and MRAS controlled system using MATLAB/Simulink.
- Testing, comparison of PID and MRAS using time domain performance analysis techniques.
- Conclusion for overall thesis work.

1.6. Thesis organization

This thesis consists of a total of five chapters and is organized as follows.

Chapter two presents the theoretical overview of continuous stirred tanker reactor (CSTR), and discusses the basic steps involved in developing both the steady state and dynamic state models of continuous stirred tanker reactor (CSTR) and its linearization technique.

Chapter 3 focuses on designing the decouplers and the controllers. Since the model developed in chapter two is strong coupling between inputs -output variables, it is required to design a decoupler and controller and this is done in detail in chapter three.

Chapter 4 discusses the results and simulations for the proposed controller.

Chapter 5 summarizes the major issues discussed throughout this thesis and consequently draws the general conclusions.

CHAPTER TWO

LITERATURE REVIEW

2.1. INTRODUCTION

A literature study was performed to obtain an overview of what research have already been reported on the chemical reactors performance, starting with what are the important parameters necessary to perform the analytical and simulation models.

Nonlinear behavior of the chemical process is not common characteristic. This feature not only heightens the control problem but also necessitates a nonlinear dynamic model of the process for control studies. PID controllers are proved to be perfect controllers for simple and linear processes. These controllers have been installed at most process plants, since they are simple, robust and familiar to the field operator.

Wei-Der Chang [14] has applied the ABC algorithm for the design of nonlinear PID controller for CSTR plant. Simulations results conclude that the proposed ABC based PID controller design outperform the real coded GA due to the robustness. Also it is observed that the results derived using ABC was consistent for all the trial runs.

Prakash et al [15] proposed CSTR nonlinear model through Fuzzy Takagi-Sugeno (T-S) fusion of linear models and also reported that the linear models are sufficient to handle the nonlinear process variations. The main disadvantage of this method is the need of local linear models identification.

Dougherty Danielle & Cooper Doug [7] has discussed multiple model adaptive strategy for multivariable model predictive control. This approach uses linear models based multi model network to predict the future performance of the system. The main drawback of the system is more tuning of prediction is required when the operating region varies.

Nahas *et al* [3] have proposed the use of internal model control for the control of chemical process. Since, it is difficult to describe the complete system behaviour using a single linear model; the performance of the control schemes based on single linear model being used to control and based on accurate model of the process, if process changes, need to change the model then the nonlinear process cannot yield satisfactory performance.

In Banerjee *et al* [13] the multiple local models are combined into a single parameter varying global model. The problem here is he assumed that each model has a controller that has been designed to perform satisfactorily within its operation region but not necessarily during a transition.

Nikolaou *et al* [9] presented the developments and further directions on linear control of non-linear processes and summarized that virtually all chemical processes are nonlinear. The problem here is he decide linear controllers adequate for those chemical process.

Wen Tan et al [18] has presented conventional PID tuning techniques viz., Z-N method, IMC method, etc., for the first order and second order process. The drawback involved in Z-N technique is PID parameters are derived from the damped oscillations gain and frequency, which is not generally preferred because the closed loop behaviour tends to be oscillatory and sensitive to uncertainty.

Among the various reactors, continuous flow stirred tank reactors have been used extensively to study the dynamic behavior of nonlinear continuous stirred tanker reactor systems. However, these systems have been based on the conventional control like PI and PID, Model Predictive Control and fuzzy logic control had used for mixing in the reactor. So that, they are not sufficient and easily condition for handling (control) for nonlinear and time variant (dynamic) MIMO continuous stirred tanker reactor

2.2. Adaptive control

It is well known that conventional control theories are widely suited for applications where the processes can be reasonably described in advance [11]. However, when the plant's dynamics are hard to characterize precisely or are subject to environmental uncertainties, one may encounter difficulties in applying the conventional design methodologies. Such as despite the difficulty in achieving high control performance, the fine-tuning of controller parameters is tedious task that always requires experts in both control theory and process information. Therefore, the control of systems with poorly known, nonlinear and uncertain dynamics has become a topics of considerable importance in the literature and presents great challenges for control engineers [6].

In the past few decades, there has been considerable interest in the development of adaptive control systems that automatically adjust controller parameters to compensate for un anticipated changes in the process dynamics. The ability of dealing with time-varying characteristics, non-linearity and uncertainties enables adaptive control algorithms to have significant potential for the operation of chemical process whose dynamics are poorly known or subject to changes in unpredictable way.

Adaptive control systems have been in existence for a few years, and a wide range of approaches have been developed. The core element of all the approaches is that they have the ability to adapt the controller to accommodate changes in the process [6, 11].

This permits the controller to maintain a required level of performance in spite of any noise or fluctuation in the process and the adaptive system has maximum application when the plant undergoes transitions or exhibits nonlinear behavior and when the structure of the plant is not known [6].

Gain scheduling is one form of adaptive control but it requires knowledge about all the process to be effective. Another alternative is to adapt the controller's parameters or when a model is available to use the system identification error to tune the controller's parameters. Consequently, tuning of the controller is indirect and necessarily requires an accurate model of the process for satisfactory performance [10].

In general, the adaptive controllers can be divided into two algorithms; direct and indirect. In direct algorithms, the parameters are updated directly. If the controller parameters are obtained indirectly via a design procedure, we use the term indirect algorithms. Sometimes, it is possible to re-parameterize the process model such that it is possible to use either a direct or indirect controller. However, the indirect methods have sometimes been called explicit self-tuning control, since the process parameters have been estimated. Direct updating of the regulator parameters has been called implicit self -tuning control. However, the most common regulators is a feedback controller with fixed parameters. Through feedback it is possible to decrease the sensitivity to parameter variations by increasing the loop gain of the system. The main drawbacks of high-gain controllers are the magnitude of the control signal and the problem of stability of the closed loop system. If there are bounds on the uncertainties of the process parameters, it is possible to design robust controllers by increasing the complexity of the controller. To use this approach, it is necessary to know the structure of the process fairly accurately and to have bounds on the variations of the parameters.

2.2.1. Gain Scheduling

In many situations it is known how the dynamics of a process change with the operating conditions of the process. One source for the change in dynamics may be nonlinearities that are known. It is then possible to change the parameters of the controller by monitoring the operating conditions of the process, this idea is called gain scheduling.

Its principle is to reduce the effects of parameter variations by changing the parameters of the regulator as function of auxiliary variables that correlate well with these changes in process dynamics. It is a nonlinear feedback controller of a special type. It has a linear regulator whose parameters are changed as a function of operating conditions in a programmed way [14].

Gain scheduling was used in special cases: such as autopilots for high-performance air-craft. The principle: It is sometimes possible to find auxiliary variable that correlate well with the changes in process dynamics. It is then to reduce the effects of parameter variations simply by changing the parameters of the regulator as functions of auxiliary variables as shown in figure below.

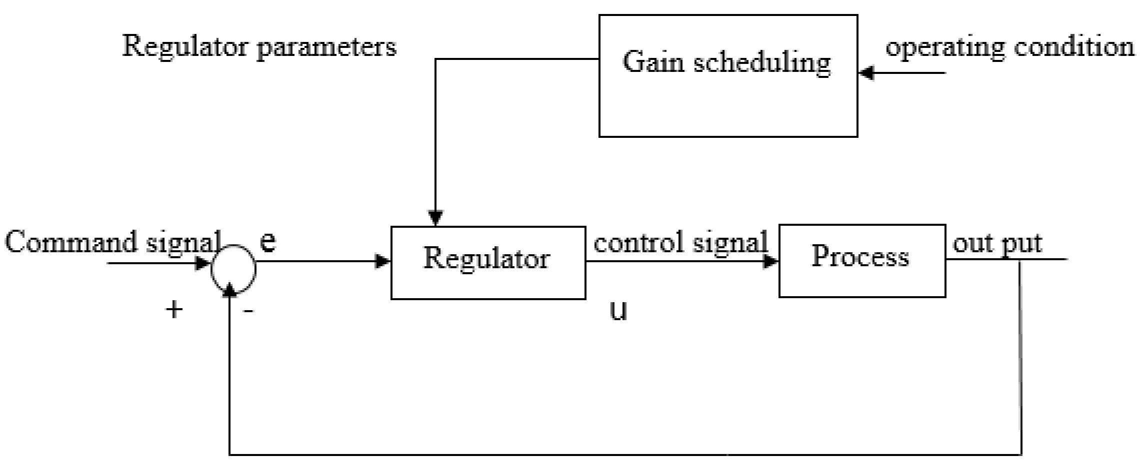


Figure 2.1 Block diagram of gain scheduling [14]

It is difficult to give general rules for designing gain scheduling regulators. The key question is to determine the variables that can be used as scheduling variables. It is clear that these auxiliary signals must reflect the operating conditions of the plant.

2.2.2. Self-tuning regulators

In an adaptive system, it is assumed that the regulator parameters are adjusted all the time. This implies that the regulator parameters follow changes in process. It is difficult to analyze the convergence and stability properties of such systems. To simplify the problem it can assume that the process has constant but unknown parameters. When the process is known, the design procedure specifies a set of desired controller parameters. The adaptive controller should converge to these parameter values even when the process is known. A regulator with this property is called Self-Tuning, since it automatically tunes the controller to the desired performance. The Self-Tuning Regulator (STR) is based on the idea of separating the estimation of unknown parameters from the design of the controller [6]. The basic idea is illustrated in figure bellow.

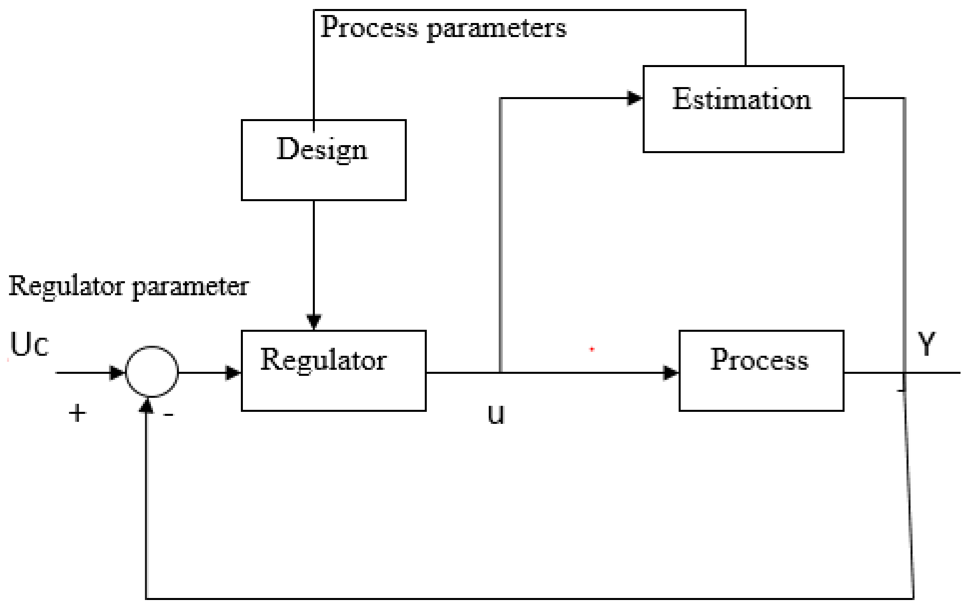


Figure 2.2 Block diagram of self-tuning regulator [6]

In the block diagram, the design block represents an on-line solution to the design problem for a system with unknown parameters. This is called the underlying design problem. The design method is chosen depending on the specifications of the closed loop systems. Different combinations of estimation methods lead to regulators with different properties.

2.2.3. Stochastic adaptive control.

Stochastic adaptive control theory is concerned with recursive estimation of unknown parameters and control for systems with uncertainties modeled as random variables or random processes and to formulate the stochastic adaptive control problem we must specify the model for the process, the admissible control signals, and the specifications (loss function) for the closed loop system. The theory is motivated by applications in such diverse areas as aerospace guidance and control, signal processing and communications and manufacturing processes. Mathematical theory of identification, filtering, control and stochastic adaptive control for models based on stochastic difference equations such as autoregressive processes and stochastic differential equations [11].

2.2.4. Model reference adaptive control system (MRAS)

Model reference adaptive controller (MRAC) is a controller used to force the actual process to behave like idealized model process. MRAC systems adapt the parameters of a normal control system to achieve this match between model and process [6].

The standard implementation of MRAC based systems contains the four key blocks shown blew. The reference model defines the desired performance characteristics of the process being controlled. The adaptation law uses the error between the process and the model output, the process output and input signal to vary the parameters of the control system. These parameters are varied so as to minimize the error between the process and the reference model.

The control system can be anything from a simple gain based controller to a more complicated parameter based transfer function or plant matrix. Whatever type of control system is used the parameters of the controller must be being varied by the adaptation law. The final element of the MRAC system is the process that is being controlled [18].

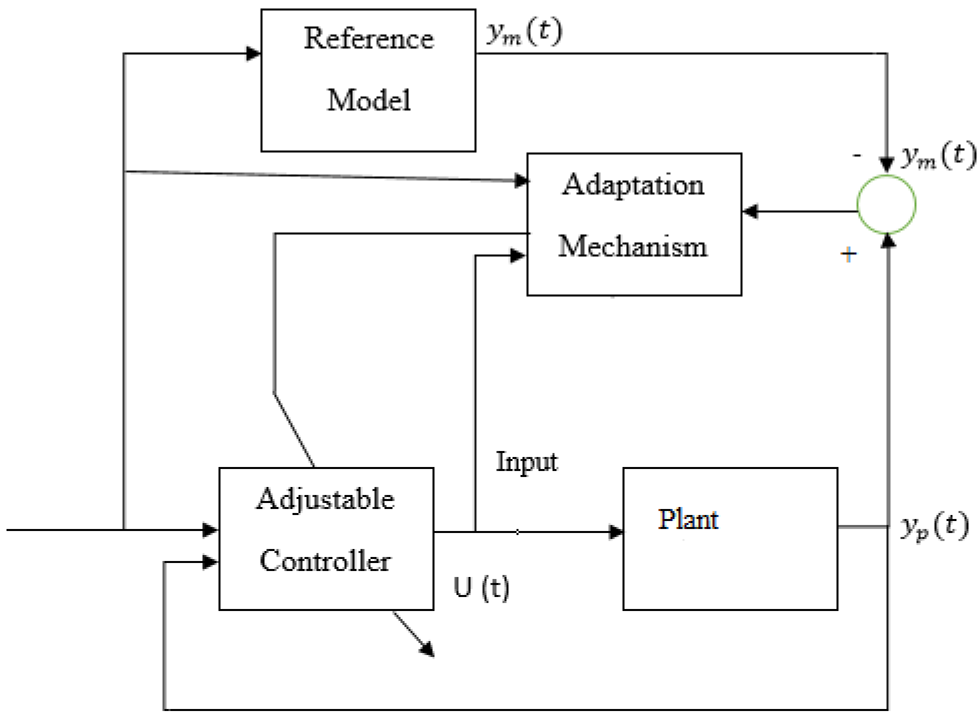


Figure 2.3 Model reference adaptive controller [18]

CHAPTER THREE

THEORETICAL BACKGROUND AND MATHEMATICAL MODELING

3.1. Introduction

A Continuous Stirred Tank Reactor (CSTR) is one of the most important unit operations in chemical industries which exhibits highly nonlinear behavior and usually has wide operating ranges. Chemical reactions in a reactor are either exothermic or endothermic and require that energy can either be removed or added to the reactor to maintain a constant temperature. The CSTR is normally run at steady state and is usually operated so as to be quite well mixed. As a result of this quality, the CSTR is generally modeled as having no special variations in concentration, temperature or reaction rate throughout the vessel. Since the temperature and concentration are identical everywhere within the reaction vessel, they are the same at the exit point as they are elsewhere in the tank. Thus the temperature and concentration at the exit are modeled as being the same as those inside the reactor [1, 2].

3.2. Component of the System

This system contains such as:

- ➢ The CSTR has two input signals:
  - ✓ Jacket coolant temperature. = T_{fc}
  - ✓ Input concentration flow rate. = q
- ➢ And two output signals:
  - ✓ Concentration in reactor tank. = C_a
  - ✓ Reactor temperature. = T

3.3. System Model (block diagram)

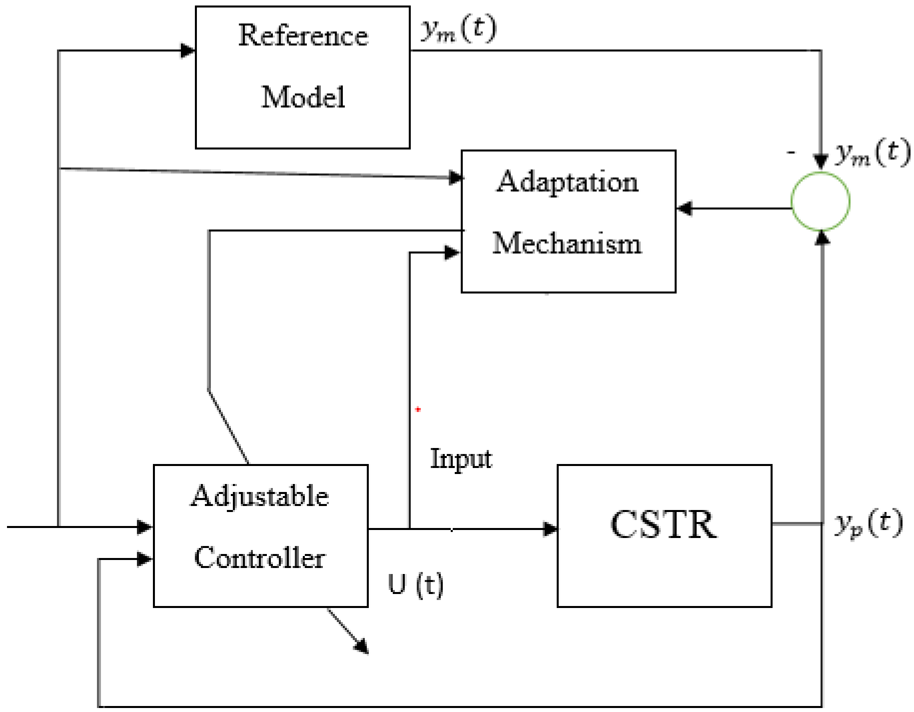


Figure 3.1 Block diagram of the system

3.4. Mathematical Modeling

To analize the behavior of a process, a mathematical representation of the physical and chemical phenomenon taking place in it is essential and this representation constitutes the mathematical model. The activities leading to the construction of the model is called modeling. The main uses of mathematical modeling is represent the system in mathematical term that is simple to apply simulation software.

3.4.1 Approach

To investigate how the behavior of a chemical process change with time under the influence of changes in the external disturbances and manipulated variables and consequently design an appropriate controller, we can use two different approaches.

3.4.1.1 Experimental approach

In this case the physical equipment of the chemical process is available to us. Consequently, we change deliberately the values of various inputs (disturbances, manipulated variables) and through appropriate measuring devices we observe. How the outputs (temperature, concentration) of the chemical processes changes with time. Such a procedure is time and effort consuming and it's usually quite costly since a large number of such experiments must be performed.

3.4.1.2 Theoretical approach

It is quite often the case that we have to design the control system for a chemical process before the process has been constructed. In such a case, we cannot rely on the experimental procedure, and we need a different representation of the chemical process in order to study its dynamic behavior. This representation is usually given in terms of a set of mathematical equations (differential, algebraic) whose solution yields the dynamic or static behavior of the chemical process we examine.

3.5. Process description (ideal CSTR)

Highly nonlinear CSTR is common in chemical and petrochemical plants. [5] In the process considered for the simulation study as shown in Figure 3.2, an irreversible, exothermic chemical reaction **A** → B occurs in constant volume reactor that is cooled by a single coolant stream. A feed material of composition c_{af} enters the reactor at temperature T_f, at a volumetric flow rate **q**. Product is withdrawn from the reactor at the same volumetric flow rate q. The mixing is assumed to be efficient enough to guarantee homogeneity of the liquid content within the reactor.

In a jacketed CSTR the heat is added or removed by virtue of the difference between the jacket fluid and the reactor fluid. Often, the heat transfer fluid is pumped through the agitation nozzles that circulate the fluid through the jacket at a high velocity. The coolant flows at a flow rate of q_c and at a feed temperature T_{fc}. The exit temperature of the coolant fluid is T_c.

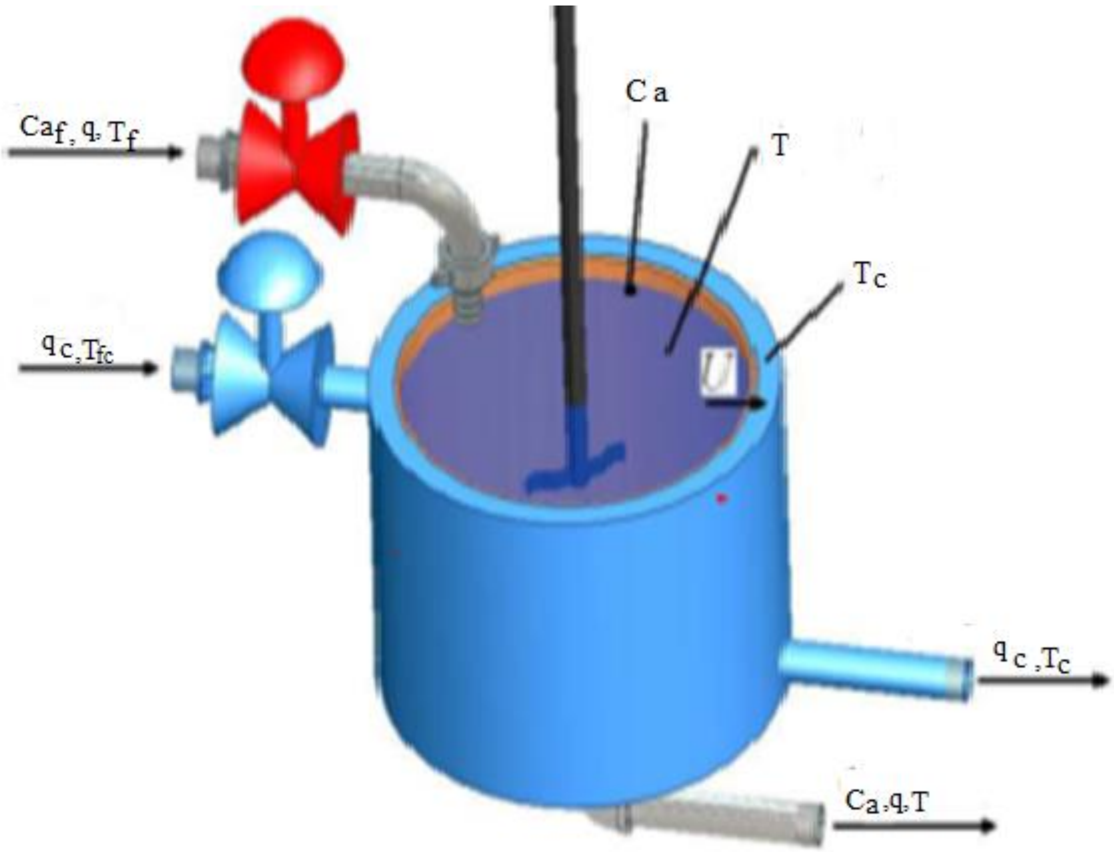


Figure 3.2 Schematic diagram of a continuous stirred tank reactor [1]

3.6. Mathematical Modeling of a Continuous Stirred Tank Reactor

Reactor Mass Balance

This is an important step because it helps to ensure that the resulting mathematical equation will have an understandable physical meaning. Just starting off by writing down equations is often liable to lead to fundamental errors, at least on the part of the beginning, all balance equations have a basic logic, as expressed by the generalised statement of the component balance given below, and it is very important that the model equations also retain this relationship. If the tank is well-mixed, the concentrations and density of the tank contents are uniform throughout. This means that the outlet stream properties are identical with the tank properties, in this case concentration c_{af} and density ρ .

The total mass in the system is given by the product of the volume of the tank contents V (m3) multiplied by the density ρ (kg/m3), thus V ρ (kg). The mass of any component A in the tank is given in terms of actual mass or number of moles by the product of volume V times the concentration of A, C_a (kg of A/m3 or kmol of A/m3), thus giving VC_a in kg or kmol.

$$\begin{pmatrix}\text{Rate of}\\ \text{accumulation}\\ \text{of mass of}\\ \text{component}\\ \text{in the system}\end{pmatrix} = \begin{pmatrix}\text{Mass flow}\\ \text{of the}\\ \text{component}\\ \text{in to}\\ \text{the system}\end{pmatrix} - \begin{pmatrix}\text{Mass flow}\\ \text{of the}\\ \text{component}\\ \text{out of}\\ \text{the system}\end{pmatrix} + \begin{pmatrix}\text{Rate of}\\ \text{production}\\ \text{of the}\\ \text{component}\\ \text{by the reaction}\end{pmatrix}$$

$$\frac{VdC_a}{dt} = \mathrm{q}\left(C_{af} - C_a\right) - Vr_A \qquad (3.1)$$

$$r_A = kC_a \qquad (3.2)$$

$$k = k_o e^{\left(-\frac{E}{RT}\right)} \qquad (3.3)$$

Where

C_a is the product (effluent) concentration of component A in the reactor and

r_A is the rate of reaction per unit volume.

The Arrhenius expression is normally used for the rate of reaction. A first order reaction results in the following expression [2].

$$r_A = k_o e^{(-\frac{E}{RT})} C_a \qquad (3.4)$$

Where,

k_o is the reaction rate constant,

E is the activation energy,

R is the ideal gas constant and

T is the reactor temperature on an absolute scale (R, Rankine or K, Kelvin).

Reactor Energy Balance

Energy balances are needed whenever temperature changes are important, as caused by reaction heating effects or by cooling and heating for temperature control. Such a balance is needed when the heat of reaction causes a change in reactor temperature. This change obviously influences the reaction rate and therefore the rate of heat evolution. Based on the law of conservation of energy, energy balances are a statement of the first law of thermodynamics. The internal energy depends not only on temperature, but also on the mass of the system and its composition. For that reason, material balances are almost always a necessary part of energy balancing. [2]

$$\begin{pmatrix} rate\ of \\ accumulation \\ of\ energy \end{pmatrix} = \begin{pmatrix} rate\ of \\ energy \\ in\ put\ due \\ to\ flow \end{pmatrix} - \begin{pmatrix} raet\ of \\ energy \\ out\ put\ due \\ to\ flow \end{pmatrix} + \begin{pmatrix} rate\ of\ energy \\ released\ due\ to \\ chemical\ reaction. \end{pmatrix} - \begin{pmatrix} rate\ of\ work \\ done\ by\ the \\ system\ on\ the \\ surrounding \end{pmatrix}$$

If specific heat capacities are assumed constant and for zero mechanical work done to the system, the energy balance equation simplifies to;

$$\frac{\mathrm{dT}}{\mathrm{dt}} = \frac{\mathrm{q}}{V}((T_f - \mathrm{T}) + \frac{-\Delta H}{C_p \rho} k_o e^{\left(-\frac{E}{RT}\right)} C_a + UA(T_{fc} - \mathrm{T}) \qquad (3.5)$$

Where

ΔH is the heat of reaction

U is the heat transfer coefficient

A is the heat transfer area

$\boldsymbol{T_f}$ is the reactant temperature

$\boldsymbol{T_{fc}}$ is the coolant temperature in the jacket.

From the equations (3.1) and (3.5), the mass balance and energy balance equations of the CSTR are obtained as follows:

$$\frac{dC_a}{dt} = \frac{\mathrm{q}}{V}\left(C_{af} - C_a\right) - k_o e^{\left(-\frac{E}{RT}\right)} C_a \qquad (3.6)$$

$$\frac{\mathrm{dT}}{\mathrm{dt}} = \frac{\mathrm{q}}{V}((T_f - \mathrm{T}) + \frac{-\Delta H}{C_p \rho} k_o e^{\left(-\frac{E}{RT}\right)} C_a + UA(T_{fc} - \mathrm{T}) \qquad (3.7)$$

The modeling equations of the CSTR such as equations (3.1) and (3.5) contain nonlinear functions of T and C_a. They are coupled and it is not possible to solve one equation independently of the other. For designing the controllers for such a nonlinear process, one of the approaches is to represent the nonlinear system as a family of local linear models. In the following section, state variable models pertaining to the nonlinear equations are derived.

3.6.1. State Variable Form of the Equations

When a system's equations of motion are rewritten as a system of first-order differential equations, each of these differential equations consists of the time derivative of the one of the state variables on the left-hand side and an algebraic function of the state variables as well as system outputs, on the right-hand side. These differential equations are called state-variable equations.

Equations (3.1) and (3.5) are represented in the standard state variable of continuous stirred tank reactor.

$$\frac{dC_a}{dt} = f_1(C_a, \mathrm{T}) = \frac{q}{V}\left(C_{af} - C_a\right) - k_o e^{\left(-\frac{E}{RT}\right)} C_a \qquad (3.8)$$

$$\frac{d\mathrm{T}}{dt} = f_2(C_a, \mathrm{T}) = \frac{q}{V}\left(\left(T_f - \mathrm{T}\right) + \frac{-\Delta H}{C_p \rho} k_o e^{\left(-\frac{E}{RT}\right)} C_a + \frac{UA}{V C_p \rho}\left(T_{fc} - \mathrm{T}\right) \qquad (3.9)$$

3.6.2. Equilibrium point

An equilibrium of dynamic system is a value of the state variable where the state variables do not change. Or an equilibrium point is the solution of the state variable that does not change with time. Using MATLAB fsolve function, the equilibrium value of continuous stirred tanker reactor of the concentration and temperature from eq. (3.8) and (3.9) at the constant parameter value which are given in the table 3.1 are 0.064 and 116.1391 respectively. The property of non-linear differential equation has multiple equilibrium point. But because of the degree of exponential function exponent is single, then the equilibrium point of concentration and temperature are also single value even if the differential equation of the system is non-linear function.

3.6.3. Linearization of Dynamic Equation

Linearization is the process of replacing the nonlinear system model by its equivalent linear counterpart in a small region about its equilibrium point. Dealing with nonlinear systems is difficult because known mathematical methods do not give us powerful enough means to analytically attack many problems which we encounter here. In order to simplify things and make them more manageable, linearization is quite often used. By linearizing nonlinear system about a single equilibrium state, the linear systems theory achievements can be explored. Method of linearization are: graphical, algebraic, analytical, harmonic, statistical and a combined one (harmonic plus statistical). For this linearization, analytical linearization is carried out by expanding the nonlinear function into a taylor series at the nominal operating point.
Let the state and input variables be defined in deviation variable form and the stability of the nonlinear equation can be determined by finding the following state space form:

$$\dot{x} = Ax + Bu \tag{3.10}$$

$$y = Cx + Du \tag{3.11}$$

We can determine the Eigen values of the A (state space) matrix and the state and input variables be defined in Deviation variable form. The matrices A and B represent the Jacobian matrices corresponding to the nominal values of the state variables and input variables and x, u and y represent the deviation variables.

In order to solve these two equations, all parameters and variables except for two ($C_a$and T) must be specified. When choosing initial guesses for numerical algorithm, it is important to use physical insight about the possible range of solutions.

For example, the feed concentration of A is 10 kgmol/ $m^3$and the only reaction consumes A; the possible range for the concentration of A is 0<C_a<10. Also it is to show that a lower bound for temperature is 298 K, which would occur if there was no reaction at all.

Table 3.1 Reactor Parameter's value [12]

No	Parameter	Unit	Value
1	q/V	hr-1	1
2	k_o,	hr-1	$10e^{15}$
3	(-ΔH),	kcal/kmol	6000
4	E,	kcal/kmol	12189
5	$C_p\rho$,	kcal/m^3'c	500
6	T_c ,	K	300
7	c_f ,	kmol/m^3	10
8	UA/V ,	kcal/m^3'c.hr	145
9	T_f ,	K	312

According to the above equilibrium point and the value of constant parameter, the state matrix A, input matrix B, output matrix C and direct transmission matrix D are given below.

The Jacobian matrix A is given as,

Let $x_1 = C_a$ and $x_2 =$ T

$$\frac{dx_1}{dt} = f_1(x_1, x_2) = \frac{q}{V}\left(C_{af} - x_1\right) - k_o e^{\left(-\frac{E}{Rx_2}\right)} x_1 \qquad (12)$$

$$\frac{dx_2}{dt} = f_2(x_1, x_2) = \frac{q}{V}\left(\left(T_f - x_2\right) + \frac{-\Delta H}{C_p\rho} k_o e^{\left(-\frac{E}{Rx_2}\right)} x_1 + \frac{UA}{VC_p\rho}\left(T_{fc} - x_2\right) \qquad (13)$$

$$A = \begin{bmatrix} A_{11} & A_{12} \\ A_{21} & A_{22} \end{bmatrix} = \begin{bmatrix} \frac{\partial f_1}{\partial x_1} & \frac{\partial f_1}{\partial x_2} \\ \frac{\partial f_2}{\partial x_1} & \frac{\partial f_2}{\partial x_2} \end{bmatrix} \quad \text{at } x_1/x_2 = x_{1e}/x_{2e} \qquad (3.14)$$

Where

x_{1e} *and* x_{2e} are the equilibrium point of the state variable of x_1 *and* x_2 respectively.

From above eq (3.14) the value of individual of A matrix's are

$$A_{11} = \left(-\frac{q}{V} - k_s\right) \qquad A_{21} = -\left[\frac{\Delta H}{\rho C_p}\right] k_s$$

$$A_{12} = \left(-x_{1e}\dot{k_s}\ \right) \qquad A_{22} = \left(-\frac{q}{V}\right) - \frac{\Delta H}{\rho C_p}\dot{k_s} x_{1e} + \frac{UA}{\rho V C_p}$$

Where

$$k_s = k_o e^{\left(-\frac{E}{Rx_{2e}}\right)} \quad \text{and} \quad \dot{k_s} = k_o e^{\left(-\frac{E}{Rx_{2e}}\right)}\left(\frac{E}{R{x^2}_{2e}}\right)$$

Then, the numerical value of matrix A at the equilibrium point is

$$\text{A} = \begin{bmatrix} -1.567 & -0.0829 \\ 1.868 & -0.3115 \end{bmatrix}$$

The Jacobian matrix B is given by

$$B = \begin{bmatrix} B_{11} & B_{12} \\ B_{21} & B_{22} \end{bmatrix} = \begin{bmatrix} \frac{\partial f_1}{\partial u_1} & \frac{\partial f_1}{\partial u_2} \\ \frac{\partial f_2}{\partial u_1} & \frac{\partial f_2}{\partial u_2} \end{bmatrix} \text{ at } u_1/u_2 = u_{1e}/u_{2e} \tag{3.15}$$

Where

u_{1e} *and* u_{2e} are the equilibrium point of the input of u_1 *and* u_2 respectively and the individual value input matrix are;

$$B_{11} = \frac{1}{V}(c_{af} - x_{1e}) \qquad B_{12} = 0$$

$$B_{21} = \frac{1}{V}(T_f - x_{2e}) \qquad B_{22} = \frac{UA}{\rho V C_p}$$

Then, the numerical value of matrix B at the equilibrium point is

$$\text{B} = \begin{bmatrix} 1.4363 & 0 \\ -13.8 & 0.3 \end{bmatrix}$$

Similar fashion, the output matrix C and the direct transmission matrix are given by

$$C = \begin{bmatrix} C_{11} & C_{12} \\ C_{21} & C_{22} \end{bmatrix} = \begin{bmatrix} \frac{\partial y_1}{\partial x_1} & \frac{\partial y_1}{\partial x_2} \\ \frac{\partial y_2}{\partial x_1} & \frac{\partial y_2}{\partial x_2} \end{bmatrix} = \begin{bmatrix} 1 & 0 \\ 0 & 1 \end{bmatrix} \tag{16}$$

$$D = \begin{bmatrix} D_{11} & D_{12} \\ D_{21} & D_{22} \end{bmatrix} = \begin{bmatrix} \frac{\partial y_1}{\partial u_1} & \frac{\partial y_1}{\partial u_2} \\ \frac{\partial y_2}{\partial u_1} & \frac{\partial y_2}{\partial u_2} \end{bmatrix} = \begin{bmatrix} 0 & 0 \\ 0 & 0 \end{bmatrix} \tag{17}$$

3.6.4. Stability Analysis

Stability can be defined as the characteristic of a system that determines the value of natural response as time approaches infinity. The concept of stability of systems is of paramount importance and from an application point of view an unstable system is of no use. A dynamic system that returns to its equilibrium state following a slight perturbation is referred to as a stable system. If it moves away and does not return to its original equilibrium state, it will be referred to as an unstable system. And if it responds to the perturbation but does not move away from its initial equilibrium state, it will be referred to as a neutrally stable system.

The stability of particular operating point is determined by finding the A-matrix for that particular operating point and finding the Eigen values of the A-matrix.

$$A = \begin{bmatrix} -1.567 & -0.0829 \\ 1.868 & -0.3115 \end{bmatrix}, \qquad \mathrm{B} = \begin{bmatrix} 1.4363 & 0 \\ -13.8 & 0.3 \end{bmatrix}$$

$$C = \begin{bmatrix} 1 & 0 \\ 0 & 1 \end{bmatrix}, \qquad D = \begin{bmatrix} 0 & 0 \\ 0 & 0 \end{bmatrix}$$

The eigen value of A are -1.4283, -0.4502 this indicates both eigen value are negative then the system is stable node.

3.6.5. Derivation of transfer function

Transfer function can be defined as the function of complex variable that represent the ratio of the Laplace transform of the output variable to the Laplace transform of the input variable when the initial condition are assume to be zero. According this idea, the transfer functions for concentration of the reactant to flow rate and jacket temperature and the temperature of the reactant to flow rate and jacket temperature are determined using MATLAB and are given by:

Transfer function relating c_a to q is

$$H_{11} = \frac{1.4364s + 1.5393}{s^2 + 1.4682s + 0.5152}$$

Transfer function relating c_a to T_c is

$$H_{12} = -\frac{0.0249}{s^2 + 1.4682s + 0.5152}$$

Transfer function relating T to q is

$$H_{21} = \frac{-13.171s - 12.5517}{s^2 + 1.4682s + 0.5152}$$

Transfer function relating T to T_c is

$$H_{22} = \frac{0.3s + 0.34}{s^2 + 1.4682s + 0.5152}$$

3.7. Relative Gain Array (RGA)

Most industrial chemical processes are multi-input multi-output (MIMO) in nature. The control of multi-input multi-output systems (MIMO) is more difficult than for single-input single output systems (SISO) due to the multitude of input-output couplings. Coupling, simply means that a change in any input leads to changes in many outputs. Nevertheless, in many cases, a simple decentralized controller is usually sufficient to achieve desired performance goals. However, there is a need for systematic techniques that can suggest the most promising configurations or pairings for the decentralized controller.

The Relative Gain Array (RGA) is the most widely used approach to decide the input-output pairs for the SISO controllers in MIMO systems. It basically measures the change in the gain between one input and one output when closing the loops between all other inputs and outputs by means of perfect control. If no large change in the gain occurs, the RGA recommends selecting those input-output as a pair for a SISO controller. This change is tested for all other inputs and outputs combinations in order to select the remaining pairs. Having selected all input-output pairs, the MIMO system can be controlled by multi-loop SISO (decentralized) controller. Therefore, decentralized and decoupled controllers, pairing is required. Relative Gain Array and condition numbers are useful techniques for pairing and its purpose is to evaluate the interaction between the loops.

RGA Analysis: 2 x 2 system

Steady-state process model for two input two output system is

$$Y_1 = H_{11} U_1 + H_{12} U_2 \qquad (3.18)$$

$$Y_2 = H_{21} U_1 + H_{22} U_2 \qquad (3.19)$$

The RGA , is defined as:

$$\Lambda = \begin{bmatrix} \lambda_{11} & \lambda_{12} \\ \lambda_{21} & \lambda_{22} \end{bmatrix}$$

Where the relative gain, λ_{ij} relates the i^{th}controlled variable and the j^{th} manipulated variable.

$$\lambda_{ij} = \frac{i^{th}\text{loop gain when allother loops are open}}{i^{th}\text{loop gain when allother loops are closed}} = \frac{\left(\frac{\partial y_i}{\partial u_j}\right)u}{\left(\frac{\partial y_i}{\partial u_j}\right)y} \qquad (3.20)$$

$\left(\frac{\partial y_i}{\partial u_j}\right) u$: Partial derivative evaluated with all of the manipulated variables except u_j, held constant (H_{ij}).

$\left(\frac{\partial y_i}{\partial u_j}\right) y$: Partial derivative evaluated with all of the controlled variables except y_i, held constant.

The properties of relative gain array are

i. λ_{ij} is dimensionless

ii. $\sum_i \lambda_{ij} = \sum_j \lambda_{ij}$=1

The Individual parameter of two input two output system can be calculated as

$$\lambda_{11} = \frac{1}{1 - \frac{H(0)_{12}\ H(0)_{21}}{H(0)_{11}\ H(0)_{22}}}, \qquad \lambda_{12} = \lambda_{21} = 1 - \lambda_{11} \qquad (3.21)$$

$$\Lambda = \begin{bmatrix} \lambda_{11} & 1 - \lambda_{11} \\ 1 - \lambda_{11} & \lambda_{11} \end{bmatrix} \qquad (3.22)$$

According to this relative gain array formula, the relative gain array of continuous stirred tank reactor is

$$\Lambda = \begin{bmatrix} 2.3 & -1.3 \\ -1.3 & 2.3 \end{bmatrix}$$

The diagonal elements are negative, this show that the system of continuous stirred tank reactor is structural unstable. In order to eliminate the interaction, A decouples D_1 (s) & D_2 (s) must be designed for those two systems for the purpose of cancellation the effect of jacket temperature on the outlet concentration (y_1) and the effect of input flow rate on the outlet temperature (y_2)

CHAPTER FOUR

Decoupler and Controller Design

4.1. Design of decoupler

One of the early approaches to multivariable control is decoupling. By adding additional controllers called de-couplers to a conventional multi-loop configuration, the design objective of reducing control loop interactions can be realized. Because, In the case of multi –loop control is not effective in reaching the desire specifications, possibility strategy for tackling the MIMO control could be to transform the transfer function matrix in to diagonal one. This strategy is called decoupling. If the result transfer function matrix were diagonal (approximately, diagonal determinant, then MIMO control would be equivalent to a set independent control loop. This help us the stability of closed loop control is determined solely by the stability characteristic of the individual feedback control loop and a set of point change for one controlled variable has no effect on the other controlled variable [21].

The condition of input–output decoupling of a system may be stated as follows. Consider the system shown in figure 4.1 and assume that there are the same number of inputs and outputs. Determine a pair of matrix D feed forward law such that every input of the closed loop system influences only one of the systems outputs, and vice-versa, every output of the closed-loop system is influenced by only one of its inputs as shown in figure 4.2

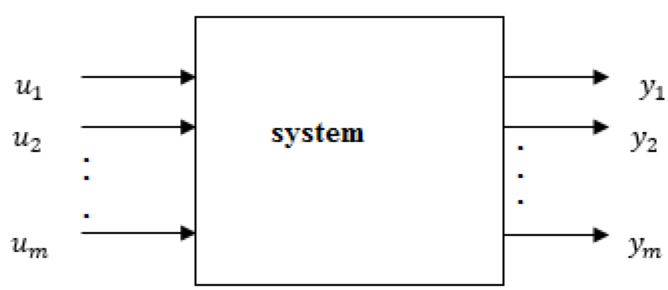


Figure 4.1 Open loop coupled system. [21]

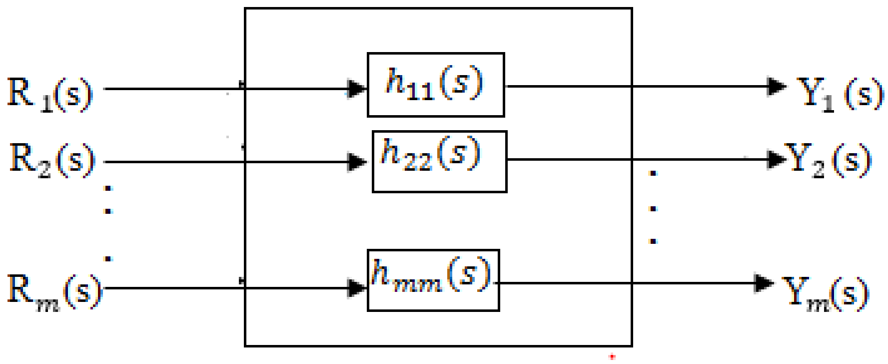


Figure 4.2 Open loop decoupled system [21]

The choice of a decoupling method is a relatively complex task since all techniques have their advantages and limitations. Simplified decoupling is by far the most popular method. Its main advantage is the simplicity of its elements. Ideal decoupling, which is rarely used in practice, greatly facilitates the tuning of the controller transfer matrix. Inverted decoupling, which is also rarely implemented, presents at the same time the main advantage of both the simplified and ideal decoupling methods.

Decoupling at the input of a two input-two output process $H(s)$ requires the design of a transfer matrix $D(s)$, such that $H(s)\, D(s)$ is a diagonal transfer matrix $T(s)$

$$H(s)D(s) = T(s) \tag{4.1}$$

$$D(s) = \begin{bmatrix} D_{11}(s) & D_{12}(s) \\ D_{21}(s) & D_{22}(s) \end{bmatrix} \tag{4.2}$$

$$H(s) = \begin{bmatrix} H_{11}(s) & H_{12}(s) \\ H_{21}(s) & H_{22}(s) \end{bmatrix} \tag{4.3}$$

$$T(s) = \begin{bmatrix} T_{11}(s) & 0 \\ 0 & T_{22}(s) \end{bmatrix} \tag{4.4}$$

$$T(s) = \begin{bmatrix} T_{11}(s) & 0 \\ 0 & T_{22}(s) \end{bmatrix} \tag{4.5}$$

$$D(s) = H(s)^{-1}T(s) = \frac{1}{H_{11}(s)H_{22}(s) - H_{12}(s)H_{21}(s)} \begin{bmatrix} H_{11}(s)T_{11}(s) & -H_{12}(s)T_{22}(s) \\ -H_{21}(s)T_{11}(s) & H_{22}(s)T_{22}(s) \end{bmatrix} \tag{4.6}$$

The elements $H_{11}(s)$,$H_{12}(s)$, $H_{21}(s)$ and $H_{22}(s)$ of equation (4.6), which represent the transfer functions of the process, are supposed to be known. The only unknown elements are $T(s)_{11}$ and $T(s)_{22}$. They represent the desired dynamics of the decoupled system.

4.1.1. Simplified decoupling

This choice makes the realization of the decoupler easy, but the diagonal transfer matrix $T(s)$ obtained is complex since its elements are the sum of transfer functions. Controller tuning can therefore be difficult. It is then often suggested to approximate each sum by a simpler transfer function to facilitate controller tuning [22].

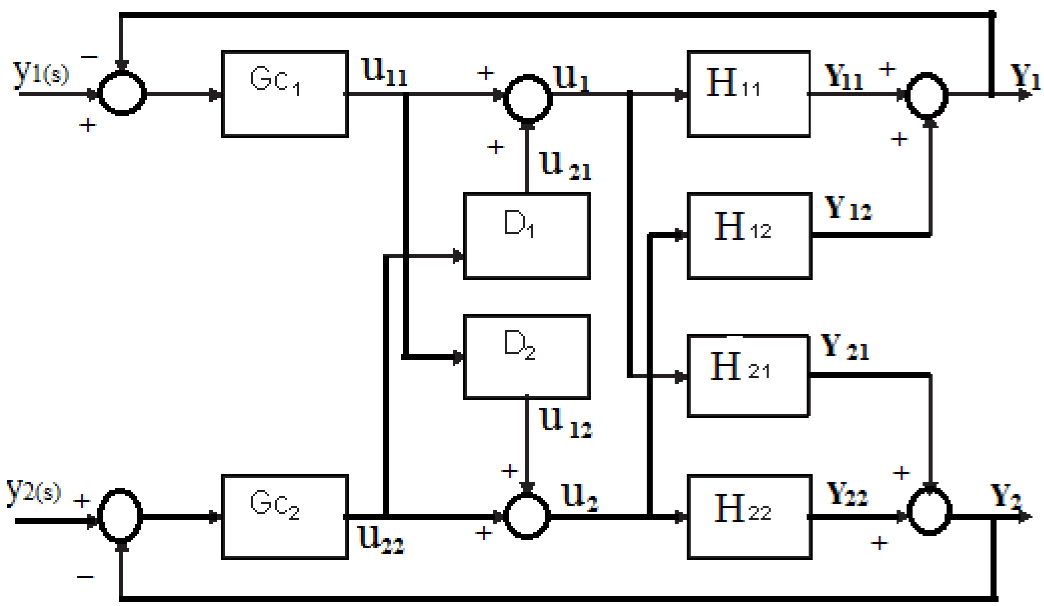


Figure 4.3 Block diagram of simplified decoupling [20]

This cancellation will occur at the y_1 summer if the decoupler output u_{21} satisfies

$$H_{21}u_{11} + H_{22}u_{21} = 0 \qquad (4.7)$$

Substituting for $u_{21} = D_1 u_{11}$ and factoring gives

$$(H_{21} + H_{22}D_1)u_{11} = 0 \qquad (4.8)$$

Note that $u_{11} \neq 0$ beecause u_{11} is control output that is time dependent. Then,

$$H_{21} + H_{22}D_1 = 0 \qquad (4.9)$$

Solving for D_1 gives,

$$D_1 = -\frac{H_{21}}{H_{22}} \tag{4.10}$$

$$D(s) = \begin{bmatrix} 1 & -\frac{\mathrm{H}_{12}(\mathrm{s})}{\mathrm{H}_{11}(\mathrm{s})} \\ -\frac{\mathrm{H}_{21}(\mathrm{s})}{\mathrm{H}_{22}(\mathrm{s})} & 1 \end{bmatrix} \tag{4.11}$$

The resulting transfer matrix *T*(*s*) is then:

$$T(s) = \begin{bmatrix} \mathrm{H}_{11}(\mathrm{s}) - \frac{\mathrm{H}_{12}(\mathrm{s})\mathrm{H}_{21}(\mathrm{s})}{\mathrm{H}_{22}(\mathrm{s})} & 0 \\ 0 & \mathrm{H}_{22}(\mathrm{s}) - \frac{\mathrm{H}_{12}(\mathrm{s})\mathrm{H}_{21}(\mathrm{s})}{\mathrm{H}_{11}(\mathrm{s})} \end{bmatrix} \tag{4.12}$$

.

Where

H(s) is plant transfer function

G_{1C} is controller one

G_{2C} is controller two

$Y_1 = c_a$

$Y_2 = T$

m_1 =q, feed concentration flow rate, = u1

$m_2 = T_c$, jacket temperature= u2

Input output relation for concentration control system and temperature control system are given by equation (4.15) and (4.16) respectively.

$$\mathrm{Y}_1 = \frac{1.4364s + 1.5393}{s^2 + 1.4682s + 0.5152}\mathrm{m}_1 + \frac{-0.0249}{s^2 + 1.4682s + 0.5152}\mathrm{m}_2 \tag{4.15}$$

$$\mathrm{Y}_2 = \frac{-13.171s - 12.5517}{s^2 + 1.4682s + 0.5152}\mathrm{m}_1 + \frac{0.3s + 0.34}{s^2 + 1.4682s + 0.5152}\mathrm{m}_2 \tag{4.16}$$

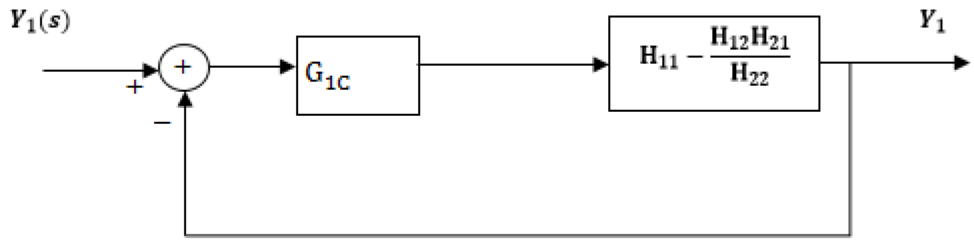


Figure 4.4 SISO concentration control system

Concentration plant transfer function

$$G_{1p} = H_{11} - \frac{H_{12}H_{21}}{H_{22}}$$

$$G_{p1} = \frac{1.4364s^2 + 4.2604s + 0.6726}{s^3 + 2.6014s^2 + 2.179s + 0.5837} \quad (4.17)$$

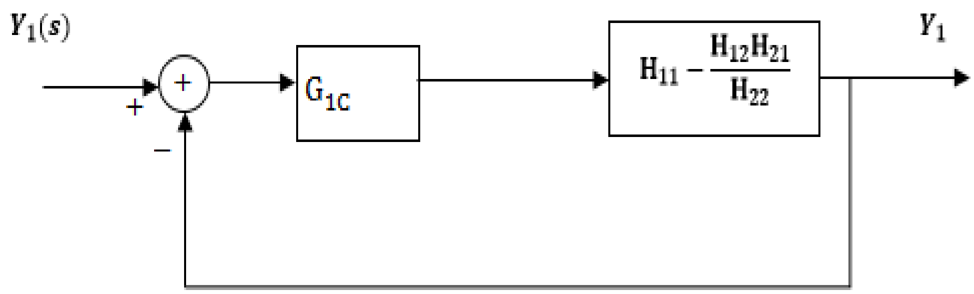


Figure 4.5 SISO temperature control system

Temperature plant transfer function

$$G_{2P} = H_{22} - \frac{H_{12}H_{21}}{H_{11}}$$

$$G_{2P} = \frac{0.3s^2 + 0.8898s + 0.1207}{s^3 + 2.5398s^2 + 2.0885s + 0.5520} \quad (4.18)$$

4.2. PID Controller

A proportional-integral-derivative controller (PID controller) is a control loop feedback mechanism (controller) widely used in industrial control systems. A PID controller calculates an error value as the difference between a measured process variable and a desired set point. The controller attempts to minimize the error by adjusting the process through use of a manipulated variable [15].

The PID controller algorithm involves three separate constant parameters, and is accordingly sometimes called three-term control: the proportional, the integral and derivative values, denoted P, I, and D. Simply put, these values can be interpreted in terms of time: P depends on the present error, I on the accumulation of past errors, and D is a prediction of future errors, based on current rate of change. The weighted sum of these three actions is used to adjust the process via a control element such as the position of a control valve, a damper, or the power supplied to a heating element [15, 17].

In the absence of knowledge of the underlying process, a PID controller has historically been considered to be the most useful controller. By tuning the three parameters in the PID controller algorithm, the controller can provide control action designed for specific process requirements. The response of the controller can be described in terms of the responsiveness of the controller to an error, the degree to which the controller overshoots the set point, and the degree of system oscillation. Note that the use of the PID algorithm for control does not guarantee optimal control of the system or system stability.

Some applications may require using only one or two actions to provide the appropriate system control. This is achieved by setting the other parameters to zero. A PID controller will be called a PI, PD, P or I controller in the absence of the respective control actions. PI controllers are fairly common, since derivative action is sensitive to measurement noise, whereas the absence of an integral term may prevent the system from reaching its target value due to the control action.

4.2.1 Proportional Term

The proportional term produces an output value that is proportional to the current error value. The proportional response can be adjusted by multiplying the error by a constant Kp, called the proportional gain constant. The proportional term is given by:

$$Pout = k_p e(t) \tag{4.19}$$

A high proportional gain results in a large change in the output for a given change in the error. If the proportional gain is too high, the system can become unstable. In contrast, a small gain results in a small output response to a large input error, and a less responsive or less sensitive controller. If the proportional gain is too low, the control action may be too small when responding to system disturbances. Tuning theory and industrial practice indicate that the proportional term should contribute the bulk of the output change

4.2.2 Integral Term

The contribution from the integral term is proportional to both the magnitude of the error and the duration of the error. The integral in a PID controller is the sum of the instantaneous error over time and gives the accumulated offset that should have been corrected previously. The accumulated error is then multiplied by the integral gain $\boldsymbol{k_i}$ and added to the controller output.

$$\boldsymbol{I_{out}} = \boldsymbol{k_i} \int \boldsymbol{e}(\tau)\mathrm{d}\tau \tag{4.20}$$

The integral term accelerates the movement of the process towards set-point and eliminates the residual steady-state error that occurs with a pure proportional controller. However, since the integral term responds to accumulated errors from the past, it can cause the present value to overshoot the set-point value.

4.2.3 Derivative Term

The derivative of the process error is calculated by determining the slope of the error over time and multiplying this rate of change by the derivative gain Kd. The magnitude of the contribution of the derivative term to the overall control action is termed the derivative gain $\boldsymbol{k_d}$. The derivative term is given by

$$D_{out} = \frac{d}{dt} e(t) \tag{4.21}$$

Derivative action predicts system behavior and thus improves settling time and stability of the system. An ideal derivative is not causal, so that implementations of PID controllers include an additional low pass filtering for the derivative term, to limit the high frequency gain and noise. Derivative action is seldom used in practice though - by one estimate in only 20% of deployed controllers- because of its variable impact on system stability in real-world applications.

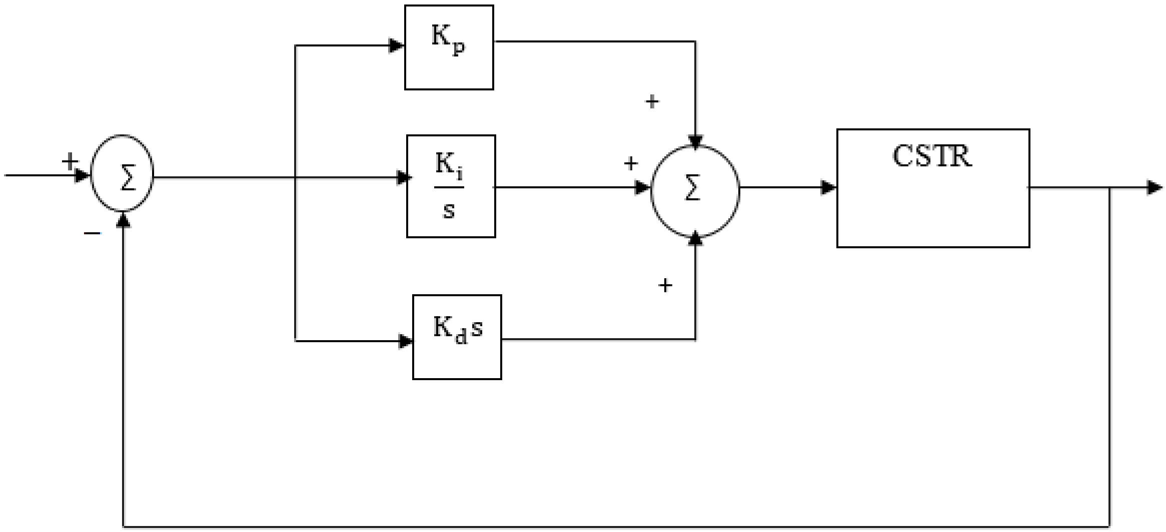


Figure 4. 6 Control loop with PID controller [15]

For the control process, better performance can be achieved by tuning the control loop, which is adjusting the control parameters to satisfy the desired control response. For PID controller, each of the three parameters (i.e.K_p,K_i and K_d) has different effect on system control which is summarized in Table 4.2 based on the situation of increasing the controller parameter individually.

Table 3.2 Effects caused by increasing the PID control parameter individually

PID Control Parameters	Rise Time (t_r)	Over Shoot (M_p)	Settling Time (t_s)	Steady State Error(e_{ss})
K_p	Decrease	Increase	Small Change	Decrease
K_i	Decrease	Increase	Increase	Eliminate
K_d	Small Change	Decrease	Decrease	Small Change

4.3. Model Reference Adaptive Control

The standard implementation of MRAC based systems contains the four key blocks. The reference model defines the desired performance characteristics of the process being controlled. The adaptation law uses the error between the process and the model output, the process output and input signal to vary the parameters of the control system. These parameters are varied so as to minimize the error between the process and the reference model.

The control system can be anything from a simple gain based controller to a more complicated parameter based transfer function or plant matrix. Whatever type of control system is used, the parameters of the controller must be varied by the adaptation law. The final element of the MRAC system is the process that is being controlled.

Adaptation law

One particular way of handling this problem is the technique of "Model Reference Adaptive Systems" (MRAS). This involves the definition of a reference process model whose dynamics in response to a reference input should be followed by the plant process. For the plant process with unknown parameters, a specific control law alters the reference input signal in order for the plant's output signal to match the one of the reference model. This control law usually features time-variant controller parameter θ which reflect the algorithm's adaptation to the given plant system.

4.3.1 MIT rule

A commonly used update law is the "MIT rule". This adaptation mechanism is defined by the controller update rule

$$\frac{d\theta}{dt} = -\alpha e y_m \tag{4.22}$$

and incorporates the output error e as well as the reference model's output y_m. To derive this update law, a cost function is defined first:

$$J = \frac{1}{2}e^2 \tag{4.23}$$

Essentially, the parameter θ is set to follow the "steepest descent" i. e. move along the negative gradient with a certain adaptation gain α :

$$\frac{d\theta}{dt} = -\,\alpha \frac{d}{d\theta} J \tag{4.24}$$

$$= -\,\alpha e \frac{d}{d\theta} e(\theta) \tag{4.25}$$

In order to calculate the sensitivity derivative $\frac{d}{d\theta}e(\theta)$, the system equations of model and plant process are inserted:

$$\frac{d\theta}{dt} = -\alpha e \frac{d}{d\theta}(y_p\text{-}y_m)$$
$$= -\alpha e \frac{d}{d\theta}(L^{-1}\{G_P(s)U(s) - G_m(s)R(s)\})$$
$$= -\alpha e \frac{d}{d\theta}(L^{-1}\{G_P(s)\theta R(s) - G_m(s)R(s)\})$$
$$=- \alpha e \frac{d}{d\theta}(L^{-1}\{G_m(s)b\theta R(s) - G_m(s)R(s)\})$$
$$=- \alpha e b L^{-1}\{G_m(s)R(s)\}$$
$$=- \alpha e b y_m \tag{4.26}$$

Where

y_m is model response

$G_m(s)$ is model transfer function

y_p is plant response

$G_p(s)$ is plant transfer function

$U(s)$ is controller

$R(s)$ is reference input

A parameter update law for the system is now derived based on the MIT rule for both θ_1 and θ_2 As in the previous section the approach

$$\frac{d\theta}{dt} = -\alpha e \frac{d}{d\theta}e(\theta) \tag{4.27}$$

Second order system with two unknown parameters

Let the second order plant transfer function is

$$y_{plant} = \frac{b_o}{s^2+b_1s+b_2} \quad \text{and}$$

The second order model reference transfer is

$$y_m = \frac{a_{1m}s + a_{0m}}{s^2 + a_{1m}s + a_{0m}}$$

$$error = e = y_{plant} - y_{model} \tag{4.28}$$

$$J(\theta) = \frac{1}{2} e^2(\theta) \tag{4.29}$$

$$\frac{d\theta}{dt} = -\gamma \frac{\delta J}{\delta \theta} = -\gamma e \frac{\delta e}{\delta \theta} \tag{4.30}$$

Let controller takes the form:

$$u = \theta_1 u_c - \theta_2 (u_c - y_{plant}) \tag{4.31}$$

$$e = y_{plant} - y_{model} = G_p u - G_m u_c \tag{4.32}$$

$$y_{plant} = G_p u = \frac{b_o}{s^2+b_1 s+b_2} \left(\theta_1 u_c - \theta_2 (u_c - y_{plant})\right) \tag{4.33}$$

$$y_{plant} = \frac{b_o u_c (\theta_1 - \theta_2)}{s^2+b_1 s+b_2 - b_o \theta_2} \tag{4.34}$$

$$e = \frac{b_o u_c (\theta_1 - \theta_2)}{s^2+b_1 s+b_2 - b_o \theta_2} - G_m u_c \tag{4.35}$$

$$\frac{\partial e}{\partial \theta_1} = \frac{b_o}{s^2+b_1 s+b_2 - b_o \theta_2} u_c \tag{4.36}$$

Since the parameters give perfect model following, we can use the approximations of

$$s^2 + b_1 s + b_2 + b_o \theta_2 \approx s^2 + a_{1m} s + a_{0m} \tag{4.37}$$

$$\frac{\partial e}{\partial \theta_1} = \frac{a_{1m} s + a_{0m}}{s^2 + a_{1m} s + a_{0m}} u_c \tag{4.38}$$

$$\frac{\partial e}{\partial \theta_1} = G_m u_c \tag{4.39}$$

$$\frac{\partial e}{\partial \theta_2} = \frac{-b_o u_c (s^2+b_1 s+b_2 - b_o \theta_2) + {b_o}^2 u_c (\theta_1 - \theta_2)}{(s^2+b_1 s+b_2 - b_o \theta_2)^2} \tag{4.40}$$

By simplifying the above equation (4.40)

$$= -\frac{a_{1m} s + a_{0m}}{s^2 + a_{1m} s + a_{0m}} (u_c - y_{plant}) \tag{4.41}$$

Because

$$s^2 + b_1 s + b_2 + b_o \theta_2 \approx s^2 + a_{1m} s + a_{0m}$$

If reference model is close to plant and we can approximate:

$$\frac{\partial e}{\partial \theta_2} = -\, G_m \left(u_c - \; y_{plant}\right) \tag{4.42}$$

From MIT rule, update rules are then:

$$\frac{d\theta_1}{dt} = -\gamma e \frac{\delta e}{\delta \theta_1} = -\gamma \left(\frac{a_{1m}s + a_{0m}}{s^2 + a_{1m}s + a_{0m}}\; u_c\right) e \tag{4.43}$$

$$= \; -\gamma G_m u_c e$$

$$\frac{d\theta_2}{dt} = -\gamma e \frac{\delta e}{\delta \theta_2} = \gamma\; G_m \left(u_c - \; y_{plant}\right) e \tag{4.44}$$

4.4 Reactor temperature control

Products within reactors usually liberate or absorb heat during processing. The amount of heat generated by an exothermic reactor increases as the reaction temperature rises. An increase in the reaction temperature will also increase heat removal, because of the increase in (ΔT) between process and coolant temperature. If an increase in a reaction temperature result in a greater increase in heat generation than in heat removal, the process is said to display positive feedback; as such it is considered to be (unstable in the open loop). In order to hold the reactor contents at the desired temperature, heat has to be added or removed by a cooling jacket or cooling pipe. Heating/cooling coils or external jackets are used for heating and cooling reactors. Heat transfer fluid passes through the jacket or coils to add or remove heat.

In order to design temperature control, first we must define the second order model that gives the characteristics of the system we want to have and the plant must be follow to the model by varying the adaption gain.

$$G_m(\mathrm{s}) = \frac{{\omega_n}^2}{s^2 + 2\omega_n \delta s + {\omega_n}^2}$$

We can develop second order model reference that has the requirement performance characteristics damping ratio $\delta = 0.93,$ natural frequency $\omega_n = 18.08$, maximum overshoot (Mp) of 0%, settling time (Ts) of 0.178 second and 2% criterion. The transfer function for the model is

$$G_m(\mathrm{s}) = \frac{327.1088}{s^2 + 33.7078s + 327.1088}$$

Plant of temperature is

$$\mathrm{G_{p2}} = \mathrm{H_{22}} - \frac{\mathrm{H_{12}H_{21}}}{\mathrm{H_{11}}}$$

$$G_{p2} = \frac{0.3s^2 + 0.8898s + 0.1207}{s^3 + 2.5398s^2 + 2.0885s + 0.5520}$$

and

$$G_m(\mathrm{s}) = \frac{327.1088}{s^2 + 33.7078s + 327.1088}$$

Then the value of $\theta_1$and θ_2 can be calculated from the eq (4.43) and eq (4.44).

$$\frac{d\theta_1}{dt} = -\gamma e \frac{\delta e}{\delta\theta_1} = -\gamma \left(\frac{a_{1m}s + a_{0m}}{s^2 + a_{1m}s + a_{0m}}\ u_c\right) e$$

$$\frac{d\theta_1}{dt} = -\gamma (\frac{327.1088}{s^2 + 33.7078s + 327.1088}) u_c\ e$$

After integrating both sides, the θ_1 value is given by

$$\theta_1 = \frac{-\gamma}{s} (\frac{327.1088}{s^2 + 33.7078s + 327.1088}) u_c\ e$$

$$\frac{d\theta_2}{dt} = -\gamma e \frac{\delta e}{\delta\theta_2} = \gamma\ G_m\ (u_c - \ y_{plant}) e$$

$$\frac{d\theta_2}{dt} = -\gamma\ \frac{327.1088}{s^2 + 33.7078s + 327.1088} \left(u_c - \frac{0.3s^2 + 0.8898s + 0.1207}{s^3 + 2.5398s^2 + 2.0885s + 0.5520}\right) e$$

After integrating both sides, the θ_2 value is given by

$$\theta_2 = \frac{\gamma}{s} \frac{327.1088}{s^2 + 33.7078s + 327.1088} \left(u_c - \frac{0.3s^2 + 0.8898s + 0.1207}{s^3 + 2.5398s^2 + 2.0885s + 0.5520}\right) e$$

4.5. Reactor Concentration control

In order to design concentration control, first we must define the second order model that gives the characteristics of the system we want to have and the plant must be follow to the model by varying the adaption gain.

$$G_m(\mathrm{s}) = \frac{327.1088}{s^2 + 33.7078s + 327.1088}$$

For the reactor concentration, we can develop second order model reference that has the requirement performance characteristics damping ratio $\delta = 0.93,$ natural frequency $\omega_n = 18.08$, maximum overshoot (Mp) of 0%, settling time (Ts) of 0.178 second and 2% criterion.

Plant of concentration is

$$\mathrm{G_{p1}} = \mathrm{H_{11}} - \frac{\mathrm{H_{12}H_{21}}}{\mathrm{H_{22}}}$$

$$\boldsymbol{G_{p1}} = \frac{1.4364s^2 + 4.2604s + 0.6726}{s^3 + 2.6014s^2 + 2.179s + 0.5837}$$

And

$$G_m(\mathrm{s}) = \frac{327.1088}{s^2 + 33.7078s + 327.1088}$$

Then the value of $\theta_1$and θ_2 can be calculated from the eq (4.43) and eq (4.44).

$$\frac{d\theta_1}{dt} = -\gamma e \frac{\delta e}{\delta \theta_1} = -\gamma \left(\frac{a_{1m}s + a_{0m}}{s^2 + a_{1m}s + a_{0m}}\ u_c \right) e$$

$$\frac{d\theta_1}{dt} = -\gamma (\frac{327.1088}{s^2 + 33.7078s + 327.1088}) u_c\ e$$

After taking Laplace transform both sides, the θ_1 value is given by

$$\theta_1 = \frac{-\gamma}{s}\left(\frac{327.1088}{s^2 + 33.7078s + 327.1088}\right)u_c\, e$$

$$\frac{d\theta_2}{dt} = -\gamma e \frac{\delta e}{\delta\theta_2} = \gamma\, G_m\, (u_c - \, y_{plant})e$$

$$\frac{d\theta_2}{dt} = -\gamma e \frac{\delta e}{\delta\theta_2} = \gamma\, \frac{327.1088}{s^2 + 33.7078s + 327.1088}\left(u_c - \, \frac{1.4364s^2 + 4.2604s + 0.6726}{s^3 + 2.6014s^2 + 2.179s + 0.5837}\right)e$$

After taking Laplace transform both sides, the θ_2 value is given by

$$\theta_2 = \frac{\gamma}{s}\frac{327.1088}{s^2 + 33.7078s + 327.1088}\left(u_c - \, \frac{1.4364s^2 + 4.2604s + 0.6726}{s^3 + 2.6014s^2 + 2.179s + 0.5837}\right)e$$

CHAPTER FIVE

5. RESULT AND DISCUSSION

In this chapter the simulation results of the system will be discussed in brief. First a simulation result for the two controlled variables without any controller i.e. open loop response is discussed. Then the results of PID controller for each controlled variable is presented. Finally, the simulation result for MRAC is discussed and compared with the results of PID controller.

5.1 Open loop Response

5.1.1 Response of SISO concentration without any controller (open loop response)

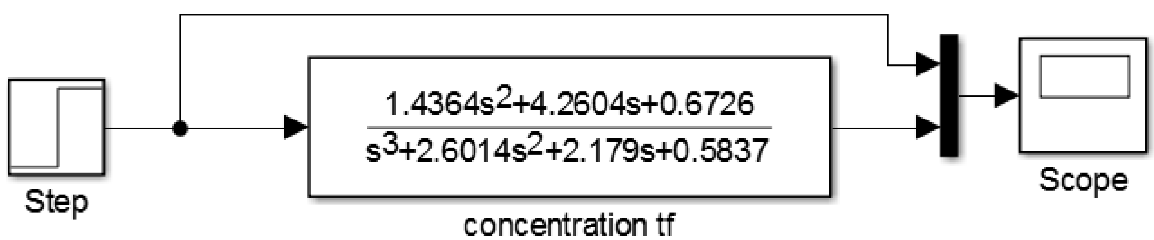


Figure 5.1 Block diagram of open loop SISO concentration

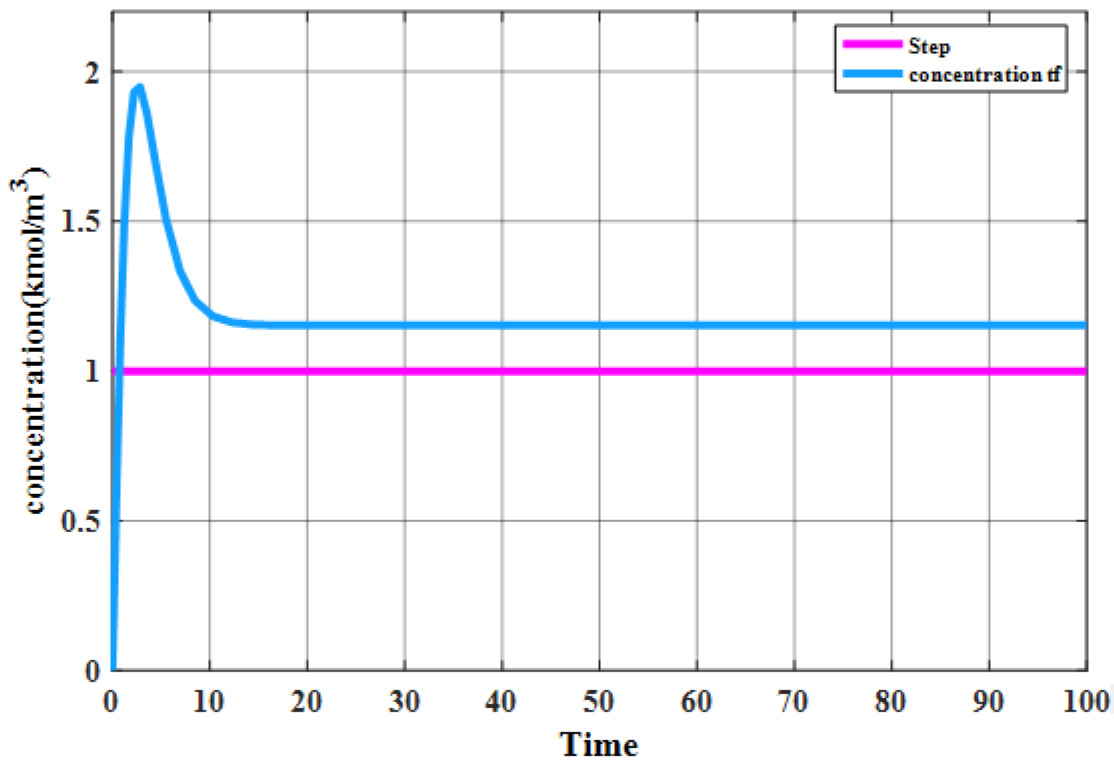


Figure 5.2 Unit step response of open loop SISO concentration

5.1.2. Response of SISO temperature without any controller (open loop response)

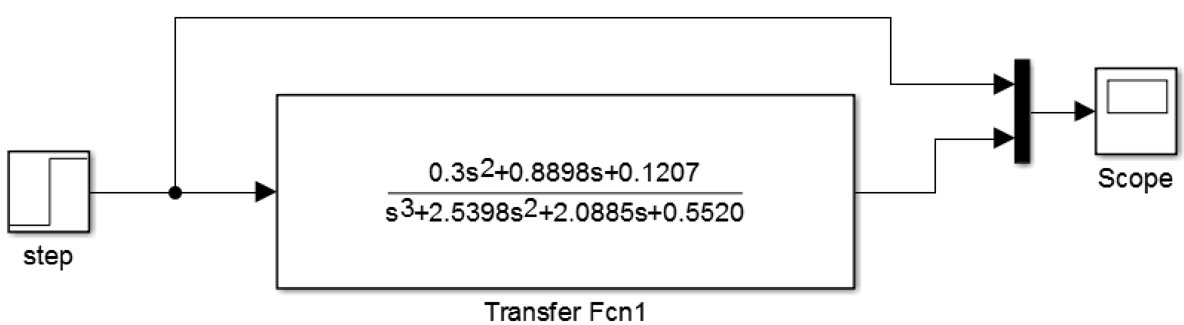


Figure 5.3 Block diagram of open loop SISO temperature

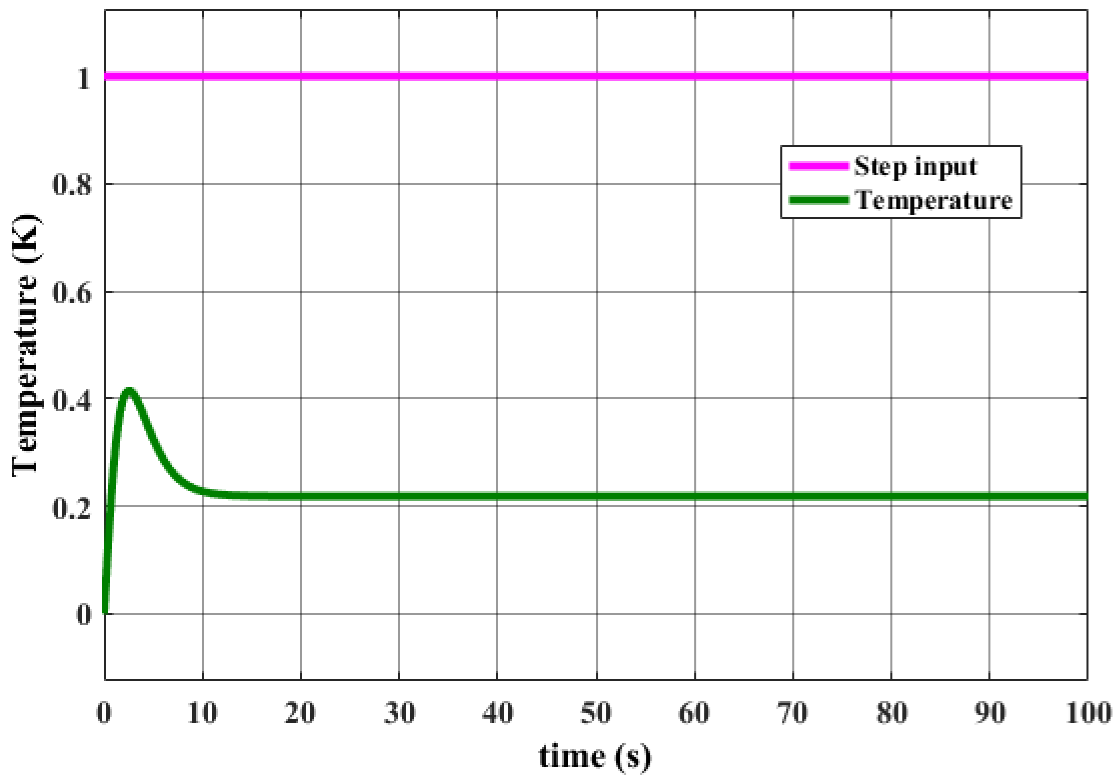


Figure 5.4 Unit step response of open loop SISO temperature

5.2. Results for conventional PID Controller

In this thesis PID controller uses as benchmark to compare its performance with performance of the MRAS controller.

5.2.1 Step response of SISO concentration under PID controller

In this thesis the PID controller gains are tuned automatically in MATLAB Simulink by developing the block diagrams for different ranges of robustness.

The closed loop plant for SISO pressure with PID controller is developed in MATLAB Simulink as shown in figure 4.5. The PID controller gains are tuned automatically in MATLAB and the values of the gains for 0.9 robustness are k_p =1.771, k_i = 2.141, k_d =0.215 and N=7.335.

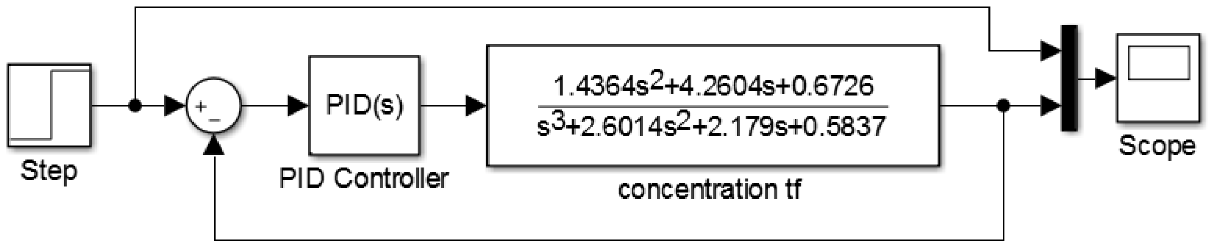


Figure 5.5 SISO concentration block diagram with conventional PID controller

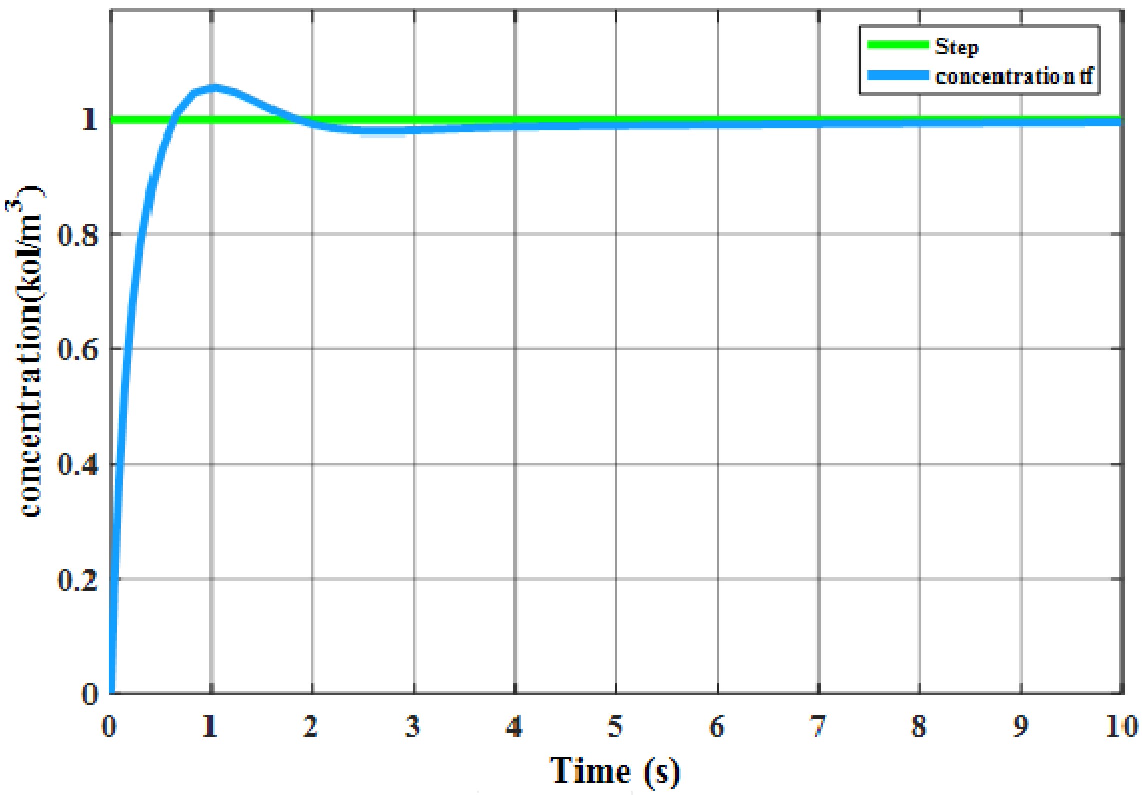


Figure 5.6 SISO response of concentration with conventional PID controller

5.2.2. Response of SISO temperature for conventional PID controller

The closed loop plant for SISO Power with PID controller is developed in MATLAB Simulink as shown in figure 4.6. The PID controller gains are tuned automatically in MATLAB and the values of the gains for a robustness of 0.9 are k_p= 1.533, k_i= 0.4727, k_d = -1.116 and filter coefficient (N) = 1.373.

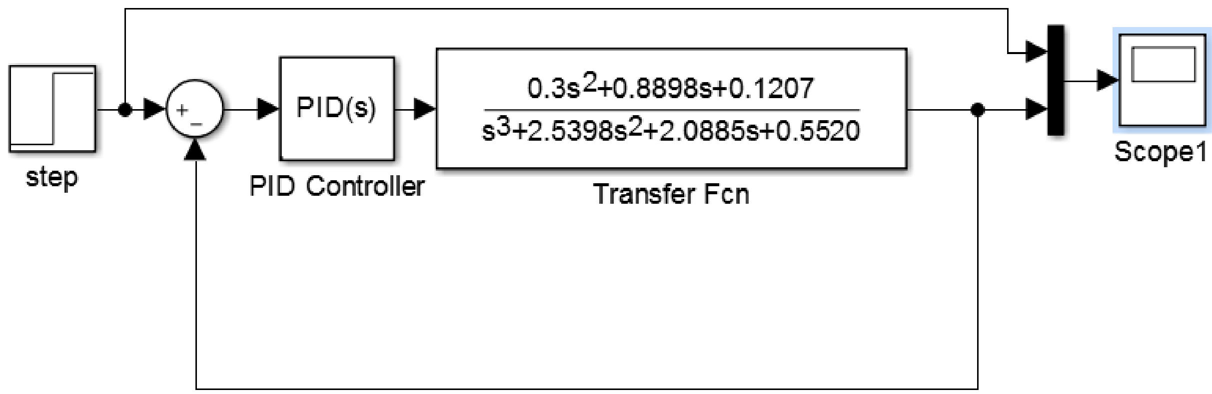


Figure 5.7 SISO temperature block diagram with conventional PID controller

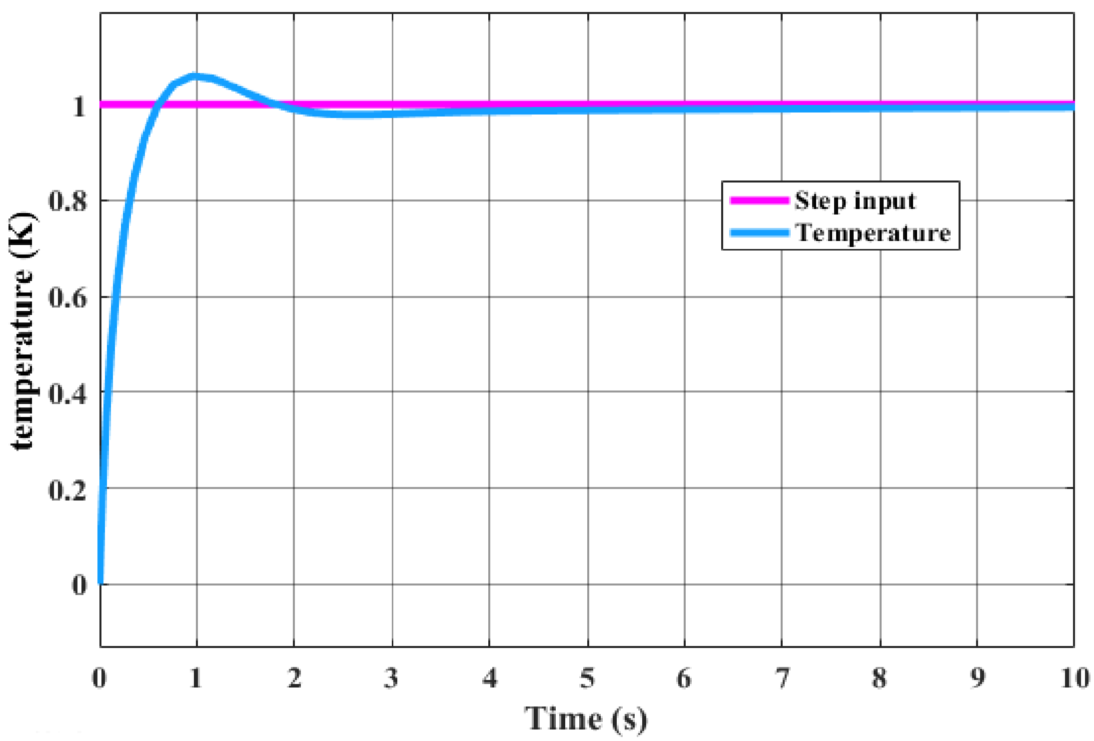


Figure 5. 8 SISO response of temperature with conventional PID controller

5.3. Results of MRAC

After decoupling the continuous stirred tanker reactor (CSTR), for SISO concentration and temperature transfer function is obtained and PID controller is designed. But the PID controller is less performance in terms of time domain performance. So MRAC is developed for each controlled variables.

The first step in MRAC is to select a reference model in which the plant is required to follow. So in order to have a plant with good performance characteristics the reference model must have good performance characteristics. Considering this implication, the system given below is taken as a reference model. Reference model is given from mathematical relationship of second order transfer function time domain specifications. Their value are rise time = 0.101 sec, overshoot = 0, settling time = 0.178 sec, steady state error = 0.

$$Y_m(\mathrm{s}) = \frac{327.1088}{s^2+33.7078s+327.1088} * U_c$$

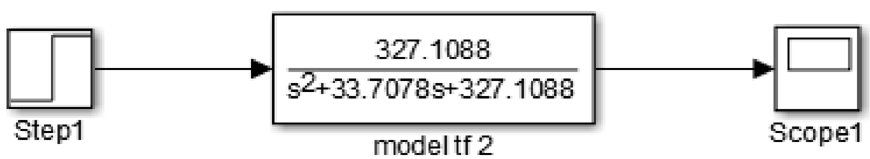


Figure 5.9 Model reference block diagram with unit step input

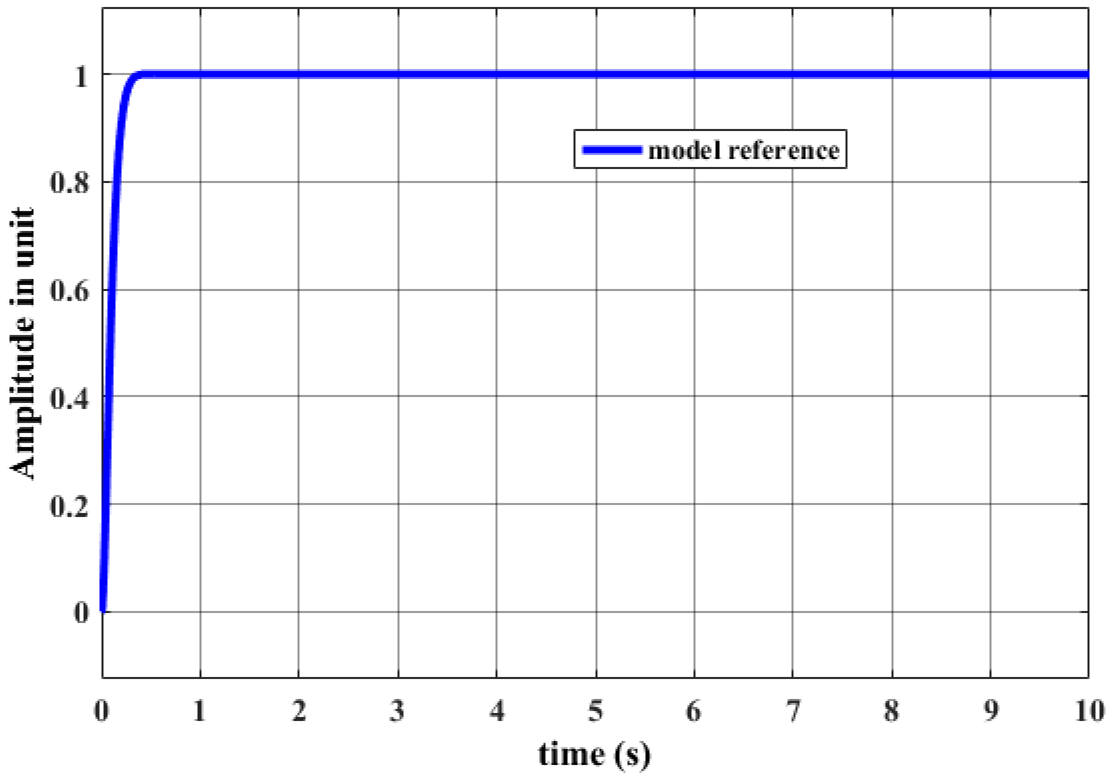


Figure 5.10 Step response of reference model.

5.3.1. MRAC for SISO concentration control

The linearized transfer function of SISO concentration is given as

$$SISO \ \text{concentration} = G_{p1} = \frac{1.4364s^2 + 4.2604s + 0.6726}{s^3 + 2.6014s^2 + 2.179s + 0.5837}$$

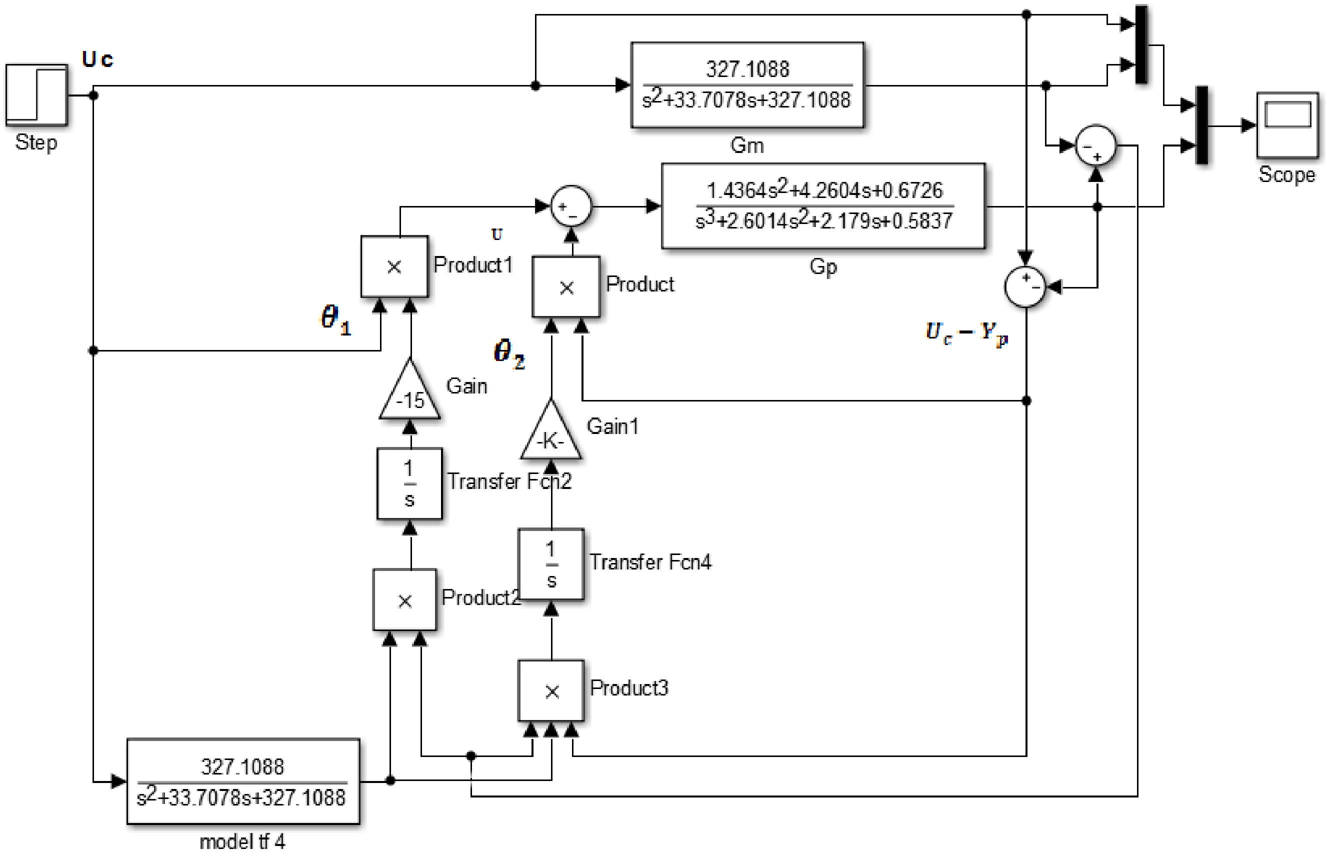


Figure 5.11MRAC controller for concentration developed in MATLAB Simulink

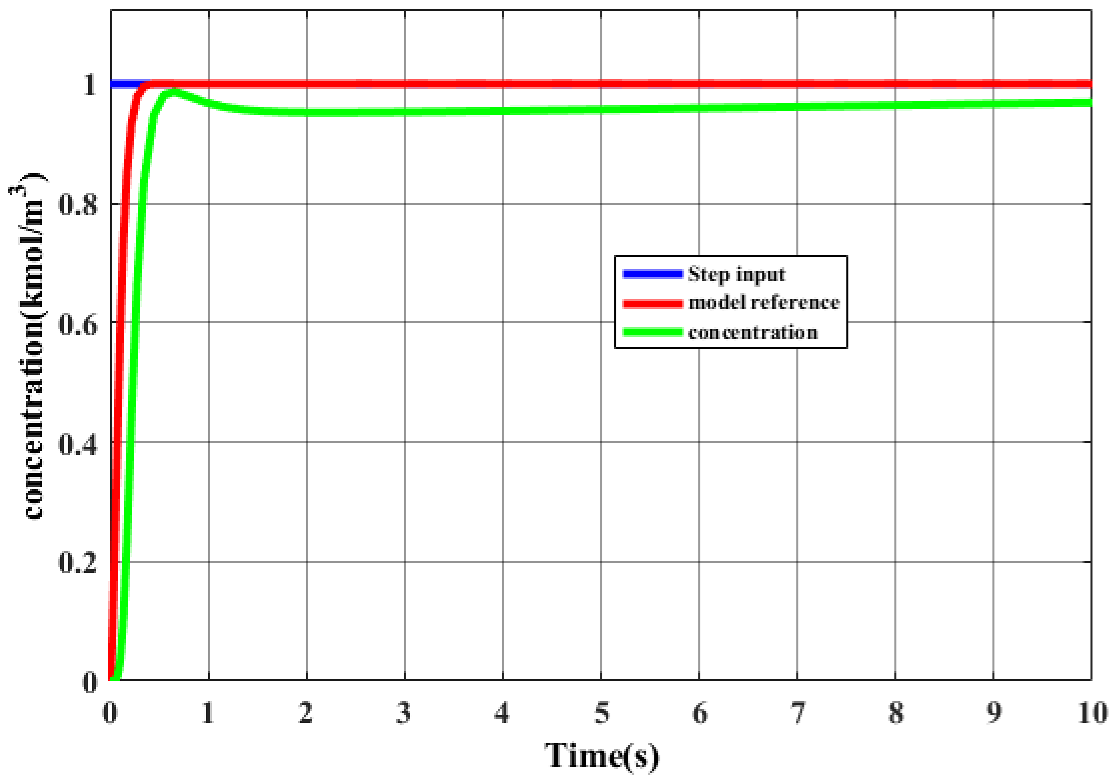


Figure 5.12 Step response of concentration using MRAS with adaptation gains of
$\gamma_1 = 1 \; and \; \gamma_2 = 100$

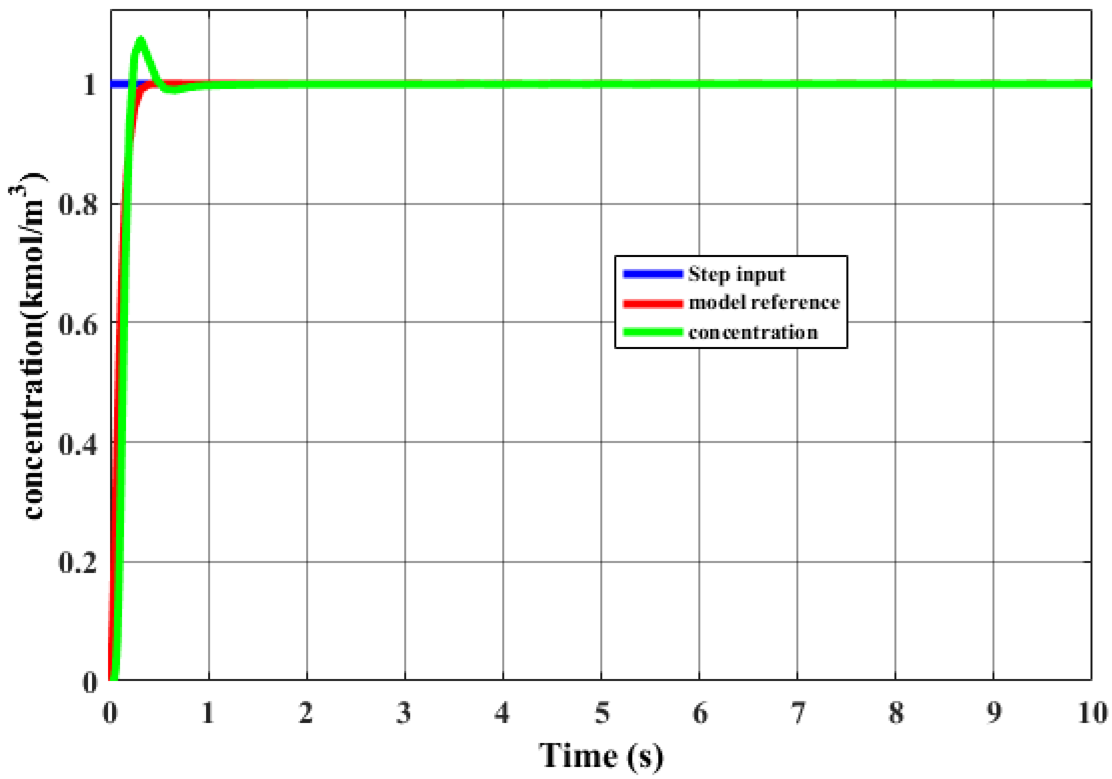


Figure 5.13 Step response of concentration using MRAS with adaptation gains of $\gamma_1 = 100\ and\ \gamma_2 = 1000$

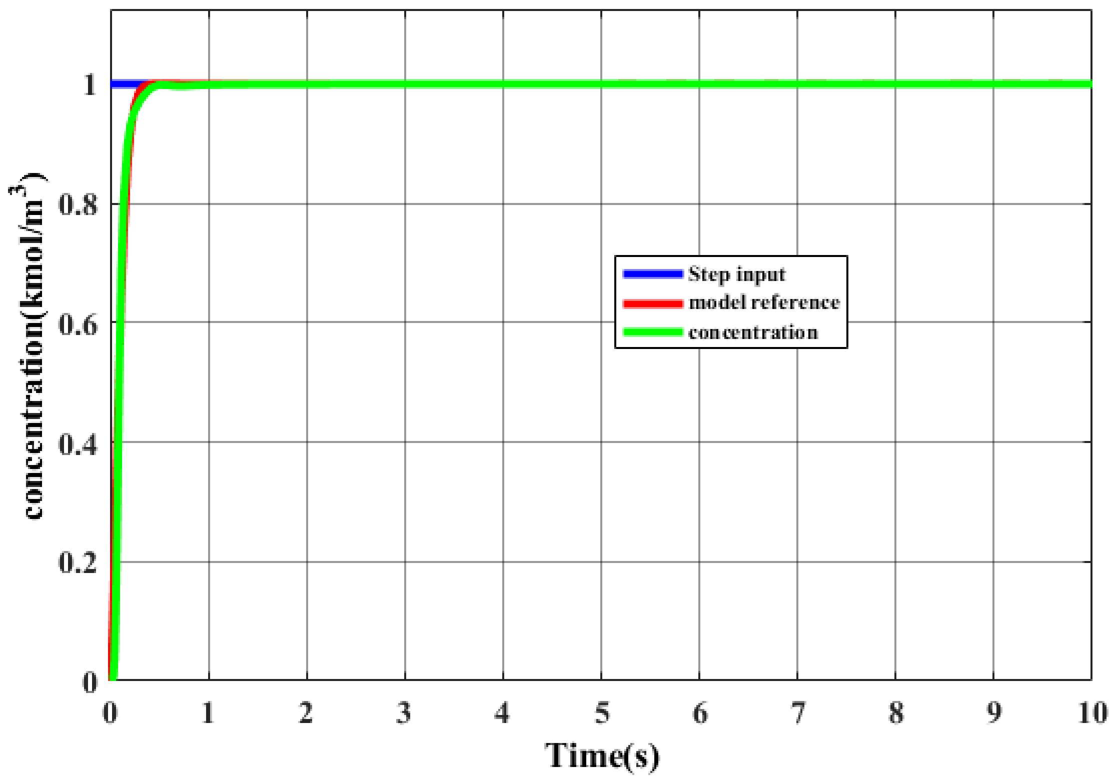


Figure 5.14 Step response of concentration using MRAS with adaptation gains of $\boldsymbol{\gamma_1 = 150 \; and \; \gamma_2 = 7000}$

Generally the accepting rang of adaption gain from the simulation show that $\boldsymbol{\gamma_1}$ is from 100 to 150 and $\boldsymbol{\gamma_2}$ is from 1000 to 7000.

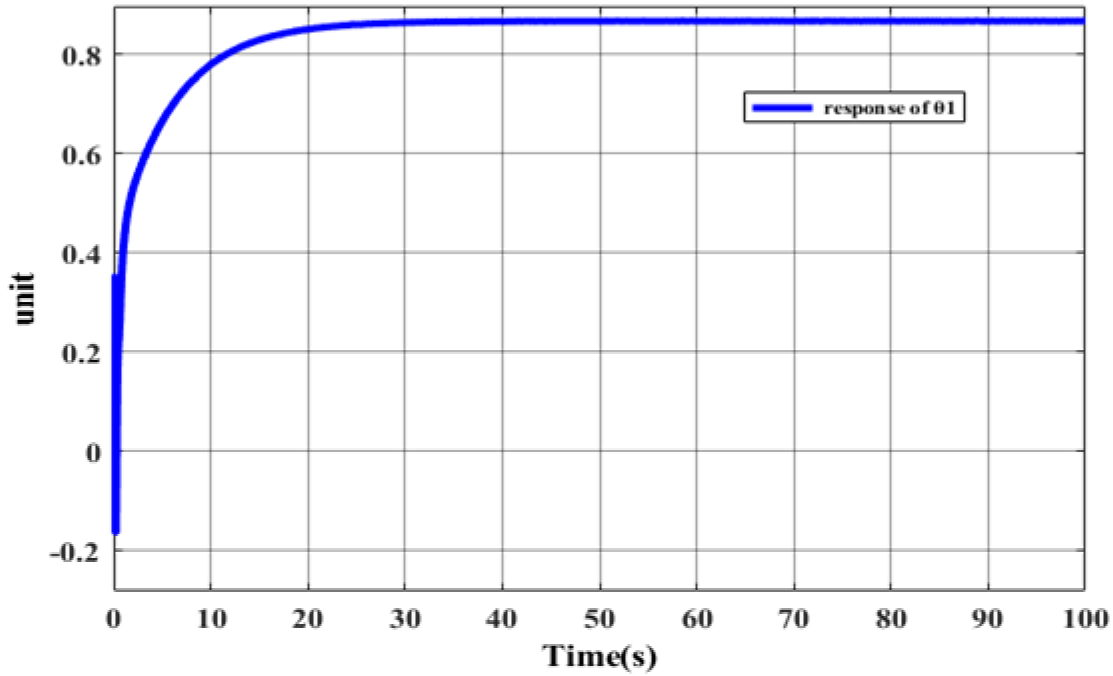


Figure 5.15 Response of $\boldsymbol{\theta_1}$ for concentration

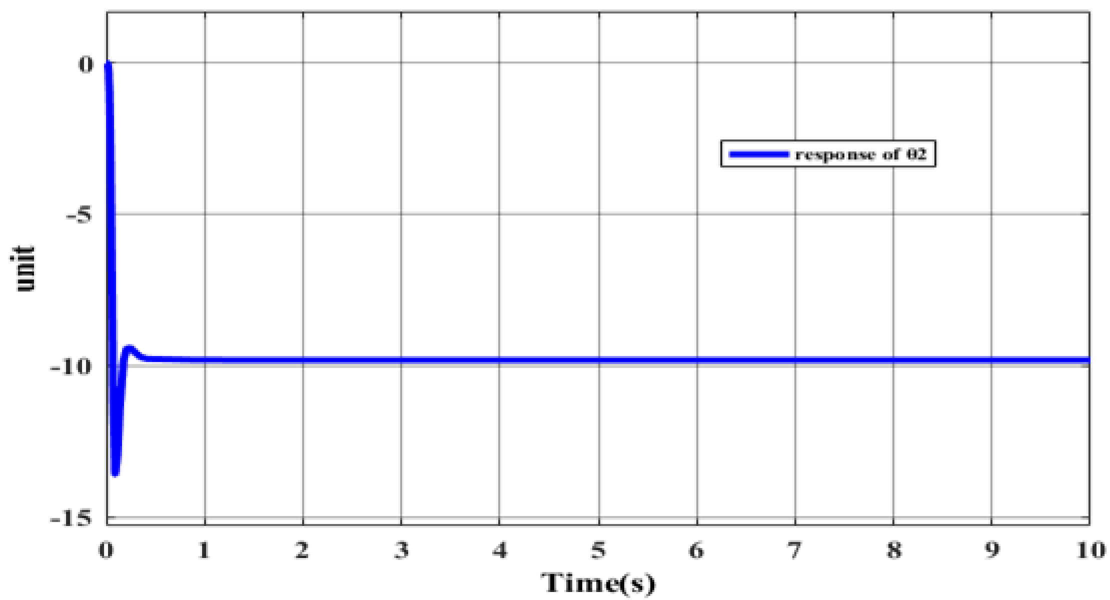


Figure 5.16 Response of $\boldsymbol{\theta_2}$ for concentration

5.3.2. MRAC for SISO temperature control

The approximated transfer function of SISO concentration is given as

$$SISO\ \text{temperature} = \boldsymbol{G_{p2}} = \frac{0.3s^2 + 0.8898s + 0.1207}{s^3 + 2.5398s^2 + 2.0885s + 0.5520}$$

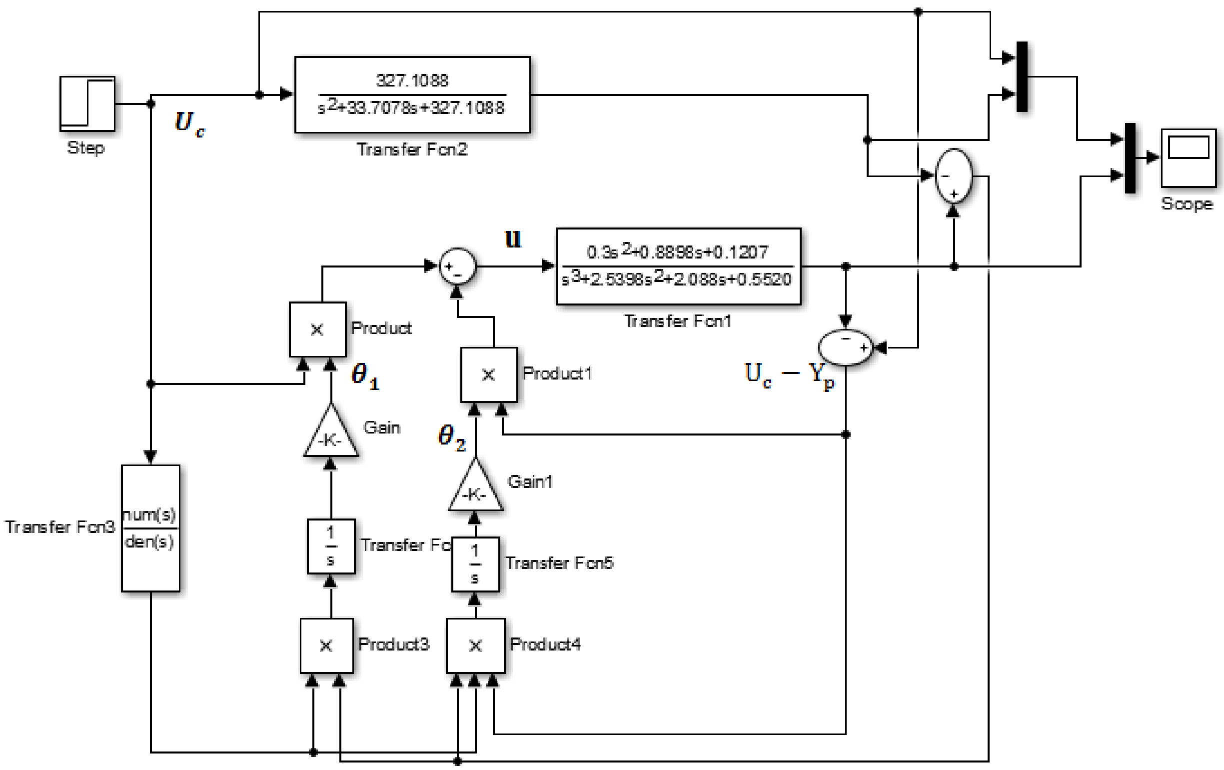


Figure 5.17 MRAC controller for temperature developed in MATLAB Simulink

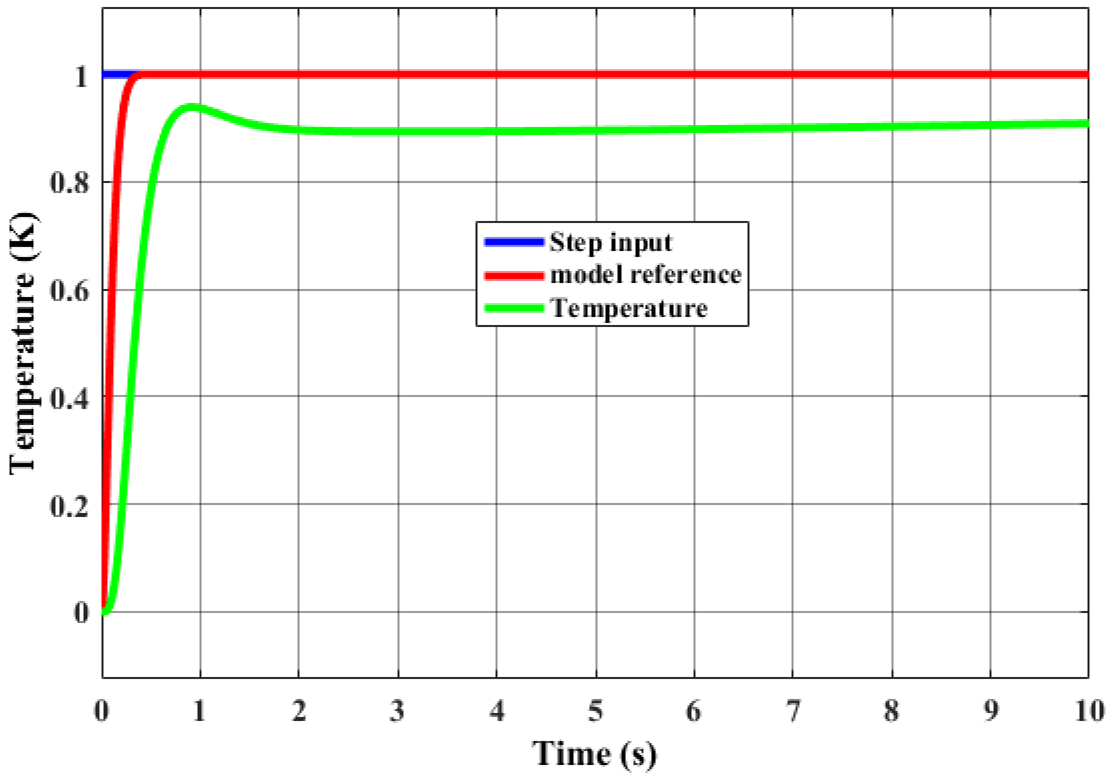


Figure 5.18 Step response of temperature using MRAS with adaptation gains of $\gamma_1 = 1$ *and* $\gamma_2 = 100$

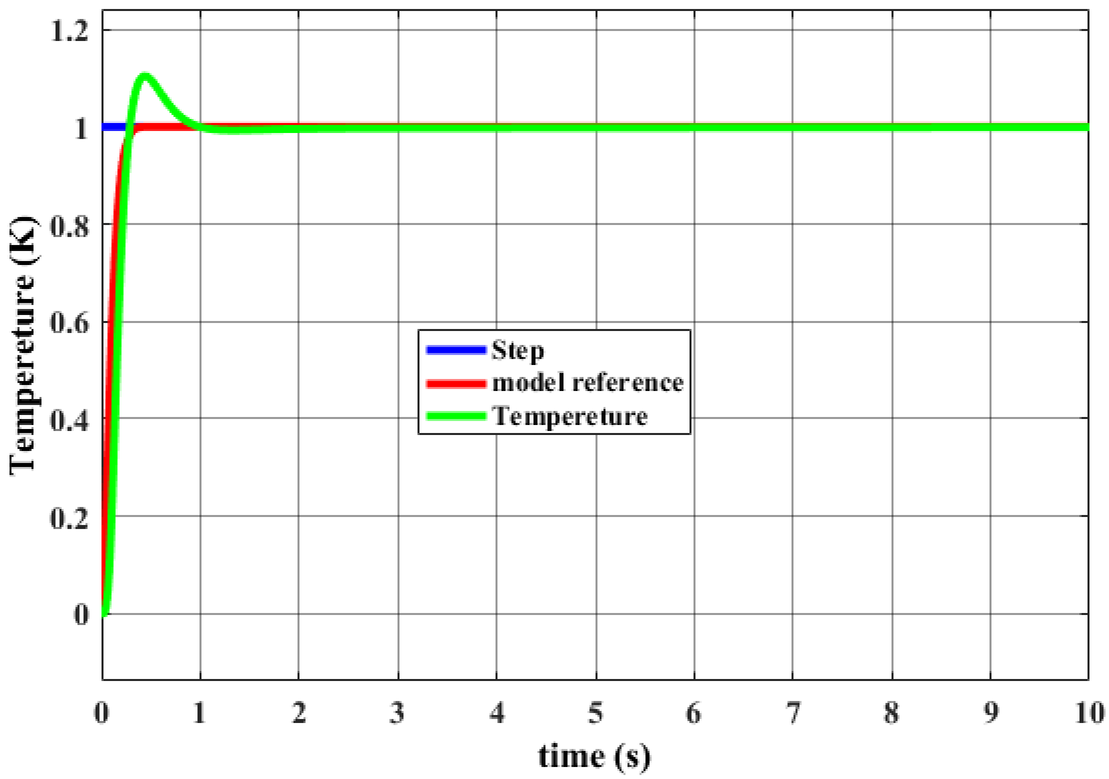


Figure 5. 19 step response of temperature using MRAS with adaptation gains of $\gamma_1 = 100\ and\ \gamma_2 = 1000$

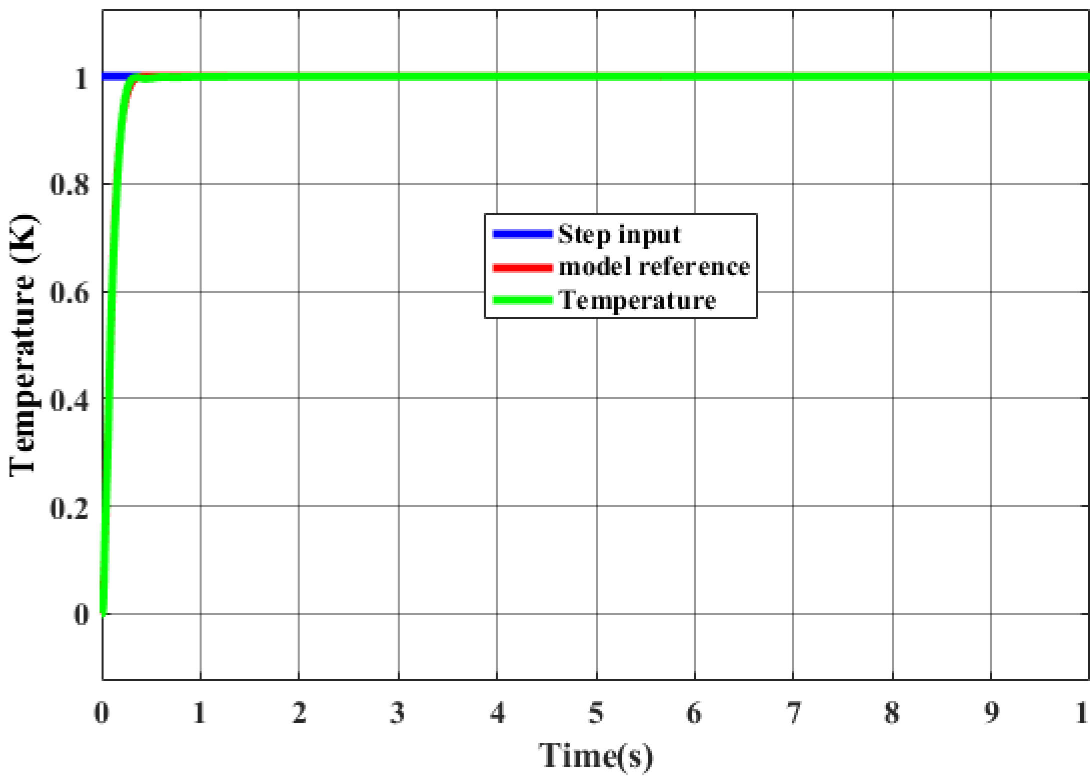


Figure 5.20 Step response of temperature using MRAS with adaptation gains of $\boldsymbol{\gamma_1} = \mathbf{1400}\ \boldsymbol{and}\ \boldsymbol{\gamma_2} = \mathbf{90000}$

Generally the accepting rang of adaption gain from the simulation show that $\boldsymbol{\gamma_1}$ is from 100 to 1400 and $\boldsymbol{\gamma_2}$ is from 1000 to 90000.

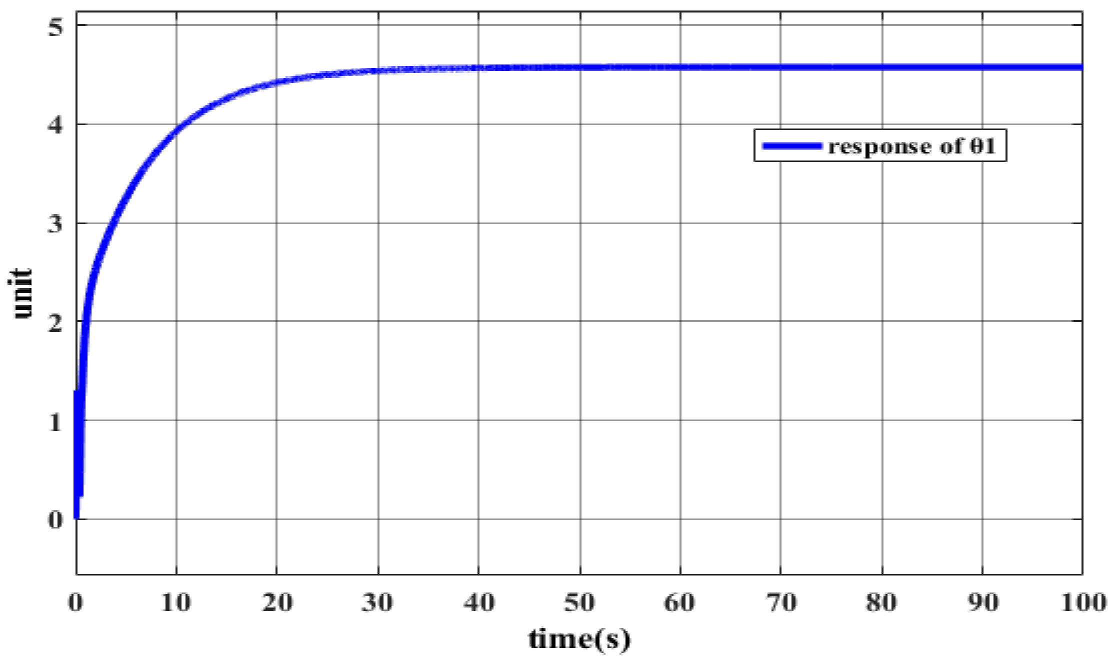


Figure 5.21 Response of $\boldsymbol{\theta_1}$ for temperature

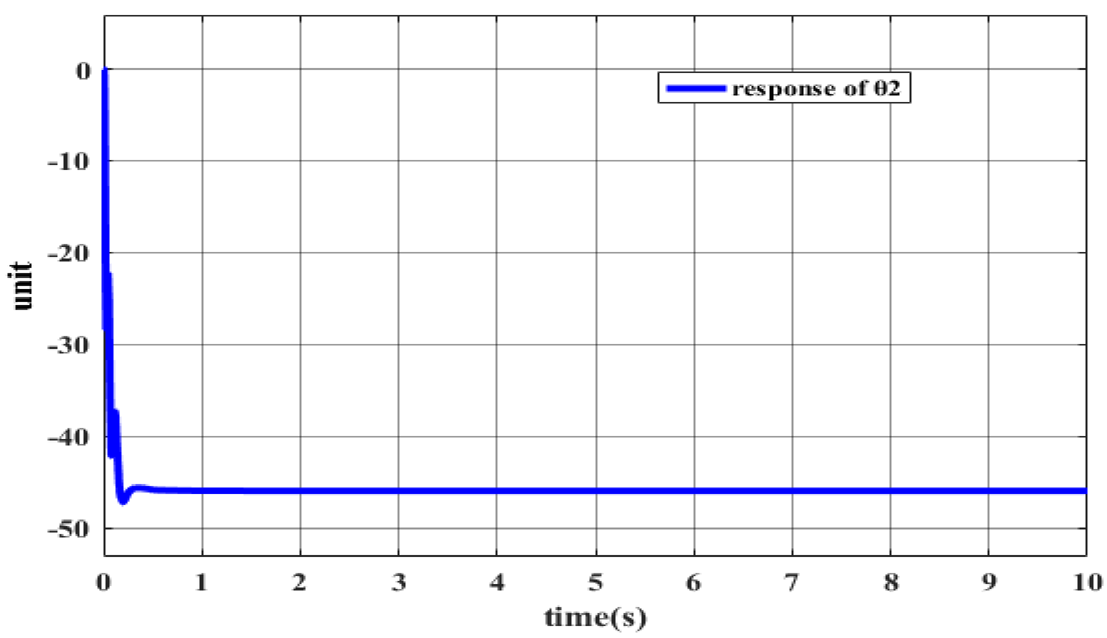


Figure 5.22 Response of $\boldsymbol{\theta_2}$ for temperature

Table 4. 1 Performance characteristics comparison of PID controller and MRAC for SISO concentration with $\boldsymbol{\gamma_1 = 150\ and\ \gamma_2 = 7000}$.

Performance characteristics of concentration	MRAS	PID
rising time	0.41 sec	0.42 sec
Over shoot	0 %	10.4%
Settling time	0.178 sec	0.48
Steady state error	0	0
Stability	Stable	Stable

Table 4.2 Performance characteristics comparison of PID controller and MRAC for SISO temperature with $\boldsymbol{\gamma_1 = 140\ and\ \gamma_2 = 90000}$

Performance characteristics of temperature	MRAS	PID
rising time	0.41	0.769 sec
Over shoot	0	2.26%
Settling time	0.177	7.86 sec
Steady state error	0	0
Stability	Stable	Stable

CHAPTER SIX

CONCLUSION

In this thesis, model reference adaptive control system is designed for continuous stirred tanker reactor (CSTR) and the design procedure consists of 4 steps: first the mathematical representation of the system (model) is linearized on selected operating point. Second, the interaction is analyzed and designed a decoupler to decouple the continuous stirred tanker reactor to two SISO system. Third, PID controller is designed for each SISO systems to compare with MRAC. Finally, MRAC with MIT rule adaptation mechanism is developed.

From simulation results show that, the characteristics plant is follow to the model characteristics. Therefore the model behavior is the same as plant characteristics or it can be substituted for the plant characteristics and we observed the effect of adaptation gain is on the time response characteristic of the second order system when the adaptation gain increases the convergence rate increase, mean we have fast response (small rise time), short settling time with small(almost zero) over shoot and the table 4.1 and 4.2 demonstrated that while the adaptive controller exhibits superior performance and the PID controller has the convergence time of typically large and there is large overshoot.

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