Effect of Model Plant Mismatch in Model Predictive Controller Performance: Continuous Stirred Tank Reactor

by

Nader Ashar Khan Bin Azman

15013

Dissertation submitted in partial fulfilment of

the requirements for the

Bachelor of Engineering (Hons)

(Chemical Engineering)

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Universiti Teknologi PETRONAS,
32610 Bandar Seri Iskandar,
Perak Darul Ridzuan

CERTIFICATION OF APPROVAL

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Approved by,

____

(Dr Lemma Dendena Tufa)

UNIVERSITI TEKNOLOGI PETRONAS

BANDAR SERI ISKANDAR, PERAK

May 2015

CERTIFICATION OF ORIGINALITY

This is to certify that I am responsible for the work submitted in this project, that the original work is my own except as specified in the references and acknowledgements, and that the original work contained herein have not been undertaken or done by unspecified sources or persons.

NADER ASHAR KHAN BIN AZMAN

ABSTRACT

Plant model is one of the important aspects in the design and implementation of Model Predictive Controller (MPC). The performance of MPC depends on the accuracy and quality of plant model. However, dynamic behaviour of a plant may change with time. Hence, plant model that are used for the design will no longer represent the plant current state after some time. In this dissertation, the effect of model plant mismatch on MPC performance will be shown by the researcher. During the conduct of this research, the researcher has developed a non-linear CSTR model by using SIMULINK. Manipulated variable and controlled variable for the CSTR model has been set by the researcher. Besides that, the researcher developed 3 different linear transfer function model using 3 different ranges. By using this 3 different transfer function model, the researcher designed 3 different MPC. The researcher has tested the plant model with 2 different tests. First, to understand the dynamic model of this CSTR, the researcher has done an open loop test to this CSTR model by adding few percentages of increment in step change to the plant input. The changes in controlled variable inside the reactor is then measured and analyzed. For the second test, the researcher done a closed loop test to measure the performance of MPC between the accurate plant models and mismatch plant models. This test is done by using MPC with plant model to control to a limit which is out of its range to represent the mismatch plant model. In the open loop test, when step change is added to the plant input, all output changes from its set point which clearly shows the non-linearity behaviour of the plant. For the MPC performance test, when mismatch is added, the controller becomes less stable and it took a longer time to reach the steady state and the new set point.

ACKNOWLEDGEMENT

First and foremost, praise to the Almighty for his grace and mercy for giving me strength in completing my Final Year Project.

A huge appreciation I bid to my supervisor Dr Lemma Dendena Tufa, who gives his best to guide and assist me in any problems I encountered during the conduction of this project. His invaluable help of constructive comments and suggestions throughout the semesters and project works have contributed to the success of my Final Year Project.

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CHAPTER 1
INTRODUCTION

1.1 BACKGROUND STUDY

Model Predictive Control (MPC) is an advance control that has been used widely because of its ability to respond to provide attractive solutions of tracking problem for linear and non-linear system [1, 2]. MPC can handle different variables and inconsistent flow of input and output of reactors [3]. MPC works by predicting the future output of a system by analyzing current state and process model of a reactor.

Performance of a model controller depends on the quality of the model [4-7]. Although model plant mismatch is impossible to avoid, it is highly beneficial to reduce the impact [7]. Model constancy plays an important role in MPC [5]. The performance of MPC relies on the model quality and the mismatch of the model-plant. However, poor model does not necessarily lead to deterioration of the performance. Furthermore, disturbance and poor tuning may also lead to degradation of performance. Therefore, it is highly advantageous to be able to segregate the function of MPM in deprived control and measure its impact. This research seeks to tackle these issues. Two rules that assist in the analysis of poor performance are also proposed. The simulation done is based on close loop data analysis from the process. This research also shows that Model Plant Mismatch (MPM) impact on control quality depends on the direction of set point. The proposed methodology is then tested to diagnose controller performance in model mismatch plant via examples of MATLAB simulation.

1.2 PROBLEM STATEMENT

Plant models play a central role in the design and implementation of model based controllers like MPC. The accuracy of the model used for design of the controller directly affects the performance of the controller. The dynamic behaviour of processes may change with time. Therefore, the model used for the design will no more represent the plant after some time. The deviation of the plant model from the plant is called model plant mismatch. While it is obvious that performance of a plant will depreciate, how much the performance of the controller will deteriorate for a given model is not properly studied. In this research, the researcher will study the effect of model plant mismatch on MPC performance. Throughout this project, the suitability of linear MPC to control non-linear CSTR will be evaluated. Therefore, this project is useful to address the problems faced by many chemical industries related to process control.

1.3 OBJECTIVE

The objective of this research is:

- To develop a non-linear model of Continuous Stirred Tank Reactor System (CSTR) using MATLAB-SIMULINK to represent the plant.
- To design and implement a model predictive controller (MPC) using a linear transfer function model to the plant.
- To introduce mismatch on the plant and investigate how the performance deteriorates.

CHAPTER 2
LITERATURE REVIEW

To determine dynamic behaviour of a model, we must understand the rate of reaction for that particular model. The hypothetic reaction below is from a paper by Dr M. J. Willis by [1]:

$$A \rightarrow B \quad (1)$$

There are 2 ways to express rate of reaction for this reaction. First way is by determining rate of disappearance of feed, which in this reaction is A.

Rate of reaction: rate of disappearance of A = $-r_A$(mol/$dm^3.s$)

Second way to calculate rate of reaction for this reaction is by determining rate of formation of product, which in this case is B.

Rate of reaction: rate of formation of B = r_B(mol/$dm^3.s$)

In the Figure 1 reaction, if rate of formation of B, is 0.4 mol per cubic centimetre per second, than the rate of disappearance of A will also be 0.4 mol per cubic centimetre per second.

i.e.: if $-r_A$= 0.4 mol/$dm^3.s$

r_B= 0.4mol/$dm^3.s$

General mole balance is defined as:

$$Rate\ of\ accumulation = Rate\ of\ mass\ in - Rate\ of\ mass\ out - Rate\ of\ generation$$

$$Rate\ of\ accumulation = V\frac{dCA}{dt} \quad (2)$$

Since Density = Mass/Volume

$$Flow\ in - Flow\ out = \frac{m}{p}(C_{Ao} - C_A) \tag{3}$$

$$Rate\ of\ Generation = V(-r_A) = VC_A k_o e^{-E/RT} \tag{4}$$

Combining all the equations, we will get

$$\frac{dCA}{dt} = V\frac{m}{p}(C_{Ao} - C_A) - VC_A k_o e^{-\frac{E}{RT}} \tag{5}$$

Model predictive control is a control plan that proposes solutions for tracking and regulation difficulty of linear or nonlinear systems [2]. Model Predictive Control (MPC) algorithms offer the way of tackling the control problem by predicting the value of output based on current state of a model and its process. Recently, MPC has become an important research topic for both academicians and engineers. To explain the basic concept of MPC, author would like to refer to Figure 1 which author take from paper titled Effect of Model-plant Mismatch on MPC Controller Performance by A.I. Nafsun and N. Yusoff [3].

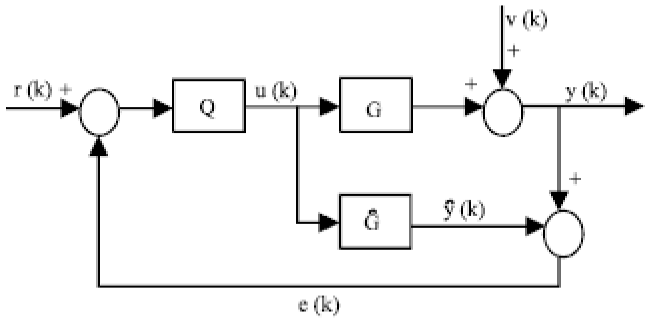

FIGURE 2.1 Model Plant Mismatch

The model plant mismatch in Figure 2.1, is described as Ĝ:

$$\Delta = G - \hat{G} \tag{6}$$

Plant output [y(k)] is defined as

$$y(k) = Gu(k) + v(k) \tag{7}$$

Where u(k) & v(k) : vector of manipulated variables.

Model output of this plant is $\hat{y}(k) = \hat{G}u(k)$ (8)

Model residuals of this plant is describes as

$$e(k) = y(k) - \hat{y}(k) = \Delta u(k) + v(k) \quad (9)$$

Few researches have been done to determine effect of model plant mismatch on controller performance. Four types of mismatch can be investigated to determine controller performance in a mismatch model. The experiment is conducted by tuning the MPC controller according to the plant [3]. Mismatch is then introduced in the plant controller. The 4 mismatch introduced are gain mismatch, reverse gain mismatch, time delay mismatch and time constant mismatch. Gain mismatch has bigger impact on controller performance than time delay and time constant. However, for set point tracking, MPC controller performs better in time delay and time constant than gain mismatch.

One way to minimize the impact of model plant mismatch is to detect the mismatch promptly [4]. Changes in reactor parameter will usually lead to mismatch model plant. Once the mismatch is detected, model will be adjusted to avoid deterioration in model control performance. This research presents detection of model plant mismatch in a run of mine ore process using correlation analysis method. By using this method, location of the mismatch can be detected accurately and the process control is restructured subsequently.

A research has been done to introduce Plant Model Ratio (PMR) which will detects model plant mismatch and ease the identification of the mismatch source whether the mismatch is from gain, time constant or time delay [6]. The objective of this research is to improve the PMR approach in a few key aspects which are experimental effort, estimation and assessment procedure by conducting theoretical studies of its approach. An assessment procedure is devised based on theoretical properties of PMR. As a result, three hypotheses tests are proposed for testing of PMR. Designed set point with minimal excitation, based on feature of Plant Model Ratio, is pivotal in the diagnosis of MPM. To check the effectiveness of the proposed methodology, the results obtained are compared with the result of the existing method.

Chemical reactors are important for chemical engineers. It is important to understand dynamic characteristic of a particular reactor to ensure that it will operate smoothly [1]. Full understanding of reactor will contribute to effective design of control system. The paper written by Dr. M. J. Willis objective is to introduce concept of CSTR dynamic model and know how to develop simulation model for CSTR. To depict the dynamic behaviour of a CSTR model, there are few parameters needed to be developed which are mass balance, component balance and energy balance equations. To develop this equation, full understanding of functional expression which is used to describe chemical reaction is necessary.

Since Continuous Stirred Tank Reactor (CSTR) can be difficult to control, there are a lot of researches done to determine the best way to control CSTR reactor. Continuous Stirred Tank Reactor (CSTR) is widely used in engineering industry. However, it is highly non-linear and is difficult to control [8-11]. There are several algorithms that can be used to construct a control design which are Lyapunov function [8, 9], Artificial Bee Colony (ABC) [10] function, Fuzzy Logic [11, 12] and etc.

Control law design and stability analysis has been widely studied in literature. Stability analysis is conducted based on mathematical methods such as linearization method or Lyapunov function. A research has been conducted to develop Lyapunov function which can be used for analyzing stability from the existing thermodynamic function [8]. In this research, numerical simulation is illustrated in the theory of closed loop control and open loop stability analysis in a single phase CSTR, which in this research is non-isothermal liquid phase. 2 chemical reaction are used to measure effectiveness of this Lyapunov function which are production of cyclopentenol from cyclopentadiene by acid-catalyzed electrophilic addition of water and acid-catalyzed hydration of 2-3-epoxy-1-propanol to glycerol. Result of this simulation is then compared to result by using classical control strategy.

There are researches that have been conducted to propose a control design by using Artificial Bee Colony (ABC) algorithm. One of the researches is conducted to propose a control design for a highly unstable and non-linear CSTR [10]. Proportional Integral Derivatives (PID) control is used in this research. The simulation is started by tuning 3 PID control gain using ABC algorithm. Using ABC algorithm, controller gains can be acquire by minimizing the cost function given. Several control operation are provided to test the effectiveness of this control method.

One of the researches is done to optimize control of CSTR reactor which is used to purify zinc [13]. In zinc purification process, zinc dust is used to eliminate impurity ions. The research is conducted by using Interacting Continuous Stirred Tank Model as a model for the simulation. Several unknown parameters required for this research is collected from numerous factory in China which is related to zinc purification. A time delayed optimal control problem is created for the zinc solution purification. In this research, optimum control law is designed to reduce impurities in zinc solution. The optimum control law is proved to be effective after undergoing several simulations.

Numerous control approaches have been practiced to CSTR parameters. A research has been conducted to present two different control strategies which are gain scheduling performance by fuzzy logic approaches & least square approaches and Imperialist Competitive Algorithm [11]. Objective of this research is to control temperature of the CSTR. This research is conducted by simulation using MATLAB software. Performance of this controller is analyzed according to Integral Absolute Error (IAE) and Sum of the Square Error (SSE) criteria.

CHAPTER 3

METHODOLOGY

The researcher has gone through 4 main steps during the conduct of this research. A non-linear CSTR model has been developed by the researcher using SIMULINK. Next, linear transfer function models are created using 3 different ranges to be used as a plant model for MPC. Next, 3 MPC controller is designed by using 3 different transfer function model. Performance of mismatch model plant is measured against good plant model.

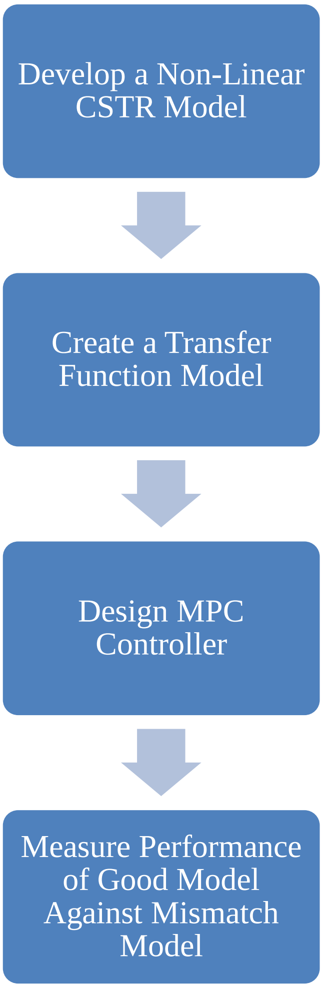

FIGURE 3.1 4 Main Steps

3.1 NON-LINEAR CSTR MODEL

This CSTR model, Figure 3.2 with the assumptions and operating conditions, Table 3.1 is taken from ‘Chemical Process Modelling and Computer Simulation’ [14]. The reaction in the CSTR is a substitution reaction which is:

$$A \rightarrow B \qquad (1)$$

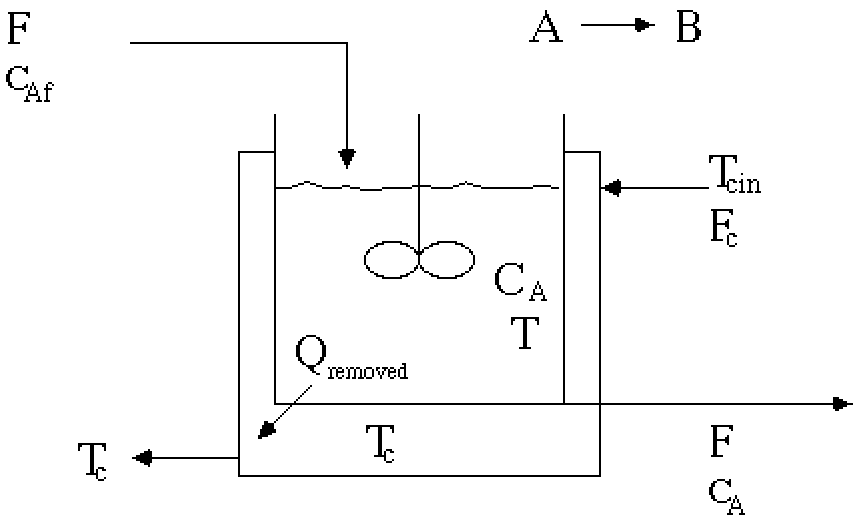

FIGURE 3.2 Schematic Representation of CSTR

There are few assumptions that are made in the development of this model [CSTR]:

- The heat loss from the process (well insulated) to the atmosphere are negligible.
- The mixture density (p) and heat capacity (C_p) are assumed constant.
- The coolant is perfectly mixed and therefore the temperature throughout the jacket is the same (T_j).
- The mass of the metal walls is negligible so the thermal inertia of the metal does not need to be considered.
- There are no spatial variations in concentration, temperature or reaction rate throughout the reactor.
- The exit stream has the same concentration and temperature as the entire reactor liquid.
- The overall heat transfer coefficient (U_i) is assumed constant.

- No energy balance around the jacket is considered. Indeed, the cooling jacket temperature (T_j) can directly be manipulated in order to control the desired reactor temperature (T).
- The reactor is flat-bottomed vertical cylinder and the jacket is around the outside and the bottom.

The operating conditions for the CSTR are as per Table 3.1.

TABLE 3.1 Steady State and Operating Conditions

	Steady sate and operating conditions	
A_c	Cross-sectional are of the reactor, m^2	4.2822
C_A	Concentration of reactor A in the exit stream, kmol/m^3	8.56303
C_{Af}	Concentration of A in the feed stream, kmol/m^3	10.0
d	Diameter of the cylindrical reactor, m	2.335
E	Activation energy, kJ/kmol	49551.112
F_i	Volumetric feed flow rate, m^3/h	10.0
h	Height of the reactor liquid, m	2.335201
$(-\Delta H)$	Heat of reaction, kJ/kmol	24936.64
R	Universal gas constant, kJ/(kmol.K)	8.314
a	Frequency factor, h^{-1}	34930800.0
pC_p	Multiplication of mixture density and heat capacity, kJ/(m^3.°C)	
T	Reactor temperature, °C	38.1771
T_f	Feed temperature, °C	25
T_j	Jacket temperature, °C	25
U_i	Overall heat transfer coefficient, kJ/(m^2.°C.h)	70
	Integration time interval = 0.005h	

The researcher develops a CSTR model based on three equations stated in 'Chemical Process Modelling and Computer Simulation'[14]. The equation is then converted into SIMULINK form. The first equation is:

$$\frac{dh}{dt} = \frac{F_i}{A_c} - \sqrt{10h/A_c} \tag{10}$$

Figure 3.3 shows the SIMULINK version for this equation.

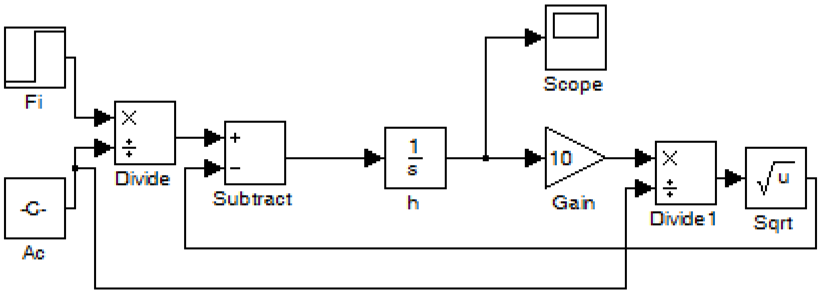

FIGURE 3.3 SIMULINK for Height Equation

Second equation used for SIMULINK model is:

$$\frac{dC_A}{dt} = \frac{F_i}{A_c h}\left(C_{Af} - C_A\right) - a \exp\left(\frac{-E}{RT}\right) C_A \tag{11}$$

Figure 3.4 shows the SIMULINK version for this equation:

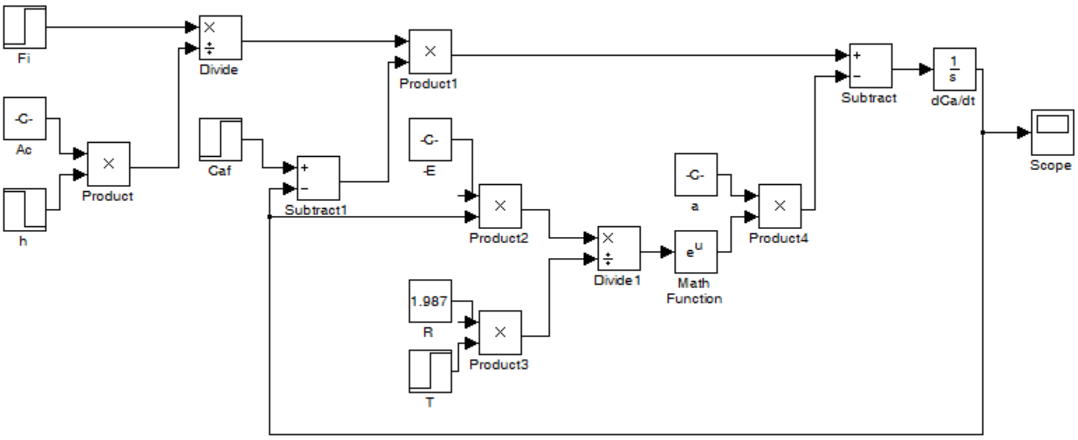

FIGURE 3.4 SIMULINK for Concentration of A Equation

Third equation used to develop CSTR in SIMULINK is:

$$\frac{dT}{dt} = \frac{F_i}{A_c h}\left(T_f - T\right) + \left(\frac{-\Delta H}{\rho C_p}\right) a \; exp\left(\frac{-E}{RT}\right) C_A - \frac{U_i A_h}{\rho C_p A_c h}\left(T - T_j\right) \qquad (12)$$

Where $\qquad A_h = A_c + \pi d h$

Figure 3.5 shows SIMULINK version of this temperature equation.

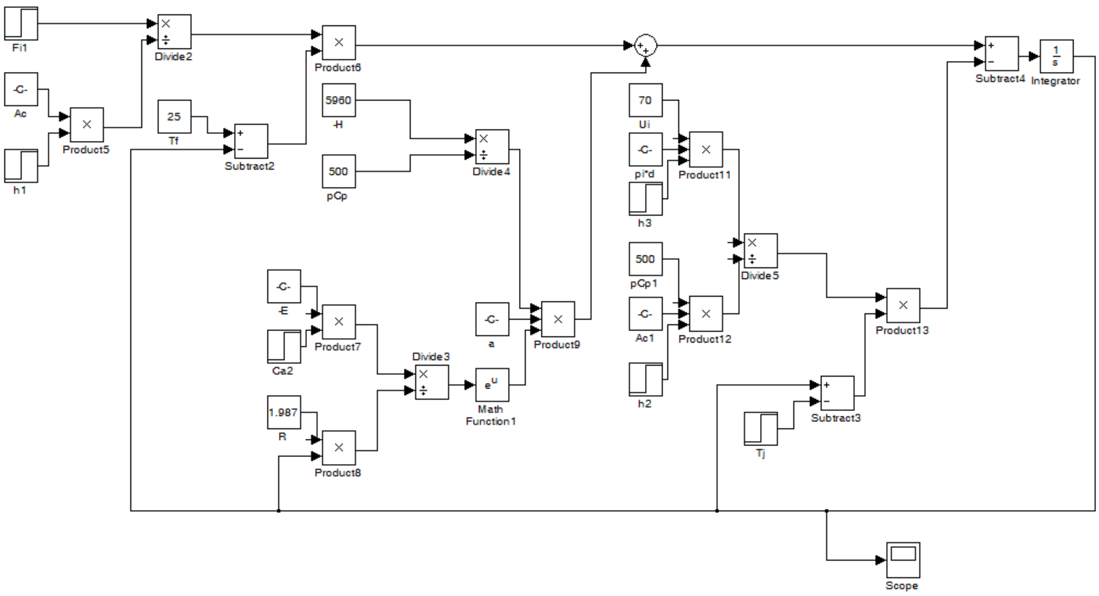

FIGURE 3.5 SIMULINK for Temperature Equation

After the researcher develops the three equations individually, the researcher combines all three equations in a single SIMULINK file. The result is shown in Figure 3.6.

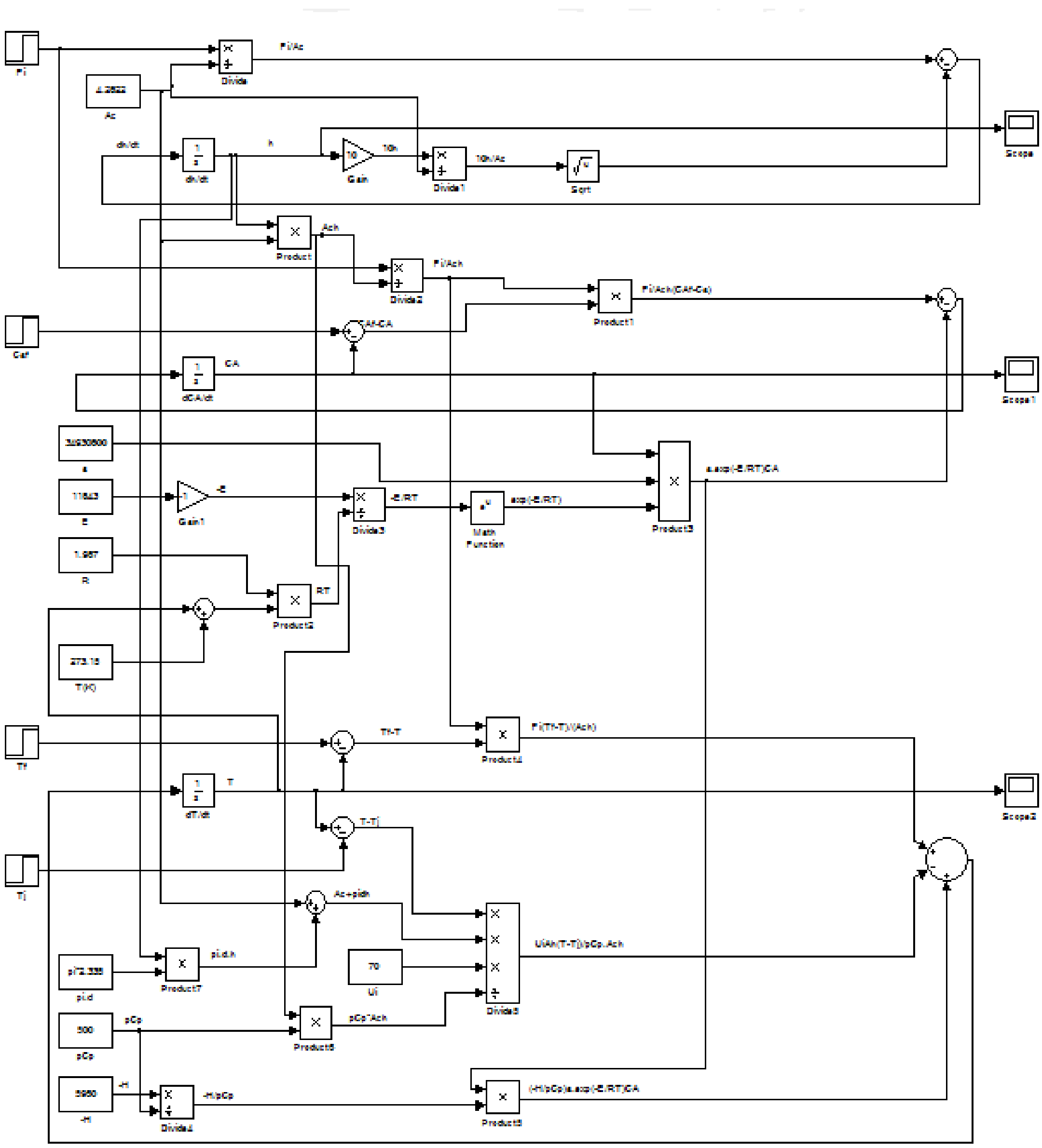

FIGURE 3.6 CSTR Model in SIMULINK Form

This CSTR model is then merged into a single subsystem.

3.2 TRANSFER FUNCTION MODEL FOR MPC

To create a transfer function model for MPC, we first must understand the dynamic behaviour of this CSTR model. CSTR controller matrix for this CSTR model is:

$$\begin{bmatrix} cA \\ h \end{bmatrix} = \begin{bmatrix} G_{11} & G_{12} \\ G_{21} & G_{22} \end{bmatrix} \begin{bmatrix} F_i \\ T_j \end{bmatrix} \tag{13}$$

where C_A and h is the output or controlled variable.

F_i and T_j are input or manipulated variable.

G11, G12, G21, G22 are transfer function model that will be created by the researcher.

F_i in the CSTR SIMULINK model was increased from 0 to 1. The value for G_{11} and G_{21} is attained by inputting source code below into MATLAB.

```
Data = iddata("output", "input", "sample time")

data11 = iddata(cA,Fi,0.1)

data21=iddata(h,Fi,0.1)
```

Next, value of F_i was returned to original state whereas value of T_j was increased from 0 to 5. The value for G_{12} and G_{22} was attained by inputting the source code below into MATLAB.

```
Data12 = iddata(cA,Tj,0.1)

Data22=iddata(h,Tj,0.1)
```

Data11, data12, data21 and data 22 are then converted into G_{11}, G_{12}, G_{21} and G_{22} by estimating the transfer model in systemic identification tools. All the transfer function model is then converted into the transfer function model general form which is $G = \frac{Kpe^{-td}}{\tau ps+1}$ by inputting the below source code into MATLAB.

```
G11 = tf([G11.Kp],[G11.Tp1 1],'iodelay',G11.Td)

G12 = tf([G12.Kp],[G12.Tp1],'iodelay',G12.Td)

G21 = tf([G21.Kp],[G21.Tp1 1],'iodelay',G21.Td)

G22 = tf([G22.Kp],[G22.Tp1 1],'iodelay',G22.Td)
```

The transfer function model is then converted into general equation of transfer model matrix by inputting the below source code into MATLAB.

```
G = [G11 G12 ; G21 G22]

P = [G21 G22 ; G31 G32]

M = [M11 M12 ; M31 M32]
```

$$[G] = \begin{bmatrix} G_{11} & G_{12} \\ G_{21} & G_{22} \end{bmatrix}$$

FIGURE 3.7 General Equation for Transfer Model Matrix

Transfer function model obtained by this limit of F_i and T_j is shown below which is denoted as TF1

$$[G] = \begin{vmatrix} \dfrac{-0.5608e^{-2.55s}}{5.203s+1} & \dfrac{-0.08638e^{-0.788s}}{3.608s+1} \\ \dfrac{0.4904}{2.156s+1} & \dfrac{9.447e^{-06}}{2s+1} \end{vmatrix}$$

FIGURE 3.8 Transfer Function 1 (TF1)

The process is repeated by changing F_i and T_j into different value to get two more Transfer function model. For $-1.5<F_i<0$ and $-2.5<T_j<0$, the transfer function obtained is shown below which is also denoted as TF2.

$$[P] = \begin{bmatrix} \dfrac{-2892e^{-2.44s}}{2.236s+1} & \dfrac{-0.05432e^{-0.755s}}{2.29s+1} \\ \dfrac{0.432e^{-0.0178s}}{1.768s+1} & \dfrac{-1.889e^{-05}}{2s+1} \end{bmatrix}$$

FIGURE 3.9 Transfer Function 2 (TF2)

For $3<F_i<0$ and $5<T_j<0$, the transfer function model obtained is shown below which is also denoted as TF3.

$$[M] = \begin{bmatrix} \dfrac{-0.2417e^{-2.34}}{1.755s+1} & \dfrac{-0.04952e^{-0.695s}}{2.164s+1} \\ \dfrac{0.397e^{-0.0347s}}{1.542s+1} & \dfrac{-9.44e^{-06}}{2s+1} \end{bmatrix}$$

FIGURE 3.10 Transfer Function 3 (TF3)

The range of how much can a transfer function model changes controlled variables into new set point can be acquired after the 3 transfer function model was developed.

Below is the tabulated version of transfer function model magnitude influence on controlled variable.

TABLE 3.2 Limit of MPC Performance

Transfer function	F_i	T_j	C_A	h
1	$0 < F_i < 1$	$0 < T_j < 5$	$-0.56 < C_A < -0.43$	$0 < h < 0.49$
2	$-1.5 < F_i < 0$	$-2.5 < T_j < 0$	$0.14 < C_A < 0.43$	$-0.65 < h < 0$
3	$-3 < F_i < 0$	$-5 < T_j < 0$	$0.25 < C_A < 0.72$	$-1.2 < h < 0$

In this research, the researcher will show the effect of model plant mismatch in MPC performance. Mismatch model means controlling output set point using model that is out of the set point range. As an example, using the second transfer function model no 2 to increase C_A up until 0.70.

3.3 DESIGN OF MPC

MPC control is attached to the CSTR model developed in SIMULINK. MPC controller is designed by making C_A as the controlled variable and F_i and T_j as manipulated variable. To design MPC controller, set point for the controlled variable must be objectified. Weight constraint is set for all input and output. The value of minimum and maximum weight constraint set is shown as per below.

TABLE 3.3 Value of Weight Constraint Set

Constraint	Minimum	Maximum
F_i	-3	1
T_j	-5	5
C_A	-0.5	0.5
h	-1	0.5

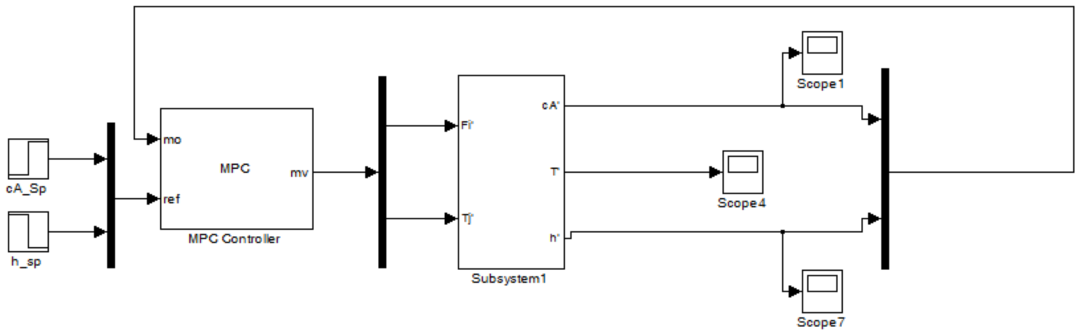

FIGURE 3.11 MPC Controller Design

The three transfer function model is use to create three different MPC model namely MPC1, MPC2 and MPC3. TF1 is use to create MPC1, TF2 for MPC2 and TF3 for MPC3.

3.4 PERFORMANCE MEASUREMENT

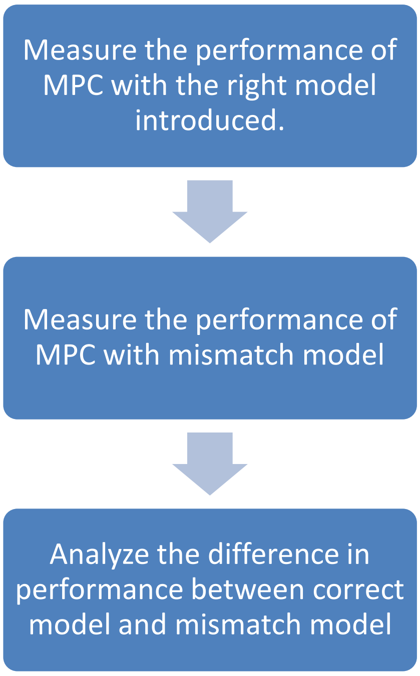

FIGURE 3.12 Performance Measurement Method

3.5 GANTT CHART

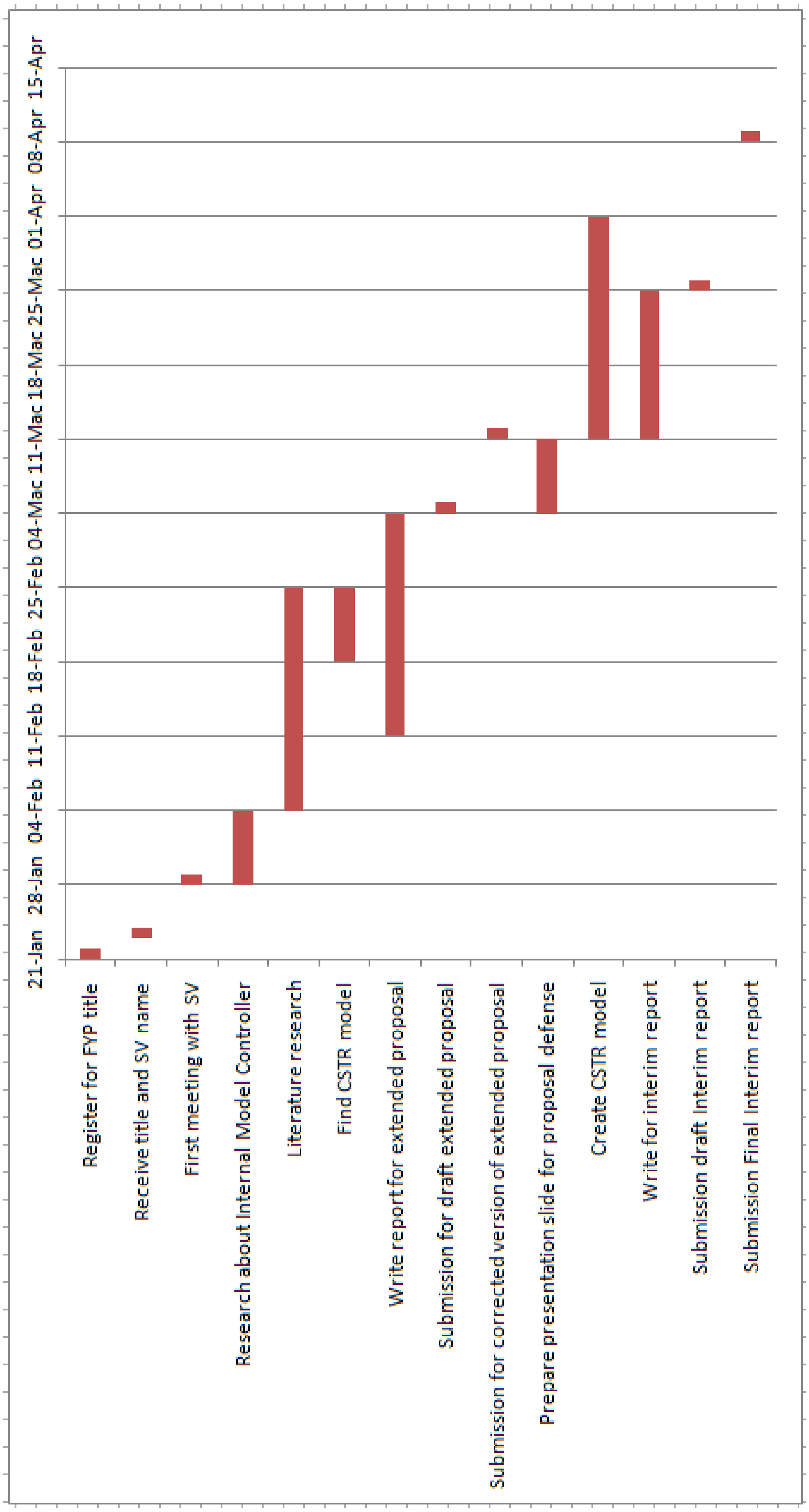

FIGURE 3.13 Gantt Chart

3.6 KEY MILESTONE

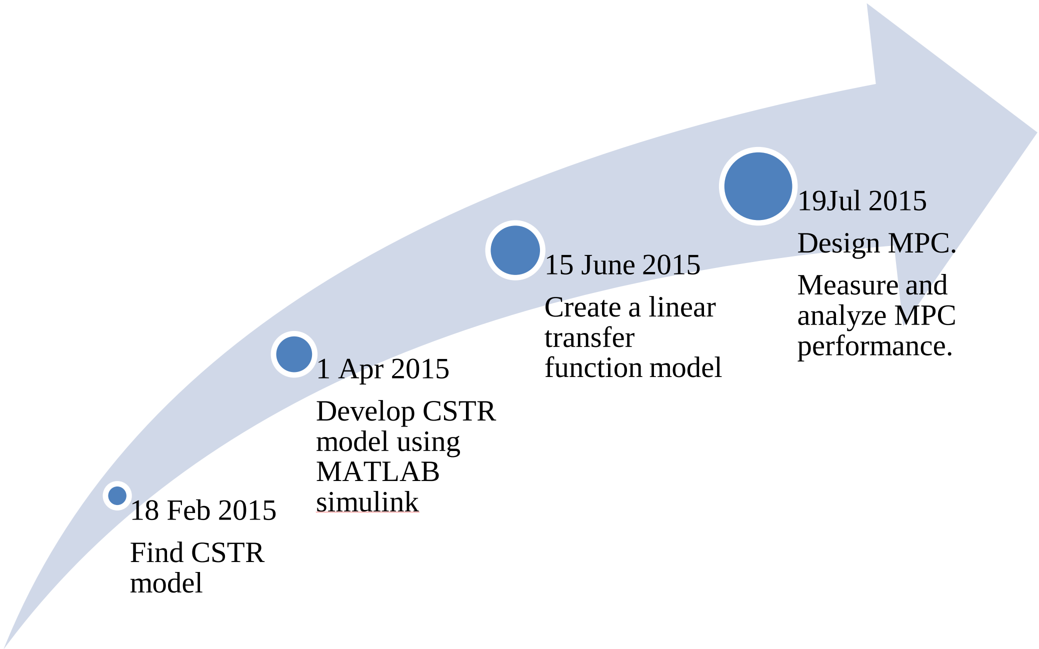

FIGURE 3.14 Key Milestone

CHAPTER 4
RESULTS AND DISCUSSION

4.1 OPEN LOOP TEST ON CSTR MODEL

To understand the dynamic model of this CSTR, the researcher has tested this model by adding 3% increment step change to the initial feed flow rate. The simulation is then started and change in value of C_A, h is observed and analysed. The test is then repeated by changing the increment value of feed flow rate to 10%.

4.1.1 Feed Flow Rate 3% Increment

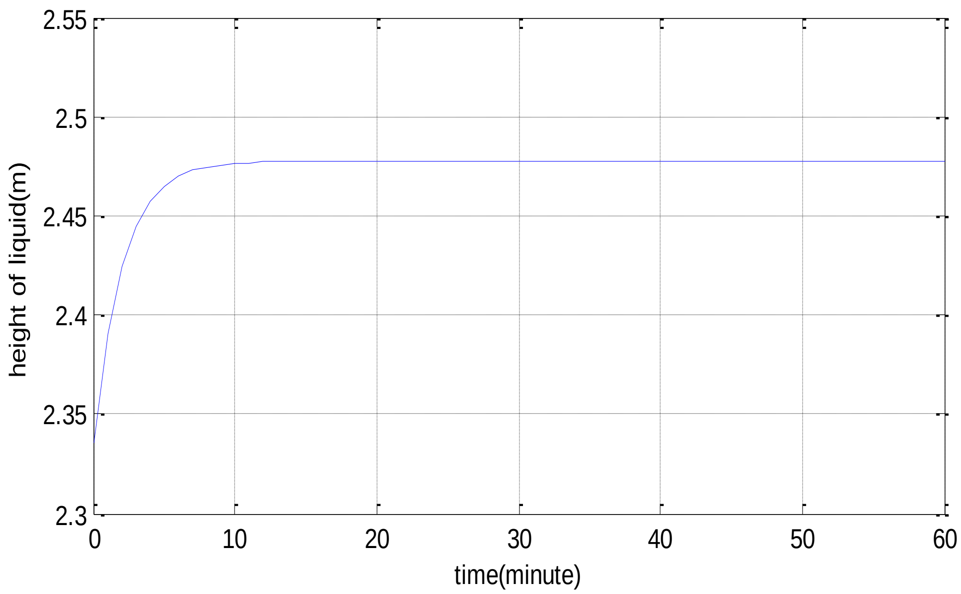

FIGURE 4.1 Changes in h When 3% Increment is Added to Volumetric Feed Flow Rate

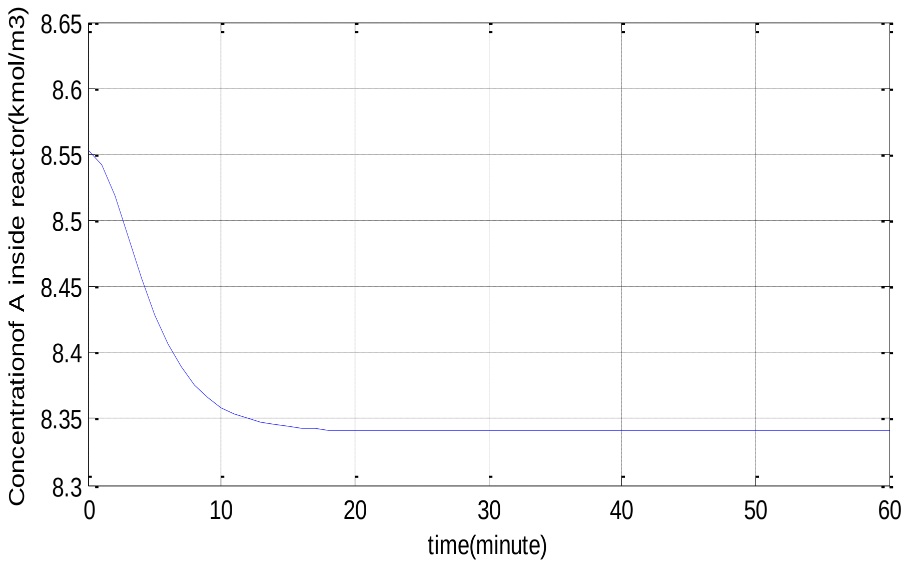

FIGURE 4.2 Changes in C_A When 3% Increment is Added To Volumetric Feed Flow Rate

When 3% increment is added to initial volumetric feed flow rate value, liquid level inside the reactor increase from the set point but concentration of reactant A in the reactor decreased from its set point. All of these parameters increase/decrease before it reaches steady state value. These changes are still bearable since none of these changes more than 10% from the initial set point.

4.1.2 Feed Flow Rate 10% Increment

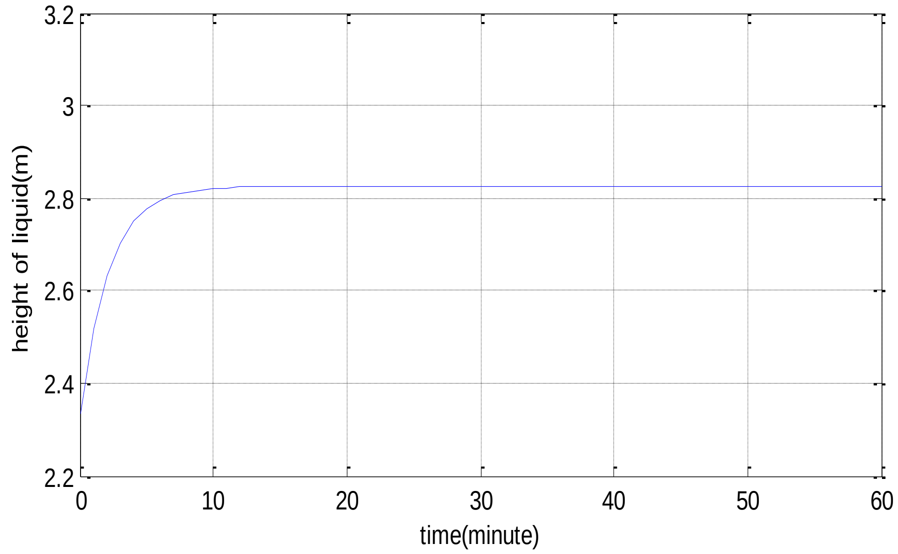

FIGURE 4.3 Changes in h When 10% Increment is Added to Volumetric Feed Flow Rate

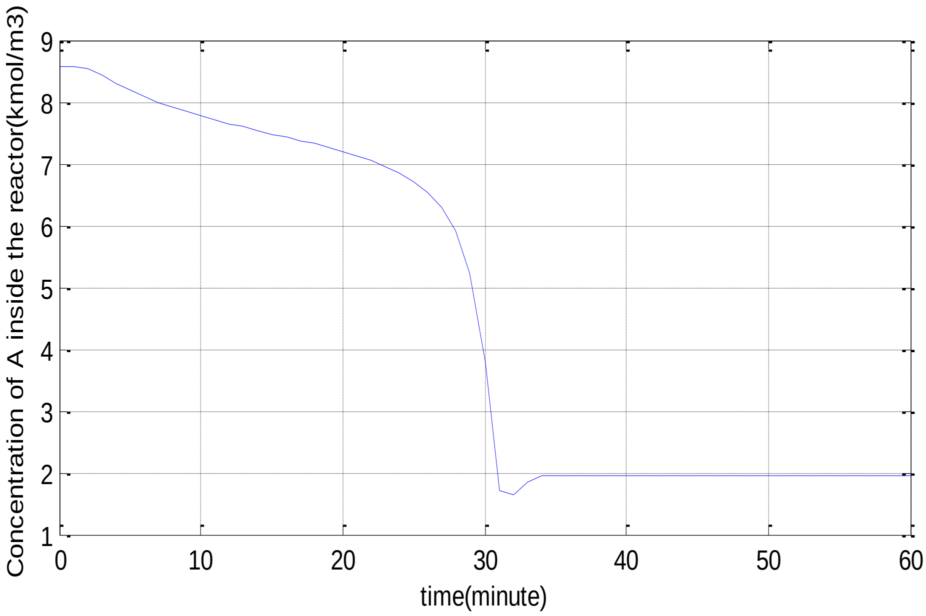

FIGURE 4.4 Changes in C_A When 10% Increment is Added To Volumetric Feed Flow Rate

When 10% increment is added to initial volumetric feed flow rate value, liquid level inside the reactor increase from the set point but concentration of reactant A in the reactor decreased from its set point. The pattern is similar to when 3% increment is added. However, liquid level increases from 2.3 to 2.8 and temperature increases from 38 to 43 before both of these parameters reach steady state. Both of these change is more than 10% of its initial set point value.

These results clearly show nonlinear characteristics of this CSTR model. Therefore, the researcher deduces that linearization system of the nonlinear system behaves quite differently and it cannot be used to represent nonlinear system of this CSTR.

4.2 MPC PERFORMANCE IN MODEL PLANT MISMATCH

4.2.1 First Transfer Function Model (TF1)

Using the first transfer function model,TF1, C_A is set to -0.5kmol. The MPC performance is measured against mismatch model, which in this case is TF2.

$$[G] = \begin{vmatrix} \dfrac{-0.5608e^{-2.55s}}{5.203s+1} & \dfrac{-0.08638e^{-0.788s}}{3.608s+1} \\ \dfrac{0.4904}{2.156s+1} & \dfrac{9.447e^{-06}}{2s+1} \end{vmatrix}$$

FIGURE 4.5 Accurate Model (TF1)

$$[P] = \begin{bmatrix} \dfrac{-2892e^{-2.44s}}{2.236s+1} & \dfrac{-0.05432e^{-0.755s}}{2.29s+1} \\ \dfrac{0.432e^{-0.0178s}}{1.768s+1} & \dfrac{-1.889e^{-05}}{2s+1} \end{bmatrix}$$

FIGURE 4.6 Mismatch Model (TF2)

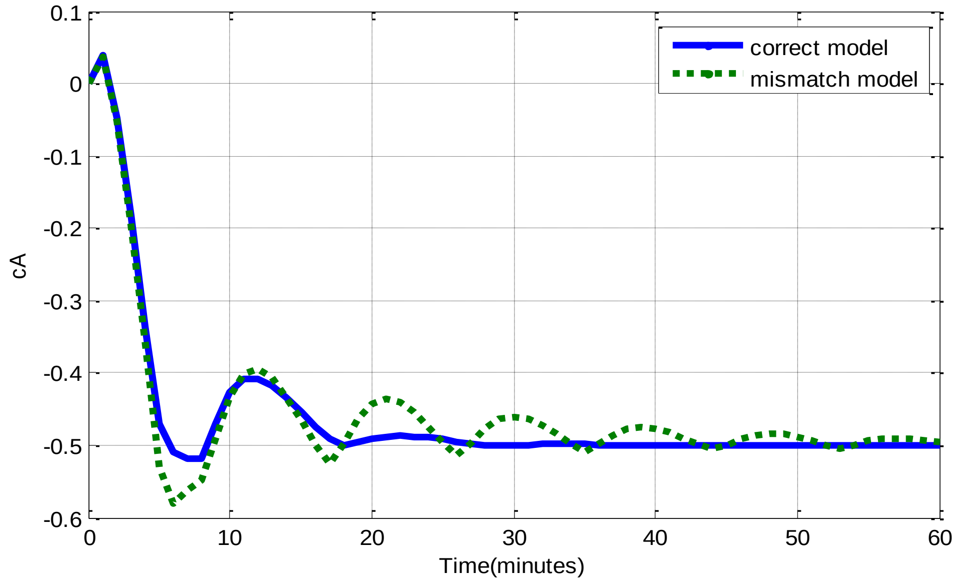

FIGURE 4.7 Graph of MPC Performance in Accurate Model and Mismatch Model

When a correct model is used, the concentration reaches its new set point and become steady state in 30 minutes. However, when a mismatch model is used, MPC performance deteriorates it became less stable as the concentration of A oscillates heavily. The concentration also did not reach and become steady state even after 60 minutes have passed.

The test is further preceded by changing TF2 with TF3, which is model mismatch for this range.

$$[G] = \begin{vmatrix} \dfrac{-0.5608e^{-2.55s}}{5.203s+1} & \dfrac{-0.08638e^{-0.788s}}{3.608s+1} \\ \dfrac{0.4904}{2.156s+1} & \dfrac{9.447e^{-06}}{2s+1} \end{vmatrix}$$

FIGURE 4.8 Accurate Model (TF1)

$$[M] = \begin{bmatrix} \frac{-0.2417e^{-2.34}}{1.755s+1} & \frac{-0.04952e^{-0.695s}}{2.164s+1} \\ \frac{0.397e^{-0.0347s}}{1.542s+1} & \frac{-9.44e^{-06}}{2s+1} \end{bmatrix}$$

FIGURE 4.9 Mismatch Model (TF3)

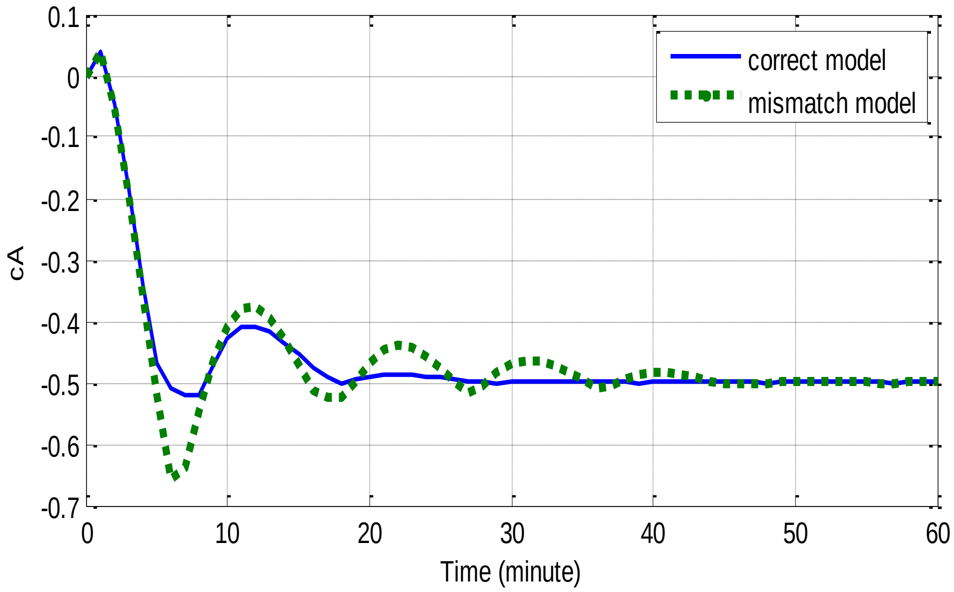

FIGURE 4.10 Graph of MPC Performance in Accurate Model and Mismatch Model

The similar behaviour is observed from Figure 4.10 during the conduct of this test. When a correct model is used, MPC control concentration of A into the new set point in 30 minutes. However, when a mismatch model is used, the concentration of A oscillates heavily and MPC needs 60 minutes to stabilize concentration of A and bring it to the new set point.

4.2.2 Second Transfer Function Model (TF2)

Using the second transfer function model,TF2, C_A is set to 0.4kmol. The MPC performance is measure against the first mismatch model, which in this case is TF1.

$$[P] = \begin{bmatrix} \dfrac{-2892e^{-2.44s}}{2.236s+1} & \dfrac{-0.05432e^{-0.755s}}{2.29s+1} \\ \dfrac{0.432e^{-0.0178s}}{1.768s+1} & \dfrac{-1.889e^{-05}}{2s+1} \end{bmatrix}$$

FIGURE 4.11: Accurate Model (TF2)

$$[G] = \begin{vmatrix} \dfrac{-0.5608e^{-2.55s}}{5.203s+1} & \dfrac{-0.08638e^{-0.788s}}{3.608s+1} \\ \dfrac{0.4904}{2.156s+1} & \dfrac{9.447e^{-06}}{2s+1} \end{vmatrix}$$

FIGURE 4.12 Mismatch Model (TF1)

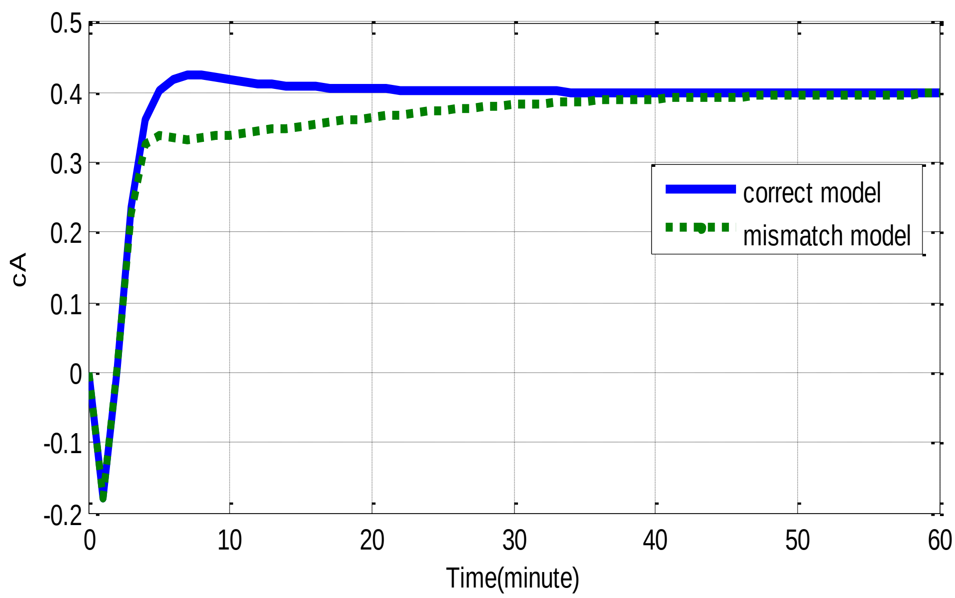

FIGURE 4.13 Graph of MPC Performance in Accurate Model and Mismatch Model

It is observed from Figure 4.13 that when correct model is used the response reaches the steady state value within 10 minutes. However, when mismatched model is used it took more than 40 minutes to reach the steady state. It is obvious from the figure that the mismatch causes the performance of the controller to deteriorate.

The test is further preceded by changing TF1 with TF3, which is model mismatch for this range.

$$[P] = \begin{bmatrix} \dfrac{-2892e^{-2.44s}}{2.236s+1} & \dfrac{-0.05432e^{-0.755s}}{2.29s+1} \\ \dfrac{0.432e^{-0.0178s}}{1.768s+1} & \dfrac{-1.889e^{-05}}{2s+1} \end{bmatrix}$$

FIGURE 4.14 Accurate Model (TF2)

$$[M] = \begin{bmatrix} \dfrac{-0.2417e^{-2.34}}{1.755s+1} & \dfrac{-0.04952e^{-0.695s}}{2.164s+1} \\ \dfrac{0.397e^{-0.0347s}}{1.542s+1} & \dfrac{-9.44e^{-06}}{2s+1} \end{bmatrix}$$

FIGURE 4.15 Mismatch Model (TF3)

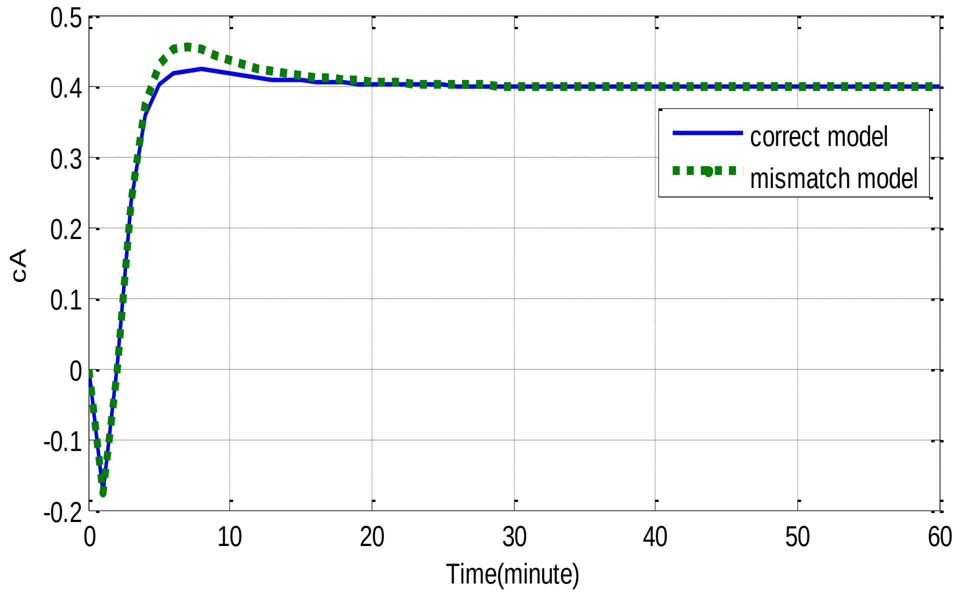

FIGURE 4.16 Graph of MPC Performance in Accurate Model and Mismatch Model

It is observed from Figure 4.16 that when correct model is used the response reaches the steady state value within 30 minutes. However, when mismatched model is used, the concentration overshoot and it took exactly 30 minutes for the output to reach the steady state. It is obvious from the figure that the mismatch causes the performance of the controller to deteriorate.

4.2.3 Third Transfer Function Model (TF3)

Using the second transfer function model, TF3, C_A is set to 0.7kmol. The MPC performance is measure against the first mismatch model, which in this case is TF1.

$$[M] = \begin{bmatrix} \frac{-0.2417e^{-2.34}}{1.755s+1} & \frac{-0.04952e^{-0.695s}}{2.164s+1} \\ \frac{0.397e^{-0.0347s}}{1.542s+1} & \frac{-9.44e^{-06}}{2s+1} \end{bmatrix}$$

FIGURE 4.17 Accurate Model (TF3)

$$[G] = \begin{vmatrix} \frac{-0.5608e^{-2.55s}}{5.203s+1} & \frac{-0.08638e^{-0.788s}}{3.608s+1} \\ \frac{0.4904}{2.156s+1} & \frac{9.447e^{-06}}{2s+1} \end{vmatrix}$$

FIGURE 4.18 Mismatch Model (TF1)

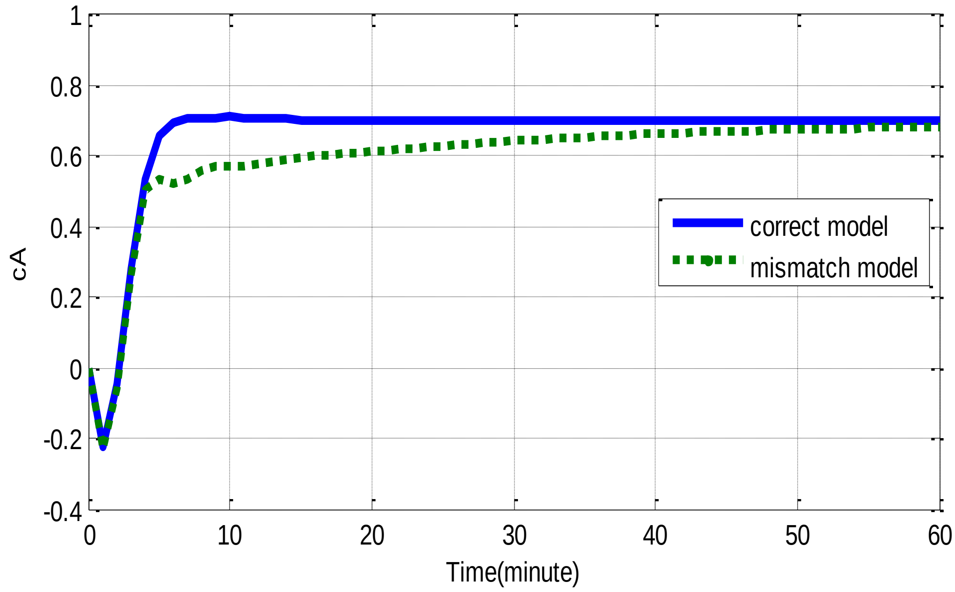

FIGURE 4.19 Graph of MPC Performance in Accurate Model and Mismatch Model

From Figure 4.19, it is observed that it took less than 10 minutes for the output to reaches steady state and new set point when a correct model is used. However, when a mismatch model is used, even after 60 minutes, the output does not reach the new set point. This shows that the mismatch cause the MPC performance to deteriorates.

The test is further preceded by changing TF1 with TF2, which is model mismatch for this range.

$$[M] = \begin{bmatrix} \dfrac{-0.2417e^{-2.34}}{1.755s+1} & \dfrac{-0.04952e^{-0.695s}}{2.164s+1} \\ \dfrac{0.397e^{-0.0347s}}{1.542s+1} & \dfrac{-9.44e^{-06}}{2s+1} \end{bmatrix}$$

FIGURE 4.20 Accurate Model (TF3)

$$[P] = \begin{bmatrix} \dfrac{-2892e^{-2.44s}}{2.236s+1} & \dfrac{-0.05432e^{-0.755s}}{2.29s+1} \\ \dfrac{0.432e^{-0.0178s}}{1.768s+1} & \dfrac{-1.889e^{-05}}{2s+1} \end{bmatrix}$$

FIGURE 4.21 Mismatch Model (TF2)

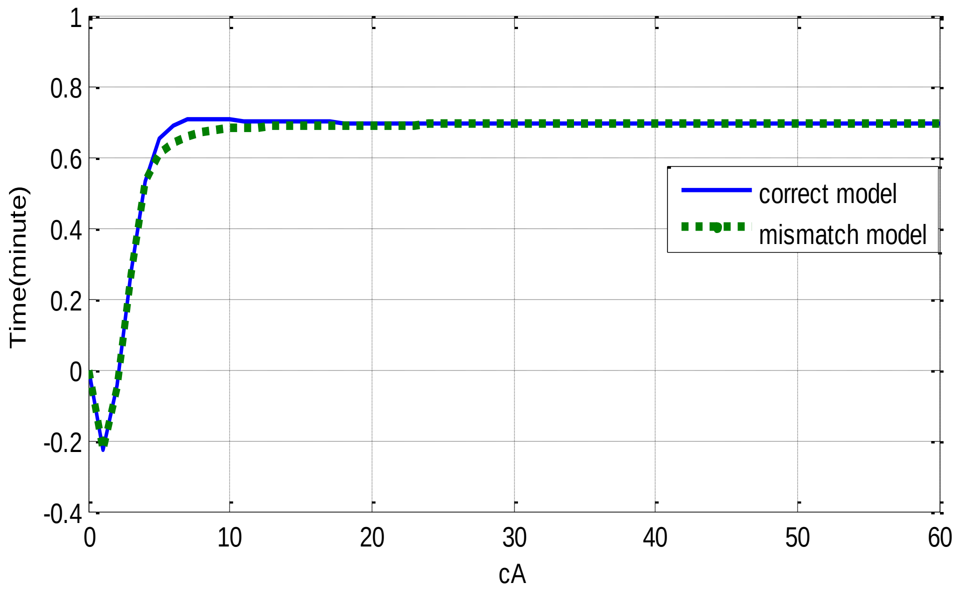

FIGURE 4.22 Graph of MPC Performance in Accurate Model and Mismatch Model

Form Figure 4.22, when a correct model is used, the output reaches steady state in less than 10 minutes. However, when a mismatch model is used, it took more than 20 minutes for MPC to bring the output to the new set point. It is clearly shown that performance of MPC is disrupted by the mismatch model.

CHAPTER 5
CONCLUSION

The accuracy of the plant model used for design of the controller directly affects the performance of the controller. However, the dynamic behaviour of processes may change with time. The deviation of the plant model from the plant is called model plant mismatch. In this research, the researcher will study the effect of model plant mismatch on MPC performance.

The objective of this research is to; 1) develop a non-linear model of Continuous Stirred Tank Reactor using SIMULINK, 2) design and implement a model predictive controller (MPC) using a linear transfer function model, 3) introduce mismatch on the plant and 4) determine how the controller performance deteriorates when mismatch is introduced. All objectives for this research is achieved.

During the conduct of this research, the researcher has developed a non-linear CSTR model by using SIMULINK. Manipulated variable and controlled variable for the CSTR model has been set by the researcher. Besides that, the researcher developed 3 different linear transfer function model using 3 different ranges. By using this 3 different transfer function model, the researcher designed 3 different MPC. The researcher has tested the plant model with 2 different tests. First, to understand the dynamic model of this CSTR, the researcher has done an open loop test to this CSTR model by adding few percentages of increment in step change to the plant input. For the second test, the researcher done a closed loop test to measure the performance of MPC between the accurate plant models and mismatch plant models.

In the open loop test, when step change is added to the plant input, all output increase or decrease from its set point which clearly shows the non-linearity behaviour of the plant. For the MPC performance test, when mismatch is added, it took longer time for MPC to bring the output to the new set point and reaches steady state. It is clearly shown that performance of MPC deteriorates when mismatch plant model is used.

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