Engineering Honours Thesis Report is submitted to the School of Engineering and Information Technology, Murdoch University, in partial fulfilment of the requirements for the Engineering Honours Degree.

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Author's Declaration

Author, Dawood Al-Kahali, declares that, apart from appropriately referenced quotations and citations, this thesis report is my own work and meet the terms with Murdoch University's academic integrity commitments and any other conditions of submission as attached to the assignment. It has not been submitted previously for assessment in another unit. I have read and understood Murdoch's Assessment Policy.

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The main body of this document has a word count of approximately 15 550 words
ABSTRACT

Process control and performance analysis are the key features of any industry. This report provides an analysis and evaluation of several methods for introducing conventional and advanced control schemes to an existing production plant for Murdoch Chemicals Ltd. The existing production plant is comprised of three heated tanks and one dye tank.

The aim of this simulation project is to study the behaviour of the system and design of different conventional and advanced controllers for the system. By performing mass, component and energy balances, the steady state system was represented by a set of first order differential equations and evaluated using MATLAB. Suitable performance measures are then implemented to evaluate the performance of a control system in relation to error analysis and disturbance rejection.

All of the project objectives have been achieved successfully. The GMC is the best implementation on the heated tank system given that the control values can cope up the limits of the manipulated variables changes.
Acknowledgements

I am deeply indebted to my academic supervisor, for her valuable advices, directions, encouragement and continuous support that she has given me.

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Finally, I want to take the opportunity to thank my family for their constant encouragement and dedicated guidance.
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<th>Explanation</th>
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<tbody>
<tr>
<td>AUTO</td>
<td>Automatic</td>
</tr>
<tr>
<td>MATLAB</td>
<td>Matrix Laboratory</td>
</tr>
<tr>
<td>Murdoch</td>
<td>Murdoch University</td>
</tr>
<tr>
<td>MV</td>
<td>Manipulated Variable</td>
</tr>
<tr>
<td>PV</td>
<td>Process Variable</td>
</tr>
<tr>
<td>SP</td>
<td>Set point</td>
</tr>
<tr>
<td>DV</td>
<td>Disturbance Variable</td>
</tr>
<tr>
<td>OP</td>
<td>Operating Point</td>
</tr>
<tr>
<td>LT</td>
<td>Level Transmitter</td>
</tr>
<tr>
<td>TT</td>
<td>Temperature Transmitter</td>
</tr>
<tr>
<td>LC</td>
<td>Level Controller</td>
</tr>
<tr>
<td>TC</td>
<td>Temperature Controller</td>
</tr>
<tr>
<td>P Controller</td>
<td>Proportional Controller</td>
</tr>
<tr>
<td>PI Controller</td>
<td>proportional–integral Controller</td>
</tr>
<tr>
<td>PID Controller</td>
<td>proportional–integral–derivative Controller</td>
</tr>
<tr>
<td>u</td>
<td>Input= MV (manipulated variable)</td>
</tr>
<tr>
<td>y</td>
<td>output= PV (Process variable)</td>
</tr>
<tr>
<td>α</td>
<td>Time delay</td>
</tr>
<tr>
<td>Pu</td>
<td>Ultimate Period</td>
</tr>
<tr>
<td>Kcu</td>
<td>Ultimate Gain</td>
</tr>
<tr>
<td>NT</td>
<td>Needle Tank</td>
</tr>
<tr>
<td>CSTR</td>
<td>Continuous Stirred Tank Reactor</td>
</tr>
<tr>
<td>C</td>
<td>Celsius</td>
</tr>
<tr>
<td>s</td>
<td>Second</td>
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<td>meter</td>
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CHAPTER 1

INTRODUCTION

1.1. Overview

1.2. Project Background

1.3. Project Scope

1.4. Project Objectives

1.5. Thesis Organisation
INTRODUCTION

1.1 Overview

The following is an introductory section detailing the background of the thesis and project objectives. The thesis background is based on knowledge of instrumentation and control systems, and reference is made to the operation and capability of the Pilot Plant at Murdoch University. This plant is a simulation of what students can expect out there in the field and offers a hands-on experience to the students.

1.2 Project Background

Figure 1 shows a real photo of the Engineering Pilot Plant at Murdoch University. This plant has been modelled to simulate the Bayer process which is used to extract alumina from mined bauxite [2]. It is an ideal place to practically learn the instrumentation and control process as the plant has been set to operate on automation and is actuated by some set parameters, which give variable inputs determined by the operator.

The plant combines the knowledge from Honeywell’s Experion Process Knowledge Suite (PKS) with a Distributed Control System (DCS). At the moment, the plant uses Experion as a means of monitoring and controlling the plant [6]. The University has acquired a simulation license for the Experion system, enabling one to program, monitor and simulate control strategies and parameters. Pilot Plant is a key in exposing the students to how an industrial plant’s control system is configured and run.
This project is a continuation of previous student who is called David Pol. He has designed and installed conductivity sensors in the three heated tanks. In addition, he has designed a proportional Integral (PI) controller using experion and Microsoft Excel Data Exchange (MEDE) softwares in order to control the conductivity in the three heated tanks.

1.3 Project Scope

The material and energy balances for any process systems, especially in reactors or heated tanks, are most important in plant design and the starting point for control design. Thus, the choice of control parameters are crucial aspects for developing a sustainable product, process development and optimization [3].
The following are the project approaches:

i. Develop dynamic model of the overall system.

ii. Studying the dynamic nature of the system,

iii. Use dynamic model for identification and controllability experiments - avoid disrupting production.

iv. Implementing and tuning PI controllers for each process variable.

v. Implementing and tuning GMC controllers for each process variable.

vi. Evaluating the performance of the controllers by introducing disturbances and set-point change in process variables.

1.4 Project Objectives

This project will attempt to identify possibilities and issues to be aware of when designing conventional and advanced control schemes, in the Engineering Pilot Plant, and check the behaviour of conductivity response to process control experiments.

A further objective is to ensure that every stage of the project followed by a methodical approach and is appropriately tested before progressing further. For this project, the methodical (Systematic) approach involved:

- Gaining a comprehensive understanding of conductivity sensors and software that will be used (Mat lab software);
INTRODUCTION

- Obtaining open loop step response data;

- Getting mass and energy balances around required systems;

- Designing conventional controllers (PI);

- Designing GMC;

- Analysing performance of controllers to the changes in conductivity and other process conditions;

1.5 Thesis Organization

This thesis report consists of seven parts, which include nine chapters. The thesis report has been organized as follows:

1. First part is an introductory part that contains the first two chapters:
   - **Chapter 1**: presents the introduction and project background.
   - **Chapter 2**: is a review of the general theoretical concepts, which would help in analysing and modelling the three hated tanks.

2. Second part consists of chapter three only.
   - **Chapter 3**: presents the literature review.

3. Third part is about model development and dynamic run, which includes chapter 4 only.
   - **Chapter 4**: presents the detailed derivation of the mathematical model that describes the operation of the three hated tanks under transient conditions and the proposed method used to control this model.
4. Fourth part presents the design work conducted in this project. It includes a detailed description of the design setup and the design procedures that were done. This part consists of the following three chapters:

- **Chapter 5**: Presents Plant wide control requirements.
- **Chapter 6**: illustrates the design, implementation and testing results of PI controllers.
- **Chapter 7**: illustrates the design, implementation and testing results of GMC controllers.

5. Fifth part presents the analysis and discussion of the controller performance. It contains chapter eight.

- **Chapter 8**: shows a comparison between the controllers using different control performance analysing techniques.


- **Chapter 9**: provides the conclusions, recommendations and future work discussion.

7. Seventh part presents the appendices.
CHAPTER 2

THEORETICAL MODELS OF CHEMICAL PROCESS

2.1. The Rational for Dynamic Process Model

2.2. General Modelling Principles

2.3. Degree of Freedom Analysis
INTRODUCTION

2.1 The Rational for Dynamic Process Model

Process dynamic models play a central role in the subject of process dynamics and control. The dynamic models can be used to:

1. Develop understanding of the process. Dynamic models and computer simulation allow transient process behavior to be investigated without having to disturb the process. Computer simulation allows valuable information about dynamic and steady-state process behavior to be acquired, even before the plant is constructed [22].

2. Train plant operating personnel. Process simulators play a critical role in training plant operators to run complex units and to deal with emergency situations. By interfacing a process simulator to standard process control equipment, a realistic training environment is created [5].

3. Develop a control strategy for a new process. Process dynamic model allows alternative control strategies to be evaluated. For instance, a dynamic model can help to identify the process variables that should be controlled and those that should be manipulated. For model-based control strategies, the process model is part of the control law.

4. Optimize process operating conditions. It can be advantageous to recalculate the optimum operating conditions periodically in order to maximize profit or minimize cost. A steady-state process model and economic information can be used to determine the most profitable operating conditions.
INTRODUCTION

2.2 General Modelling Principles

It is significant to remember that a process dynamic model is a mathematical abstraction of a real process. The process model equations are the best approximation to the real process as expressed by the adage that "all models are wrong, but some are useful" [5]. Consequently, the model cannot incorporate all of the features, whether macroscopic or microscopic, of the real process. Modelling inherently contains a compromise between model complexity and accuracy on one hand, and the cost and effort required to develop the model and verify it on the other hand. The required compromise must take into consideration the following factors:

- The modelling objectives,
- The expected benefits from use of the model, and
- The background of the intended users of the model (for example; plant engineers versus research specialists) [5].

2.3 Degree of Freedom Analysis

To simulate a process, it is necessary first ensure that its model equations (differential and algebraic) constitute a solvable set of relations. In other words, the output variables, normally the variables on the left side of the equations, could be solved in terms of the input variables on the right side of the equations.
In order for the model to have a unique solution, the number of unknown variables must equal the number of independent model equations. An equivalent statement is that all of the available degrees of freedom must be utilized [5]. The number of degrees of freedom, \( N_{F'} \) can be calculated from the expression (2-1)

\[
N_{F'} = N_V - N_E
\]  

(2 – 1)

Where;

\( N_V \) is the total number of process variables and

\( N_E \) is the number of independent equations.

A degrees of freedom analysis allows modeling problems to be classified according to the following categories:

i. \( NF = 0 \): The process model is precisely identified. If \( NF = 0 \), then the number of equations is equal to the number of process variables and the set of equations has a solution.

ii. \( NF > 0 \): The process is underspecified. If \( NF > 0 \), then \( N_V > N_E \), so there are more process variables than equations. Accordingly, the \( N_E \) equations have an infinite number of solutions, because \( NF \) process variables can be identified randomly.
iii. $NF < 0$: The process model is over specified. For $NF < 0$, there are fewer process variables than equations, and therefore the set of equations has no solution.

Degrees of Freedom Analysis can be determined by track the following procedure:

1. List all quantities in the model that are known constants (or parameters that can be specified) based on equipment dimensions, known physical properties, and so on [5].

2. Define the number of equations $N_E$, and the number of process variables, $N_V$.

3. Determine the number of degrees of freedom using equation $(2 - 1)$,

4. Identify the $N_E$, output variables that will be found by solving the process dynamic model.

5. Identify the $NF$ input variables that must be identified as either disturbance variables (DV) or manipulated variables (MV), in order to utilize the $NF$ degrees of freedom.
3.2 Overview

3.3 Background of the Engineering Pilot Plant

3.4 Process Modelling Approaches – Modelling of the three heated Tanks

3.5 Control Approaches Review
LITERATURE REVIEW

3.1 Overview

This chapter discovers the literature that is relevant in order to understand the development and interpreting the results of this project. This chapter delivers an overview of previous researchers on the Engineering Pilot Plant that includes the following:

i. Description and background review of the process plant;

ii. Overview of the Honeywell Experion PKS;

iii. Process Modelling Approaches of the three heated tanks;

iv. Control Approaches which include:
   - Control Loop Selections
   - Tuning Methodologies
   - Control strategies which include:
     - Conventional Control
     - Generic Model Control
3.2 The Engineering Pilot Plant

3.2.1 Pilot Plant Background

A schematic diagram of the pilot plant which is located in the Engineering Building at Murdoch University is shown in Figure 2. The Supply Tanks are the primary source for providing the facility with the feed. Alternate tanks can be utilised to provide a source of disturbances to test control strategies [1].

![Pilot Plant Schematic Diagram](image-url)
The Pilot Plant has two large storage tanks that supply the system with water. Water leaves the supply tanks through an electric pump, which is controlled by a variable speed drive, to the Ball Mill. Once the Ball Mill has reached its capacity, it overflows into the Ball Mill Tank [1]. Ball mill and Ball mill tank are shown in Figure 3.

This tank has a level sensor, density sensor and a cone-shaped base in which the Ball Mill Pump is attached. The Ball Mill Pump is also controlled by a variable speed drive and propels water into a cyclone valve [2].

The cyclone valve, depending on the calibration, sends a percentage of water in one direction when the inflow is above a particular flow rate. The cyclone underflow diverts water downwards into the Cyclone Underflow Tank [2].

This tank also has a level sensor, density sensor, and a cone-shaped base. Leaving Cyclone Underflow Tank are two variable speed driven pumps, one recycles back to
the Storage Tank, and the other recycles back to the Ball Mill Tank. The Cyclone overflow sent into a Lamella Tank. This tank is a square-shaped vessel with a base that resembles an upside-down pyramid [2].

The water can be pumped from this tank back to the storage reservoir, or it can overflow into the Needle Tank [2]. Figure 4 below shows Cyclone under Flow Tank, Lamella Tank, and Needle Tank.

An important note about the Lamella Tank is that inflow pipe is curved into the tank itself and situated at the bottom. Therefore, it causes the tapping effect when the Ball Mill Pump is turned off causing the Lamella Tank to empty back into the Cyclone Underflow Tank [2].

Once adequate water is in the Lamella Tank, the overflow enters the Needle Tank [2]. This tank is a long, lean tank and also contains a level sensor, density sensor

![Figure 4: Cyclone under Flow Tank, Lamella Tank, and Needle Tank [2].](image-url)
and cone-shaped base. This vessel can also be filled up with the Non-linear Supply Tank. At the base of the Needle Tank, there is a pump which is also VSD driven. The VSD driven pump drives water into the Continuously Stirred Tank Reactor (CSTR) system (Heated Tanks).

The heated tank system includes three heated tanks that containing temperature sensors. The three tanks are cascaded downwards, the highest tank receiving flow from the Needle Tank. Once this tank has reached its capacity, it overflows into the second heated tank, and finally, once that has reached capacity it overflows into the final tank which contains a level sensor. The water is now pumped to the drain. Figure 5 shows the three steam-heated tanks while Figure 6 shows the entire Pilot Plant setup.

Figure 5: Three Steam-Heated Tanks (CSTRs) [2].
2.2.1.1 Alumina Refinery

The pilot plant at Murdoch University is a miniaturised physical setup of the common industrial alumina refinery [2]. An alumina refinery creates alumina from mined bauxite through a chemical procedure known as the Bayer process [3]. The pure alumina is used to produce aluminium metal which is utilised in many applications.

2.2.1.2 Bayer Process

The process derives its name from the Austrian Chemist Karl Josef Bayer, who initially developed and patented the process around 1887 (British patent) and has since been popularly used in the production and beneficiation of Aluminium from bauxite [3][4]. The process is traditionally composed of some stages, which includes milling, digestion clarification, precipitation, evaporation, classification and calcification as shown in Figure 7.
3.2.2 Honeywell Experion Process Knowledge (PKS)

3.2.2.1 Overview of the Honeywell Experion Process Knowledge System

Honeywell’s Experion Process Knowledge (PKS) System is a Distributed Control System (DCS) integrating an advanced process automation platform with some software applications used for the configuring, monitoring and control of an industrial process [6].

A DCS characterises a system controlled by one or more controllers distributed throughout a process, connected to a network to acquire, communicating and monitoring data [7]. A simplified DCS representation is shown in Figure 8.
3.2.2.2 Honeywell Experion Station

The Honeywell Experion PKS C300 control system had been installed to run the pilot plant [6]. An overview of the control architecture can be viewed in Figure 9. The control architecture displays how the pilot plant server communicates along a bus to all the Honeywell controllers which are then connected to various transducers and actuators.
The software provided for the operator to communicate with this system is the Honeywell Experion Station. The station software communicates to the pilot plant server to change the values of various state variables. Within this software, a page displays the overview of the plant and some others that show different segments in more details [6]. These can be seen in Figure 10 to Figure 12.

Figure 10 shows the overview page which provides level indications on all of the tanks. So, these levels can be monitored without having to switch pages. The trend line screen page can also be used for this and will be discussed later. At the top left corner of the page, there are navigation buttons that can be used to switch pages.

The following pages can also be navigated through via the arrows protruding in and out of the various flow lines.
Figure 10: Overview of Pilot Plant in Honeywell Station [2].

Page number five displays the Needle Tank (NT) and Nonlinear Tank (NLT). Flow into the NT can be monitored with FT_401 and the flow out can be controlled with the NT Pump. This pump provides flow to the BMT and the Continuously Stirred Tank Reactors (CSTR’s) depending on the position of the feed valve to the CSTR’s. The pump on the underflow of the NLT can be used to inject water into the NT and CSTR’s. These flows can be monitored with FT_523 and FT_569. Figure 11 shows Needle Tank screenshot.
The final page contains all three Heated Tanks (CSTRs). These tanks have control valves for steam inflow which are labelled FCV_622, FCV_642 and FCV_662. The temperature of the material in the tanks can be controlled by these valves and monitored by the indication on the tank. There is a pump controlling the outflow from CSTR 3 and two valves, FCV_690 and FCV_688, which can defer water from the pump back to the upper CSTR’s or out of the system as a product. The flow through these valves can be monitored with FT_687 and FT_689. Figure 12 below shows CSTR Bank Control screenshot.
3.2 Process Modelling Approaches – Modelling of the three heated Tanks

3.2.1 General Modelling Concept

In order to study the behaviour of a system, a set of the mathematical equation describing the dynamic behaviour of the system is needed. The model can be used to investigate the operation of the system at different set points. A mathematical model can be developed from mass balance and energy balance equations [18] which will be based on in this thesis project. However, there are different ways to obtain a model of the process which are as follow:
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- Open Loop Step Test followed by Auto Relay tuning which has been done by Ben [1] using Microsoft Excel Data Exchange (MEDE) software.
- Linear Least Squares Regression Method and solving for a first order with delay model that has been done by Pol [9] using Simulink and Experion software. Moreover, Pol has done open loop step test using MEDE software.

3.2.2 Heated Tank’s Level and Temperature Models

Before implementing any control scheme on the plant, data was taken by Ben in order to investigate how the plant operates in an open-loop state [1]. This data was later used to validate the open-loop model. Each of the main pumps and valves were stepped and the plant data was recorded.

In order to check the behaviour of the heated tank 3 level and the temperatures in all of the three heated tanks, open loop tests have been done for all of the process variables. Figure 13 shows a sample open loop test. Open loop tests have done by Ben are as follow:

- Product Pump (PP_681) has been stepped up by 5% (from 15% to 20%).
- Steam Valve (FCV_622) has been stepped up by 5% (from 10% to 15%).
- Steam Valve (FCV_642) has been stepped up by 5% (from 10% to 15%).
- Steam Valve (FCV_662) has been stepped up by 5% (from 10% to 15%).
3.2.3 Heated Tank's Conductivity Models

3.2.3.1 Conductivity Concept

Conductivity or conductance of an electrolyte solution can be defined as a measure of its ability to conduct electricity [9]. The SI unit of conductance is Siemens while a fluid conductivity is expressed in micro Siemens per centimetre (µS/cm) [11].

3.2.3.2 Conductivity Measurement Principle

Conductivity measurements are usually used in many environmental and industrial applications due to a fast, cheap and dependable method of measuring the ionic
content in a solution such as the measurement of product conductivity is a typical way in order to monitor and consistently trend the performance of water purification systems [12].

The electrical conductivity of an electrolyte solution can be measured by defining the solution resistance between two cylindrical or flat electrodes separated by a fixed distance as shown in Figure 14 [13]. An alternating voltage (AC) is used for the measurement rather than the direct voltage (DC) because DC in would cause electrolysis [14]. The resistance can be measured by a conductivity meter [13].

![Figure 14: Conductivity Measurement Principle [13].](image-url)
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3.2.3.3 Conductivity Sensors Implementation

It was required to install three Indumax CLS50D conductivity sensors by Pol [9] in the three heated tanks in the pilot plant, obtain data and analyse results before developing simulations and control strategies.

3.2.3.3.1 Physical Installation

Figure 15 shows the location where the conductivity sensor was installed on heated tank 2 in the pilot plant.

![Figure 15: Installed Conductivity Sensor in Pilot Plant Heated Tank 2 [9]](image)

Four factors have been taken into consideration when choosing a location to install the sensor in, which are as follows [9]:

i. Proximity to agitator
   - To mitigate any risk of damage;

ii. Proximity to steam coil;
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- To provide a medium for the steam coil to cool rapidly

iii. Proximity to metal surfaces
- Due to the measuring principle of the instrument;

iv. Cable route.
- Due to an installation factor to measured values

The temperature was considered, as according to instrument (Indumax CLS50D) specifications which can be seen in Appendix B. The sensor has a maximum process temperature of 125 Degree C [9].

3.2.3.4 Applications

Measured conductivity gives a good indicator of the availability of conductive ions in solution. Conductivity measurements are used widely in many industries [12]. They are used:

- to monitor quality in public water supplies,
- in hospitals,
- in boiler water and
- in industries which depend on water quality such as brewing.

This type of measurement (brewing) is considered as not ion-specific. In spite of the fact, if the solution composition and its conductivity behaviour are known, it is sometimes used to find the amount of total dissolved solids (T.D.S.) [11].
3.2.3.5  **Effect of Temperature**

The significant factor affecting conductivity is temperature, such that, a high temperature increases the rate of flow of the ions between the plates, hence a higher conductivity. Given that conductance is a measure of ions in the fluid, it becomes an important parameter in the liquid analysis as it can be used to determine the degree of contaminant concentration and chemical concentration in the fluid under study. Thus, one can come up with a chart of purity levels of the fluid depending on the conductivity of the fluid under specified atmospheric conditions.

It is a common practice in water purification plants where, for example, bottled water can be classified based on the concentration of minerals in the water. The conductivity of a solution is adjusted to show what it would be at 25 Degree C and enables a conductivity measurement to show more reliably if a change in conductivity is due to a change in concentration, not temperature [15].

3.2.3.6  **Effect of Electrolyte Solution**

A suitable chemical is required to manipulate the conductivity of the process liquid. As a solid conductivity of depends on the free movement of electrons, the fluid conductivity depends on the free movement of ions. These liquids are known as electrolytic solutions [16].

An electrolyte is a substance whose aqueous solution dissociates into ions [9]. They can be divided into acids, bases and salts. More specifically, an acid is an ionic compound that yields hydrogen (H+) ions when dissolved, a base provides an
excess of hydroxide (OH-) ions, and salt is an ionic compound commonly produced by an acid-base reaction [17].

The same qualities that make these substances conductive are also what makes them have corrosive or oxidising properties. So, when charged ions can move around, it will increase the rate of ionic reactions and electrolysis [17]. It is well known that salt water improves the rate of corrosion on many metals, notably iron, as well as an acids ability to dissolve metal resulting in hydrogen gas and salt. Obviously, this is an undesired situation in the pilot plant.

There are different electrolytes which are as follows [9]:

i. Sodium Chloride (NaCl) – Common table salt;
ii. Potassium Chloride (KCl) – Readily available salt;
iii. Sodium Hypochlorite (NaOCl) – Commonly available as disinfectant and pool chlorine;
iv. Sodium Bicarbonate (NaHCO3) - Bicarbonate soda;
v. Acetic Acid (CH3COOH) – Household vinegar;
vi. Hydrochloric Acid (HCl) - Pool Acid.

3.2.3.7 Concentration versus Conductivity

Conductivity as defined previously, it is the measurement of how well electrical current passes through. It is in direct relation to the ion concentration of most solutions. The higher the ion concentration in the solution, the better electricity it will conduct.
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Although some highly concentrated solutions do not follow the linear relationship of concentration-to-conductivity, the temperature can affect conductivity. So, a standard conversion factor can be used to make the best estimation of concentration in case if conductivity is known [16].

Specific conductance of any solution contains one electrolyte which depends on the concentration. Therefore, the following equation shows that it is convenient to divide the specific conductance by concentration [12].

\[
\Lambda_m = \frac{k}{C}
\]  

(3—1)

Where;

\(\Lambda_m\) is molar conductivity

\(k\) is a specific conductance

\(C\) is solution concentration

3.2.3.8 Conductivity Models

The dye tank was calculated to have a volume of 20.5L calculated from the diameter and height measurements of 295mm and 300mm respectively. Based on this, a solution was created comprising 325g of table salt and 18L of tap water. These were arbitrary quantities chosen based on the equipment that was available at the time [9]. This test was conducted by setting dye tank pump speed to 100% with all flow going
to CSTR3. It was intended only to find the maximum and minimum values of conductivity, to assess if the concentration of NaCl solution was adequate to achieve a measurable response [9].

Multiple steps were performed to obtain the process response and model when the system is subject to different levels [9].

The average of all derived process models was taken to create a generalised transfer function that could be used for model-based tuning [9]:

\[
G_p = \frac{46 \times e^{-7s}}{533s + 1} \tag{3-2}
\]

This model was used as the basis for all model-based controller tuning [9].

---

**Figure 16:** Conductivity Open Loop Simulink Model [9].
3.2.4 Software Platform

3.2.4.1 Microsoft Excel Data Exchange (MEDE)

As mentioned previously in process model identification, Ben was used Microsoft Excel Data Exchange software. MEDE is identified as one of Honeywell’s preferred forms of data exchange. MEDE allows for the capture of real-time data point values and historical information from the Experion server, for display within an Excel spreadsheet [6]. Additionally, MEDE provides the ability to send values to data points on the Experion server. Figure 16 displays the resulting process variable response (PV), from a step in input, or manipulated variable (MV), to a first order system control module.

3.2.4.2 Simulink

Also, mentioned previously in process model identification, Pol was used Simulink software. Simulink is contained within the MATLAB umbrella is MathWorks’ block diagram environment for simulation and control, Simulink. Simulink offers a range of function block libraries and solvers for the effective modelling, simulation and control of dynamic systems [6]. Simulink was used extensively for testing processes and control strategies developed within the Experion environment.

3.2.4.3 Experion

The process model was also implemented using the Experion first order simulation Control Model [9]. Experion software detailed explanation can be found in section (3.2.2).
3.3 Control Approaches Review

3.3.1 Control Loop Pairing (Loop Selections)

The control loop pairings that were selected for the automatic control of the heated tanks in the pilot plant can be seen in Table 2. These particular control loops leave the needle tank pump (NTP_561) and recycle flow rate as free variables that can be used as disturbance variables by manually adjusting at any point to change the flow out of the pumps.

Table 2: Control Loop Pairing.

<table>
<thead>
<tr>
<th>Process Variable (PV)</th>
<th>Manipulated Variable (MV)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Heated Tank-1 Level (LT_667)</td>
<td>Product Pump (PP_668)</td>
</tr>
<tr>
<td>Heated Tank-1 Temperature (TT_623)</td>
<td>Steam Valve-1 (FCV_622)</td>
</tr>
<tr>
<td>Heated Tank-2 Temperature (TT_643)</td>
<td>Steam Valve-2 (FCV_632)</td>
</tr>
<tr>
<td>Heated Tank-3 Temperature (TT_663)</td>
<td>Steam Valve-2 (FCV_662)</td>
</tr>
<tr>
<td>Heated Tank-1 Conductivity (CA_623)</td>
<td>Dye Tank Pump (DP_611)</td>
</tr>
<tr>
<td>Heated Tank-2 Conductivity (CA_643)</td>
<td>Dye Tank Pump (DP_611)</td>
</tr>
<tr>
<td>Heated Tank-3 Conductivity (CA_663)</td>
<td>Dye Tank Pump (DP_611)</td>
</tr>
</tbody>
</table>

3.3.2 Tuning Methodologies

3.3.2.1 Auto Relay Tuning method

3.3.2.1.1 Overview

PI control was implemented in the spreadsheet by using the Velocity Form [1]. The velocity form calculates the change in the controller position at each sampling point.
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and adds it to the previous value which then sent to the actuator as the controller action. This is why there is a page in the spreadsheet that is dedicated to retrieving historical values from the server [1]. The equation for the velocity form is shown below [10]:

\[
\Delta u(k) = K_c \left[ \left( 1 + \frac{\Delta t}{\tau_i} + \frac{\tau_d}{\Delta t} \right) \varepsilon(k) - \left( \frac{2\tau_d}{\Delta t} + 1 \right) \varepsilon(k - 1) + \frac{\tau_d}{\Delta t} \varepsilon(k - 1) \right]
\]

(3 - 3)

Where:

\( K_c \) is the controller gain;

\( \tau_i \) is the integral time;

\( \tau_d \) is the derivative time;

\( \Delta t \) is the timeframe in which the program is recalculating;

\( \varepsilon(k) \) is the error signal at a \((k)\) point in time;

and \( \Delta u(k) \) is the change in the manipulated variable that should be implemented in the current cycle [6].

Relay tuning was chosen to tune the controllers as this method can be used to tune controllers without the process model [1]. The following steps were taken to implement relay tuning:

i. Identify the process variable and manipulated variable.

ii. Set the high limits and low limits on both the process variable and actuator action.

iii. Write an equation in Excel spreadsheet to calculate values to be sent to the actuator. For example, if the process variable is less than the lower limit, then the actuator action must be set to its high point, and if the process variable is
greater than the upper limit, then the actuator action must be set to its high limit. Otherwise, the actuator action must take the previous value.

iv. Obtain a graph of both process variable and actuator action versus time. A sample chart is shown in Figure 17.

![Graph of process variable and actuator action versus time](image)

**Figure 17: Relay Tuning Illustration [10].**

As can be seen in Figure 17 above, the components needed to calculate the parameters of the controller are

- A is the amplitude
- $P_u$ is the period
- $2h$ is the difference in the actuator limits.
The following two equations are used to calculate the ultimate gain and period

\[ K_{cu} = \frac{4h}{\pi A} \]  

\[ P_u = \text{ultimate Period} \]

v. To obtain the control parameters, the Ziegler-Nichols tuning rules were applied as can be seen in Table 3.

<table>
<thead>
<tr>
<th>Controller Type</th>
<th>( K_c )</th>
<th>( \tau_i )</th>
<th>( \tau_d )</th>
</tr>
</thead>
<tbody>
<tr>
<td>P</td>
<td>0.5K_{cu}</td>
<td>-</td>
<td>-</td>
</tr>
<tr>
<td>PI</td>
<td>0.45K_{cu}</td>
<td>( P_u/1.2 )</td>
<td>-</td>
</tr>
<tr>
<td>PID</td>
<td>0.6K_{cu}</td>
<td>( P_u/2 )</td>
<td>( P_u/8 )</td>
</tr>
</tbody>
</table>

2.4.2.1.2 Finding Level Controller Parameters of Heated Tank 3 – Using Auto Relay Tuning Method

In order to determine correct tuning parameters for each of the controllers, it was decided to perform the Auto Relay Tuning method on each of the tanks [1]. Auto relay tuning is a method of tuning that involves forcing the process variable to oscillate about a desired set point by aggressively changing the manipulated variable, as can be seen in Figure 18 [6].

Figure 18 shows a graph of water level and Production pump action versus time for Heated Tank 3.
Once this has been performed, the data is analysed to determine the ultimate period of oscillation (\(P_u\)) as well as the amplitude of the wave. From the plot in Figure 18, the parameters are extracted to obtain the following values:

\[
A = \frac{95-90}{2} = 2.5
\]
\[
H = \frac{10-1}{2} = 4.5
\]
\[
P_u = 880 - 472 = 408
\]
Using Table 3 Ziegler Nichols Tuning Rules to obtain the values of (Kc) and $(\tau_i)$

\[ K_{cu} = \frac{4 \times h}{\pi \times A} \]

\[ K_{cu} = \frac{4 \times 4.5}{\pi \times 2.5} \]

\[ K_c = 2.29 \]

\[ \tau_i = \frac{408}{1.2} = 340 \text{ sec} \]

As the controllers are PI, the PI controller tuning parameters were used and control parameters were obtained for each controller [1].

**Table 4: Auto Relay Tuning Results [1]**

<table>
<thead>
<tr>
<th>Tank Name</th>
<th>$P_u$</th>
<th>$h$</th>
<th>$A$</th>
<th>$K_{cu}$</th>
<th>$K_c$</th>
<th>$\tau_i$</th>
</tr>
</thead>
<tbody>
<tr>
<td>Heated Tank-1 (Level)</td>
<td>408</td>
<td>4.5</td>
<td>2.5</td>
<td>5.7096</td>
<td>$-2.29$</td>
<td>340</td>
</tr>
<tr>
<td>Heated Tank-1 (Temperature)</td>
<td>319</td>
<td>10</td>
<td>0.385</td>
<td>33.0712</td>
<td>14.8820</td>
<td>265.8333</td>
</tr>
<tr>
<td>Heated Tank-2 (Temperature)</td>
<td>285</td>
<td>10</td>
<td>0.54</td>
<td>23.5785</td>
<td>10.6103</td>
<td>237.501</td>
</tr>
<tr>
<td>Heated Tank-3 (Temperature)</td>
<td>220</td>
<td>10</td>
<td>0.53</td>
<td>24.0234</td>
<td>10.8105</td>
<td>183.333</td>
</tr>
</tbody>
</table>
3.3.2.2 **Direct Synthesis Method**

The direct synthesis approach to control design has been chosen when developing control algorithms for conductivity [9]. Direct synthesis can be defined as a model based tuning method that provides the ability to choose the precise response of the process by tuning the controller with respect to a chosen desired model [41]. Direct synthesis was chosen instead of other tuning methods due to the ease of implementation, availability of process models and the ability to select a desired response. The following section details the derivation of tuning parameters. For this system, a first order response was desired; therefore, no overshoot or oscillations were expected.

A process reaches 99.7% of its final value in approximately $5\tau$ [9]. Therefore, a $\tau$ value of 150 would depict a rise time of 750s or 12.5 minutes. This was the original value chosen and a simplified derivation of controller tuning parameters is shown below:

\[
\text{Process Response} = G_p = \frac{46}{522s + 1} \tag{3 - 5}
\]

\[
\text{Process Response} = q(s) = \frac{1}{150s + 1} \tag{3 - 6}
\]

\[
\text{Controller Required} = G_c = \frac{1}{g_p \left( \frac{q}{1 - q} \right)} \tag{3 - 7}
\]

**Solving for $G_c$:**

\[
G_c = \frac{1}{46} \left( \frac{1}{522s + 1} \right) \left( \frac{1}{150s + 1} \right) \left( \frac{1}{1 - \frac{1}{150s + 1}} \right) \tag{3 - 8}
\]
The full derivation of tuning parameters for a first-order system can be found in Appendix C. Rise time can be changed easily by changing the tau value of the desired response. This was a major factor in the selection of tuning methods, if upon implementation control action was too aggressive or conservative, it could be easily changed in the future [9].

### 3.3.3 Control Strategies

Two different control schemes will be used and compared in this thesis project which are Conventional Control and Generic Model Control (GMC). These two control schemes have been widely employed in industrial applications for many years, and each has areas in which they are most useful [1].
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3.3.3.1 Conventional Control

3.3.3.1.1 General Concept

There are several closed-loop structures which include feedback, feedforward, cascade feedback and model-based control. The most common and widely employed technique of closed-loop control is feedback. Figure 19 shows the structure of feedback control.

![Figure 19: Structure of the Feedback Control [10].](image)

In Figure 19 above, (y) is the process variable that is to be controlled. So, it comes as close as possible to the setpoint SP. The sensor measures the value of y and sends it to a comparator. The difference between the measured value of (y) and the set point is called an error (e) which is sent to the controller.

The output of the controller is (U) which sent to the actuator for control action. The conventional types of feedback controller are the proportional only controller (P), proportional Integral (PI) and proportional Integral derivative (PID) controller. Following is a brief explanation of the conventional types of feedback controllers.
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3.3.3.1.2 Proportional only Controller

Proportional only control is a simple controller to implement and reacts fast to disturbances. The main problem of the proportional only controller is the offset. The offset is the difference between set-point and the process variable at steady state. As the proportional only controller results in an offset, it makes it difficult for the process variable to reach the set point. The equation of the P controller is shown below [10]:

\[ c(s) = K_c \varepsilon(s) \]  \hspace{1cm} (3 – 12)

3.3.3.1.3 Proportional-Integral Controller

The proportional Integral (PI) controller has both the proportional term and the integral of the error. The integral term eliminates the offset experienced when using the proportional only controller. The fact that a PI controller can keep the process variable at the desired set point makes it a better option than a proportional only controller [10].

The disadvantage of the PI controller is that if not properly tuned it can make the system unstable and slow to respond to the disturbances. The equation of the PI controller is shown below [10].

\[ c(s) = K_c \left( 1 + \frac{1}{\tau_1 s} \right) \varepsilon(s) \]  \hspace{1cm} (3 – 13)
3.3.3.1.4 Proportional-Integral Derivative Controller

The PID has a derivative term which uses the derivative of the error to evaluate the controller action. The derivative term improves the response time to disturbance [10]. If tuned correctly, the PID is better controller as it compensates for the shortfalls of the P and PI controllers. The problem with PID is the derivative term amplifies noise which can make the system unstable. The equation of the PID is given below [10].

\[ c(s) = K_c \left( 1 + \frac{1}{\tau_i s} + \tau_d s \right) \varepsilon(s) \]  \hspace{1cm} (3 - 14)

3.3.3.2 Level and Temperature Control

Once the controller tuning was finished, the new control parameters were entered into each cell for the different stages and the controllers were tested to ensure correct operation. Once everything was confirmed to work correctly, closed-loop test results were taken [1].

3.3.3.3 Conductivity Control

The conductivity control designs have been done by Pol starting with a single-tank control design. It means that each tank would have its independent control mechanisms. Conductivity control was done using the Modbus communication and the conductivity sensors inside the tank. The pilot plant’s Experion network made it possible to integrate the two [9].
2.3.3.3.1 Simulink Implementation

Figure 20 shows developed a simulation model of conductivity with PI Control.

![Figure 20: Conductivity Simulink Simulation with PI Control [9].](image)

Using the developed simulation model where open loop performance had already been proven correct, PI control was added, shown in Figure 40, and then its performance analysed. This was the first step performed to assess if the controller parameters derived would provide the desired response [9].

2.3.3.3.2 Experion Implementation

Once the controller tuning parameters had been confirmed, it was desired to test controller implementation in Experion. This was required as Simulink and Experion implemented PI controllers with different algorithms as shown in Equation (3 – 15) and Equation(3 – 16). Documentation provided by Honeywell specified the control algorithm was of the form shown in Equation 7 as well as stating that T1, integral time, is in minutes [39]. It was not found in MATLAB documentation if integral time was in minutes, but was presumed to be in seconds, and required to be confirmed.
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\[
G_c = K \left(1 + I \times \frac{1}{s}\right) \quad (3-15)
\]

\[
G_c = K \left(1 + \frac{1}{T1} \times \frac{1}{s}\right) \quad (3-16)
\]

where \( T1 \) is integral time in minutes

Final used controller parameters values are shown in Table 5.

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Simulink</th>
<th>Experion</th>
</tr>
</thead>
<tbody>
<tr>
<td>( K_c )</td>
<td>0.07565</td>
<td>3.7825</td>
</tr>
<tr>
<td>( \tau_t )</td>
<td>0.0019557</td>
<td>8.7</td>
</tr>
</tbody>
</table>

3.3.4 Generic Model Control

3.3.4.1 General Concept

Generic Model Control (GMC) was developed for multivariable and nonlinear process model control. This control system is based on the best available process model. The main objective of the control system is to find values for the manipulated variable to follow the desired reference trajectory [19].

The GMC performance objective is to minimise the difference between the desired derivative of the process output and the actual derivative.

The reference system equation is as follows:
The two most important factors considered from this equation are:

1. If the process variable deviates from the setpoint, it acts quickly to bring the setpoint back.
2. Ensure offset free performance.

$K_1$ and $K_2$ values define the trajectory and speed of the response in reference system equation. For the pilot plant, these values are derived from the steady state step changes.

2.6.2.2 GMC Design for Level and Temperature

Advanced control systems (GMC) had not been designed by Ben in order to control the level of heated tank 3 and to control the temperature in all of the three heated tanks. However, Dynamic Matrix Control was implemented on the temperature system of the heated tanks [1]. In this project, GMC will be designed, implemented and tested to control the level and the temperature of the heated tanks.
2.6.2.3 GMC Design for Conductivity

Advanced conductivity control systems (GMC) had not been designed by Pol [9]. However, GMC will be designed, implemented and tested to control the concentration of the three heated tanks in this project. In addition, conductivity will be calculated from the concentration.
PART THREE: MODEL DEVELOPMENT AND DYNAMIC RUN

CHAPTER 4

DEVELOPMENT OF THE PROCESS MODEL AND DYNAMIC RUN

4.1. Overview

4.2. Heated Tanks Process Design

4.3. Development of the Process Model

4.4. Degree of Freedom Analysis

4.4. Dynamic Model Implementation in MATLAB
4.1 Overview

This part, chapter, of this report is associated with studying the dynamic nature of the process. Simulation model for dynamic system will be implemented in MATLAB. The process is modelled by nonlinear set of equations. Regulation of each of these units i.e. heated tank 3 water level, temperatures of all heated tanks, concentrations inside each tank and the conductivities, by itself might seem straight forward. However, when these units are interconnected in a plant, the process variables’ steady-state and dynamic behaviours change significantly due to the high level of interactions among them.

The approach includes the development of the process models, implementation of the dynamic model and studying the dynamic nature of the system.

4.2 Heated Tanks Process Design

The process design of this project consists of three heated tanks concocted in series and one dye tank connected to each of the heated tanks. There are two different feed streams. The first feed stream is non-concentrated stream (Pure water) which coming from the needle tank and it feeds directly into the first heated tank. The second feed stream is concentrated stream (Salty water) which coming from the needle tank and feed to all of the heated tanks. The three heated tanks heat up the feed and sending it as a concentrated product through the outlet of the last heated tank. From the outlet of the third heated tank, a part of the concentrated product is drawn and rest of the product is recycled back into the first heated tank as can be seen in Figure 21.
Figure 21: Heated Tanks Process Design (Author).
4.3 Development of the Model and Mathematical Validation – Modelling of the three heated Tanks

4.3.1 General Modelling Concept

In order to study the behaviour of a system, a set of the mathematical equation describing the dynamic behaviour of the system is needed. The model can be used to investigate the operation of the system at different set points and implementation of various control strategies. A mathematical model can be developed from mass balance and energy balance equations [18).

4.3.2 Heated Tank-1 Models

4.3.2.1 Mass Balance

In order to develop a model for the tank’s level, a mass balance is required.

\[
\frac{dm}{dt} = m_{in} + m_d - m_2 + m_{recycle} \tag{3-2}
\]

Where: \( m = \rho V, \; in = \rho F \) \tag{3-3} 

\[
\rho_{sw1} \frac{dV_1}{dt} = \rho_{pw} \cdot F_{in} + \rho_{sw} \cdot \frac{F_d}{3} - \rho_{sw1} \cdot F_1 + \rho_{sw3} \cdot F_{recycle} \tag{3-4}
\]

Where: \( \rho_{pw} \) is pure water density

\( \rho_{sw} \) is salty water density
4.3.2 Component Balance

\[
\frac{dV_1}{dt} = \frac{1}{\rho_{sw1}} \left[ \rho_{pw} \cdot F_{in} + \rho_{sw3} \cdot \frac{F_d}{3} - \rho_{sw1} \cdot F_1 + \rho_{sw3} \cdot F_{recycle} \right] \quad (3 - 5)
\]

4.3.2.2 Component Balance

\[
\frac{dV_1 C_1}{dt} = V_1 \frac{dC_1}{dt} + C_1 \frac{dV_1}{dt} = F_{in} \cdot 0 + \frac{F_d}{3} \cdot C_d - F_1 \cdot C_1 + F_{recycle} \cdot C_{out} \quad (3 - 6)
\]

Where: \( V_1 \) is constant

\[
\frac{dC_1}{dt} = \frac{1}{V_1} \left[ \frac{F_d}{3} \cdot C_d - F_1 \cdot C_1 + F_{recycle} \cdot C_{out} \right] \quad (3 - 7)
\]

4.3.2.3 Energy Balance

In order to develop a model for the tank’s temperature, energy balance for all of the three Heated Tanks is required which is as follows:

\[
\begin{align*}
\text{Rate of energy accumulation} &= \text{Rate of input of energy} - \\
&\quad \text{Rate of output of energy} + \text{Rate of heat input} - \\
&\quad \text{heat loss to atmosphere} \\
&= \rho_{sw1} c_p \frac{d}{dt} [V_1 T_1] \\
&= \rho_{pw} c_p \cdot F_{in} T_{in} + \rho_{sw3} c_p \cdot F_{recycle} T_{out} + \rho_{sw} c_p \cdot \frac{F_d}{3} T_{in} \\
&\quad - \rho_{sw1} c_p \cdot F_1 T_1 + Q - UA \cdot (T_1 - T_{atm}) \\
&= \rho_{sw1} c_p \cdot F_{in} T_{in} + \rho_{sw3} c_p \cdot F_{recycle} T_{out} + \rho_{sw} c_p \cdot \frac{F_d}{3} T_{in} \\
&\quad - \rho_{sw1} c_p \cdot F_1 T_1 + Q - UA \cdot (T_1 - T_{atm}) \quad (3 - 9)
\end{align*}
\]
4.3.3 Heated Tank-2 Models

4.3.3.1 Mass Balance

\[
\frac{d\dot{m}_1}{dt} = \dot{m}_1 + m_d - \dot{m}_2 \tag{3 - 12}
\]

Where: \( m = \rho V, \dot{m} = \rho F \)

\[
\rho_{sw2} \frac{dV_2}{dt} = \rho_{sw1} \cdot F_1 + \rho_{sw} \cdot \frac{F_d}{3} - \rho_{sw2} \cdot F_2 \tag{3 - 13}
\]

\[
\frac{dV_2}{dt} = \frac{1}{\rho_{sw2}} \left[ \rho_{sw1} \cdot F_1 + \rho_{sw} \cdot \frac{F_d}{3} - \rho_{sw2} \cdot F_2 \right] \tag{3 - 14}
\]

Where: \( F_1 = F_1 + \frac{F_d}{3} + F_{recycle} \tag{3 - 15} \)

\( F_2 = F_1 + \frac{F_d}{3} \tag{3 - 16} \)
DEVELOPMENT OF THE PROCESS MODEL

4.3.3.2 Component Balance

\[ \frac{dV_2 C_2}{dt} = V_2 \frac{dC_2}{dt} + C_2 \frac{dV_2}{dt} = F_1 \cdot C_1 + \frac{F_d}{3} \cdot C_d - F_2 \cdot C_2 \]  \hspace{1cm} (3 - 17)

Where: \( V_2 \) is constant

\[ \frac{dC_2}{dt} = \frac{1}{V_2} \left[ F_1 \cdot C_1 + \frac{F_d}{3} \cdot C_d - F_2 \cdot C_2 \right] \]  \hspace{1cm} (3 - 18)

4.3.3.3 Energy Balance

Rate of energy accumulation

= Rate of input of energy – Rate of output of energy
+ Rate of heat input
– heat loss to atmosphere  \hspace{1cm} (3 - 19)

\[ \rho_{sw2} C_p \frac{d}{dt} [V_2 T_2] \]

\[ = \rho_{sw1} C_p \cdot F_1 \cdot T_1 + \rho_{sw} C_p \cdot \frac{F_d}{3} \cdot T_{in} - \rho_{sw2} C_p \cdot F_2 \cdot T_2 + Q \]

– \( UA \cdot (T_2 - T_{atm}) \)  \hspace{1cm} (3 - 20)

\[ \frac{dT_2}{dt} = \frac{1}{\rho_{sw2} C_p V_2} \left[ \rho_{sw1} C_p \cdot F_1 \cdot T_1 + \rho_{sw} C_p \cdot \frac{F_d}{3} \cdot T_{in} - \rho_{sw2} C_p \cdot F_2 \cdot T_2 + Q \right. \]

– \( UA \cdot (T_2 - T_{atm}) \]  \hspace{1cm} (3 - 21)

4.3.4 Heated Tank-3 Models
DEVELOPMENT OF THE PROCESS MODEL

4.3.4.1 Mass Balance

Rate of Accumulation of mass
\[ \frac{dm}{dt} = \dot{m}_2 + m_d - m_{\text{out}} \] (3 – 22)

Where: \( m = \rho V \), \( \dot{m} = \rho F \)

\[ \rho_{sw3} \frac{dV_3}{dt} = \rho_{sw2} \cdot F_2 + \rho_{sw} \cdot \frac{F_d}{3} - \rho_{sw3} \cdot F_{\text{out}} \] (3 – 24)

\[ \frac{dV_3}{dt} = \frac{1}{\rho_{sw3}} \left[ \rho_{sw2} \cdot F_2 + \rho_{sw} \cdot \frac{F_d}{3} - \rho_{sw3} \cdot F_{\text{out}} \right] \] (3 – 25)

Where: \( V_3 = h \cdot A_3 \) (3 – 26)

\[ \frac{dh}{dt} = \frac{1}{\rho_{sw3} \cdot A_3} \left[ \rho_{sw2} \cdot F_2 + \rho_{sw} \cdot \frac{F_d}{3} - \rho_{sw3} \cdot F_3 \right] \] (3 – 27)

Where: \( F_3 = F_{\text{out}} - F_{\text{recycle}} \) (3 – 28)

4.3.4.2 Component Balance

\[ \frac{dV_3 C_{\text{out}}}{dt} = V_3 \frac{dC_{\text{out}}}{dt} + C_{\text{out}} \frac{dV_3}{dt} \]
\[ = F_2 \cdot C_2 + \frac{F_d}{3} \cdot C_d - F_{\text{out}} \cdot C_{\text{out}} \] (3 – 29)

Substitue equations (3 – 24) & (3 – 26) into equation (3 – 29)
DEVELOPMENT OF THE PROCESS MODEL

\[
V_3 \frac{dC_{out}}{dt} = F_2 \cdot C_2 + \frac{F_d}{3} \cdot C_d - F_{out} \cdot C_{out} - C_{out} \left( \rho_{sw2} \cdot F_2 + \rho_{sw} \cdot \frac{F_d}{3} \right)
- \rho_{sw3} \cdot F_{out} \quad (3 - 30)
\]

\[
h \cdot A_3 \frac{dC_{out}}{dt} = F_2 \cdot C_2 + \frac{F_d}{3} \cdot C_d - F_{out} \cdot C_{out} - C_{out} \left( \rho_{sw2} \cdot F_2 + \rho_{sw} \cdot \frac{F_d}{3} \right)
- \rho_{sw3} \cdot F_{out} \quad (3 - 31)
\]

\[
\frac{dC_{out}}{dt} = \frac{1}{h \cdot A_3} \left[ F_2 \cdot C_2 + \frac{F_d}{3} \cdot C_d - F_{out} \cdot C_{out}
- C_{out} \left( \rho_{sw2} \cdot F_2 + \rho_{sw} \cdot \frac{F_d}{3} \right) - \rho_{sw3} \cdot F_{out} \right] \quad (3 - 32)
\]

4.3.4.3 Energy Balance

\[\text{Rate of energy accumulation} = \text{Rate of input of energy} - \text{Rate of output of energy} + \text{Rate of heat input} - \text{heat loss to atmosphere} \quad (3 - 33)\]

\[
\rho_{sw3} c_p \frac{d}{dt} [V_3 T_{out}] = \rho_{sw2} c_p \cdot F_2 T_2 + \rho_{sw} c_p \cdot \frac{F_d}{3} T_{in} - \rho_{sw3} c_p \cdot F_{out} T_{out} + Q
- U A \cdot (T_{out} - T_{atm}) \quad (3 - 34)
\]

Where:
\[
\frac{dV_3 T_{out}}{dt} = V_3 \frac{dT_{out}}{dt} + T_{out} \frac{dV_3}{dt} \quad (3 - 35)
\]
\[V_3 = h \cdot A_3\]

Using equation (3 – 24) from mass balance
DEVELOPMENT OF THE PROCESS MODEL

\[ \rho_{sw3} \cdot C_p \cdot h \cdot A_3 \frac{dT_{out}}{dt} \]
\[ = \rho_{sw2} \cdot C_p \cdot F_2 \cdot T_2 + \rho_{sw} \cdot C_p \cdot \frac{F_d}{3} \cdot T_{in} - \rho_{sw3} \cdot C_p \cdot F_{out} \cdot T_{out} + Q \]
\[ - U A \cdot (T_{out} - T_{atm}) \]
\[ - T_{out} \left( \rho_{sw2} \cdot F_2 + \rho_{sw} \cdot \frac{F_d}{3} - \rho_{sw3} \cdot F_3 \right) \quad (3 - 36) \]

\[ \frac{dT_{out}}{dt} = \frac{1}{\rho_{sw3} \cdot C_p \cdot h \cdot A_3} \left[ \rho_{sw2} \cdot C_p \cdot F_2 \cdot T_2 + \rho_{sw} \cdot C_p \cdot \frac{F_d}{3} \cdot T_{in} \right. \]
\[ - \rho_{sw3} \cdot C_p \cdot F_{out} \cdot T_{out} + Q - U A \cdot (T_{out} - T_{atm}) \]
\[ - T_{out} \left( \rho_{sw2} \cdot F_2 + \rho_{sw} \cdot \frac{F_d}{3} - \rho_{sw3} \cdot F_3 \right) \]  \quad (3 - 37) \]

*Where: Q = Qs \cdot \lambda s*

Qs is steam flowrate

\( \lambda s \) is latent heat (constant)

*U can be calculated by putting equation (3 \(- 37) at steady state then re – arrange and solve for U.*

### 4.4 Degree of Freedom Analysis

In order to determine whether the system can be simulated with the given set of equations for the model, the degree of freedom analysis needs to be done to establish that there is a finite set of result [20].

The number of degree of freedom can be calculated by subtracting the total number of independent (process) equations from the total number of unknown /process variables [20].
DEVELOPMENT OF THE PROCESS MODEL

A system is only realisable i.e. a finite solution, if the number of unknown / process variables are equal to the number of independent equations. The solution obtained from such a system may not necessarily be unique when dealing with nonlinear set of equations.

In this project, the total number of system variables is 23 (Appendix F), which are listed as below:

- \( \text{Fin, Fout, Frecycle, Fd, F1, F2, F3} \)
- \( \text{Tin, T1, T2, Tout} \)
- \( \text{Q1, Q2, Q3} \)
- \( \text{Cin, C1, C2, Cout} \)
- \( \text{Cond1, Cond2, Cond3} \)
- \( \text{h3, P} \)

The numbers of dependent equations for the system are found to have been 13 (Appendix G), which means the degree of freedom is 9. As such, eight of these variables need to be identified which are classified as below:

- Manipulated variables: \( \text{Fout, Q1, Q2, Q3, Fd} \)
- Disturbances variables \( \text{Fin, Frecycle, Tin, Cin} \)

4.5 Dynamic Model Implementation in Matlab

The dynamic mathematical model of the system was implemented in MATLAB in a function file using \( xdot \) vector. The MATLAB solver ‘ode45’ in the main file calls the function file and solve it within defined time span. The initial steady state conditions
DEVELOPMENT OF THE PROCESS MODEL

(Appendix F) for the process variables $h$, $T_1$, $T_2$, $T_{out}$, $C_1$, $C_2$ and $C_{out}$ were given as initial states to ‘ode45’. Each process variable is then put in an array and plotted.

4.5.1 Steady State Analysis

Using MATLAB inbuilt function `fsolve`, the steady state values for the process variables $h$, $T_1$, $T_2$, $T_{out}$, $C_1$, $C_2$ and $C_{out}$ were computed using the model ODE. The steady state computation using `fsolve` on MATLAB. The following shows the MATLAB outputs on Command Windows:

![MATLAB Outputs](image)

Figure 22: fsolve performance results using ODE on Command Windows.
DEVELOPMENT OF THE PROCESS MODEL

4.5.2 Dynamic Response Analysis

Using the steady state solutions obtained from Part 1, a dynamic response of the three heated tanks had been simulated and it was done by using MATLAB inbuilt function ode45. To do so, two MATLAB scripts were written; one m.file encoded with the function ode45, while the other m.file contains a function that holds the model ODEs shown in previous section.

The MATLAB scripts used for generating the dynamic responses can be referred in Appendix H. After running the program, the Figure 23 in the next page shows the system steady state run performance using the solution obtained from Part 1, including a plot of the given sets of steady state inputs.

Once steady state run had accomplished, the dynamic nature of the system was tested via disturbance change as follows;

- Feed flow rate (±5 %),
- Recycle flow rate (±5 L/min),
- Feed temperature (±5 Degree C) and
- Feed concentration (±0.005 Kg/L) individually.
Figure 23: Steady State Run.
After a positive step change of 5% on feed flow rate, $F_{in}$ which was initially 5.964 L/min, the following shows the system response parameters such as $T_1, T_2, T_{out}, C_1, C_2,$ and $h$, along with its disturbance variables.

Figure 24: Dynamic response to positive Fin step change.
Figure 25: Dynamic response to positive Freecyle step change.
Figure 26: Dynamic response to negative Fre cycle step change.
DEVELOPMENT OF THE PROCESS MODEL

According to Figures 24, 25 and 26, step change in feed flow-rate or recycle flow-rate has produced changes in all of the process variables, which means that the model is working fine.

4.6 Summary

In this chapter, the process is modelled by nonlinear set of equations. Implementation of the process model for dynamic system has been done in MATLAB. In addition, the dynamic nature of the process has been investigated by introducing some of the disturbance changes.

Therefore, development of the process models, implementation of the dynamic model and studying the dynamic nature of the system have been completed successfully in this chapter.
CHAPTER 5

PLANTWIDE CONTROL

5.1. Overview

5.2. Plant-wide Control System Design Procedures

5.3. Heated Tanks Unit Plant

5.4. Process Parameters Specifications
5.1 Overview

In this chapter, a hierarchical design procedure will be described that can be used to develop multiloop and multivariable measurement and control strategies for plantwide control systems. The procedure assists the engineer in determining how to choose the best controlled, manipulated, and measured variables in the plant, when to use advanced control techniques such as GMC. The proposed design procedure is based on the hierarchy of process control activities.

The goal is a plantwide control system design that is no more complicated or expensive than necessary and that, when built, can be operated easily by typical plant operators. Finally, the only definitive way of validating a selected plantwide control system design is by plant tests and by the operating plant's performance.

5.2 Procedures for the Design of Plant-wide Control System

The design of a plant-wide control system consists of three major steps:

1. The **overall specifications for the plant** and its control system are stated.

2. The control system structure is developed. This step includes selecting **controlled, measured, and manipulated variables**; choosing multiloop or multivariable control; deciding how to control production rate, product quality, and inventories; and handling operating constraints. Decomposition of the plantwide control problem into smaller problems for the purpose of analysis may also be employed here [22].
3. Design is followed by a detailed specification of all instrumentation/hardware and software, cost estimation, evaluation of alternatives, and the ordering and installation of equipment [22].

4. Following design and implementation of controllers and plant tests which will be discussed in the next two chapters.

Figure 27: Hierarchical Design Procedures [17].
5.3 Heated Tanks Unit Plant

The principles from the previous section can be applied now to the plant of the three heated tanks.

5.3.1 Heated Tanks Process Design Overall Specifications

5.3.1.1 Plant Process Description

As explained earlier in section 4.2 that the heated tanks process design have the following specifications:

- Three heated tanks connected in series.
- One dye tank connected to each of the heated tanks.
- Two different feed streams:
  - Non-concentrated feed stream (Pure water) coming from the needle tank and feeds directly into the first heated tank.
  - Concentrated feed stream (Salty water) coming from the dye tank and feeds into all of the three heated tanks.
- The three heated tanks heat up the mixed feed and sending it as a product through the outlet of the last heated tank.
- From the third heated tank a part of the concentrated product is drawn and rest of the product is recycled back into the first heated tank as can be seen in Figure 15.
Figure 28: Schematic Diagram of Heated Tanks Process Design (Author).
5.3.1.2 Assumptions

The following are the assumptions that have been made to heated tank process:

- Mixing is perfect in each of the heated tank.
- Temperature is constant in each of the three heated tanks.
- Tank 1 and Tank 2 overflow at all times.
- Stream Fin is pure water (non-concentrated fluid).
- Stream Fd is salty water (Concentrated fluid).
- The delay in the piping is negligible; materials move from the dye tank to all of the heated tanks in zero time.

5.3.1.3 Density and Concentration Relationship

Density measures the amount of mass per unit of volume in a substance. Concentration describes the amount of a substance dissolved in another substance. Changing the concentration of a solution changes the density of the solution.

5.3.1.3.1 Concentration

The concentration in a solution \( C_{sw} \) = \( \frac{\text{Mass of solute (salt)}}{\text{Volume of solution (water)}} \) (5 - 1)

\[
C_{sw} = \frac{325 \text{ [g]}}{18 \text{ [L]}} = 18.056 \frac{g}{L}
\]

\[
C_{sw} = \frac{0.325 \text{ [Kg]}}{18 \text{ [L]}} = 0.018056 \frac{Kg}{L}
\]

\[
C_{sw} = \frac{0.325 \text{ [Kg]}}{0.018 \text{ [m}^3\text{]}\}} = 18.056 \frac{Kg}{m^3}
\]

The unit of \( \frac{Kg}{L} \) has been decided to use.
5.3.1.3.2 Density

Density \( (\rho_{sw}) \) = \( \frac{\text{Mass of a substance \ (Total mass)}}{\text{Volume of the substance \ (Total Volume)}} \) (5 - 2)

\[
= \frac{0.325 \ [\text{Kg}] + 18 \ [\text{Kg}]}{18 \ [\text{L}]}
\]

\[
= \frac{18.325 \ [\text{Kg}]}{18 \ [\text{L}]}
\]

\( \rho_{sw} = 1.01806 \left[ \frac{\text{Kg}}{\text{L}} \right] \)

Known: \( \rho_{pw} = 1.0 \left[ \frac{\text{Kg}}{\text{L}} \right] \)

5.3.1.3.3 Salt Water Density \( (\rho_{sw}) \) and Salt Water Concentration \( (C_{sw}) \) Relationship

Mass Fraction, \( x_{salt} = \frac{\text{Mass of solute \ (salt)}}{\text{Mass of solution \ (water)}} = \frac{m_{salt}}{m_{solution}} \) (5 - 3)

\( V_{solution} = V_{water} + V_{salt} \) (5 - 4)

\( V_{solution} = \frac{m_{solution}}{\rho_{sw}} \) (5 - 5)

\( V_{water} = \frac{m_{water}}{\rho_{water}} \) (5 - 6)

\( V_{salt} = \frac{m_{salt}}{\rho_{salt}} \) (5 - 7)
Substitute equations (5 – 5), (5 – 6) and (5 – 7) in equation (5 – 4):

\[
\frac{m_{\text{solution}}}{\rho_{\text{sw}}} = \frac{m_{\text{water}}}{\rho_{\text{water}}} + \frac{m_{\text{salt}}}{\rho_{\text{salt}}} \quad (5 – 8)
\]

where, \( m_{\text{water}} = m_{\text{solution}} – m_{\text{salt}} \) \quad (5 – 9)

Equation (5 – 8) becomes:

\[
\frac{m_{\text{solution}}}{\rho_{\text{sw}}} = \frac{m_{\text{solution}}}{\rho_{\text{water}}} - \frac{m_{\text{salt}}}{\rho_{\text{water}}} + \frac{m_{\text{salt}}}{\rho_{\text{salt}}} \quad (5 – 10)
\]

Divide both sides of equation (5 – 10) by \( m_{\text{solution}} \):

\[
\frac{1}{\rho_{\text{sw}}} = \frac{1}{\rho_{\text{water}}} - \frac{m_{\text{salt}}}{m_{\text{solution}} \cdot \rho_{\text{water}}} + \frac{m_{\text{salt}}}{m_{\text{solution}} \cdot \rho_{\text{salt}}} \quad (5 – 11)
\]

Equation (5 – 11) becomes:

\[
\frac{1}{\rho_{\text{sw}}} = \frac{1}{\rho_{\text{water}}} + \frac{m_{\text{salt}}}{m_{\text{solution}}} \left( \frac{1}{\rho_{\text{salt}}} - \frac{1}{\rho_{\text{water}}} \right) \quad (5 – 12)
\]

where, \( x_{\text{salt}} = \frac{m_{\text{salt}}}{m_{\text{solution}}} \), Equation (5 – 12) becomes:

\[
\frac{1}{\rho_{\text{sw}}} = \frac{1}{\rho_{\text{water}}} + x_{\text{salt}} \left( \frac{1}{\rho_{\text{salt}}} - \frac{1}{\rho_{\text{water}}} \right) \quad (5 – 13)
\]

where, \( x_{\text{salt}} = \frac{c_{\text{sw}}}{\rho_{\text{sw}}} \) \quad (5 – 14)

Equation (5 – 13) becomes:

\[
\frac{1}{\rho_{\text{sw}}} = \frac{1}{\rho_{\text{water}}} + \frac{c_{\text{sw}}}{\rho_{\text{sw}}} \left( \frac{1}{\rho_{\text{salt}}} - \frac{1}{\rho_{\text{water}}} \right) \quad (5 – 15)
\]
Multiply both sides of equation (5 – 10) with $\rho_{sw}$:

$$1 = \frac{\rho_{sw}}{\rho_{water}} + c_{sw} \left( \frac{1}{\rho_{salt}} - \frac{1}{\rho_{water}} \right) \quad (5 – 16)$$

Rearrange and solve for $\rho_{sw}$:

$$\rho_{sw} = \rho_{water} \left[ 1 - c_{sw} \left( \frac{1}{\rho_{salt}} - \frac{1}{\rho_{water}} \right) \right] \quad (5 – 17)$$

So, the salt water density ($\rho_{sw}$) for the three heated tanks related to the concentrations ($c_{sw}$) are as follows:

1. Salt water density ($\rho_{sw1}$) of Heated Tank – 1

$$\rho_{sw1} = \rho_{water} \left[ 1 - c_{sw1} \left( \frac{1}{\rho_{salt}} - \frac{1}{\rho_{water}} \right) \right] \quad (5 – 19)$$

2. Salt water density ($\rho_{sw2}$) of Heated Tank – 2

$$\rho_{sw2} = \rho_{water} \left[ 1 - c_{sw2} \left( \frac{1}{\rho_{salt}} - \frac{1}{\rho_{water}} \right) \right] \quad (5 – 20)$$

3. Salt water density ($\rho_{sw3}$) of Heated Tank – 3

$$\rho_{sw3} = \rho_{water} \left[ 1 - c_{sw3} \left( \frac{1}{\rho_{salt}} - \frac{1}{\rho_{water}} \right) \right] \quad (5 – 21)$$

5.3.1.4 Conductivity and Concentration Relationship

As mentioned earlier that the previous student (David Pol) has designed PI controller to control the conductivity in the three heated tanks. In this project, concentration in the three heated tanks will be controlled. However, conductivity in each heated tank
PLANTWIDE CONTROL SYSTEM DESIGN

will be calculated based on the measured concentrations using the following equation.

\[ TDS = SC \times f \]  \hspace{1cm} (5 - 22)

where,

\[ TDS = \text{Total Dissolved Solids in} \left[ \frac{mg}{L} \right] \]

\[ SC = \text{Specific Conductance in} \left[ \frac{\mu S}{cm} \right] \]

\[ f = \text{Conversion Factor} = 0.65 \]

Therefore, Conductivity can be calculated through the following equation:

\[ SC = \frac{[TDS \times 1000]}{f} \]  \hspace{1cm} (5 - 23)

Calculated conductivity for each of the three heated tanks obtained from the measured concentrations can be found in Appendix L.

5.3.2 Detailed Specification of all instrumentation/hardware and software

Before the pilot plant can be controlled adequately, the associated pumps and control valves must be calibrated, allowing the operator to know the correct rate of fluid flow. In order to achieve this, a relationship between the valve opening percentage and the corresponding fluid flow rate need to be determined. To achieve this, two differing methods of calibration are employed: Manual and Data Calibration.

As some of the pipes feature both a pump and a control valve pairing, and others feature only a pump, the decision was made for consistency to set all control valves
to be open to 100% and measure flow rate against the operating point of the corresponding pump. This removes them from the calculation and avoids any unwanted disturbances when calibrating the pumps.

The unit operators which must be calibrated for this project are: NTP_561, PP_681, FCV_622, FCV_642 and FCV_662.

5.3.2.1 Manual Calibration

The first and more preferable method is to perform the calibrations manually. Manual calibration involves the physical collection of fluid in a measuring cylinder over a fixed time interval for a given pump operating point and repeated over the range of operation of the pump. To ensure reliable and repeatable measurements can be achieved, several steps are taken to eliminate possible sources of error, including:

- Allowing sufficient time after adjusting the pump operating point before taking a measurement such that fluid flow rate has reached steady state.
- Adjusting the pump operating point over its entire range in both positive and negative increments to construct the hysteresis plot.
- Repeating the test a sufficient number of times and calculating an average value to reduce manual timing and measurement errors.

Calibration of FCV_642, the steam control valve that determines the flow rate of steam in to Heated Tank 2 was completed using this method. The fixed time interval was decided to be 10 seconds and three measurements were taken at each interval,
averaged and converted from ml to litres per minute. The result of this data collection can be seen in Table 6.

<table>
<thead>
<tr>
<th>% Open</th>
<th>V(L)</th>
<th>F(L/min)</th>
</tr>
</thead>
<tbody>
<tr>
<td>20%</td>
<td>0.41</td>
<td>0.205</td>
</tr>
<tr>
<td>30%</td>
<td>0.55</td>
<td>0.275</td>
</tr>
<tr>
<td>40%</td>
<td>0.68</td>
<td>0.34</td>
</tr>
<tr>
<td>50%</td>
<td>0.81</td>
<td>0.405</td>
</tr>
<tr>
<td>60%</td>
<td>0.95</td>
<td>0.475</td>
</tr>
<tr>
<td>70%</td>
<td>1.125</td>
<td>0.5625</td>
</tr>
<tr>
<td>80%</td>
<td>1.325</td>
<td>0.636</td>
</tr>
<tr>
<td>90%</td>
<td>1.46</td>
<td>0.73</td>
</tr>
<tr>
<td>100%</td>
<td>1.55</td>
<td>0.775</td>
</tr>
</tbody>
</table>

Using Excel, this data is fit to a plot and a trend line can be established to relate valve position to flow rate in litres per minute. The plot is shown in Figure 29.
The relationship between valve position and flow rate in litres per minute of steam flow-rate to the heated tank 2 is given by equation (5 – 23) where \( y \) is flow rate and \( x \) is valve position.

\[
y = 0.7324x + 0.0498 
\]  
\( (5 – 24) \)

Calibration of PP_681, the product pump that determines the flow rate of production in to Heated Tank 3 (CSTR3) was completed using this method. The result of this data collection is detailed in Table 7.

<table>
<thead>
<tr>
<th>%Open</th>
<th>V(L)</th>
<th>F (L/min)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>10</td>
<td>2.0448</td>
<td>4.0896</td>
</tr>
<tr>
<td>20</td>
<td>1.4304</td>
<td>5.7216</td>
</tr>
<tr>
<td>30</td>
<td>2.1456</td>
<td>8.5824</td>
</tr>
<tr>
<td>40</td>
<td>2.6256</td>
<td>10.5024</td>
</tr>
<tr>
<td>50</td>
<td>3.12</td>
<td>12.48</td>
</tr>
<tr>
<td>60</td>
<td>3.456</td>
<td>13.824</td>
</tr>
</tbody>
</table>

Using Excel, this data is fit to a plot and a trend line can be established to relate valve position to flow rate in litres per minute. The plot is shown in Figure 30.
The relationship between pump opening and flow rate in litres per minute of product pump to the heated tank is given by equation (5 – 24) where $y$ is flow rate and $x$ is valve position.

$$ y = 0.2061x + 2.354 \quad (5 – 25) $$

### 5.3.2.1 Data Calibration

The second method relies upon the use of Experion to read information about the level of the tank (can be either volume drop of preceding tank, or volume increase of proceeding tank) and recording the rate at which this changes over a fixed time interval. This rate of change can then be compared to the physical dimensions of the tank which (measured manually) to provide the actual fluid flow rate. This method is less accurate; however as it is subject to several nuances that must be factored to ensure reliability of the data.

The rest of the calibration data can be found in Appendix B.
5.4.2 Process Parameters Specifications

Table 8 lists all the process and manipulated variables.

**Table 8: Process and Manipulated Variables**

<table>
<thead>
<tr>
<th>S. No.</th>
<th>Variable</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>$F_{in}$</td>
<td>Feed flow rate</td>
</tr>
<tr>
<td>2</td>
<td>$F_{out}$</td>
<td>Product flow rate</td>
</tr>
<tr>
<td>3</td>
<td>$F_{recycle}$</td>
<td>Circulating (Recycle) flow rate</td>
</tr>
<tr>
<td>4</td>
<td>$D_{d}$</td>
<td>Dye Tank Flowrate</td>
</tr>
<tr>
<td>5</td>
<td>$F_1$</td>
<td>$F_{in} + F_{recycle} + D_{d}$</td>
</tr>
<tr>
<td>6</td>
<td>$F_2$</td>
<td>$F_1 + D_{d}$</td>
</tr>
<tr>
<td>7</td>
<td>$F_3$</td>
<td>$F_{out} + F_{recycle}$</td>
</tr>
<tr>
<td>8</td>
<td>$T_{in}$</td>
<td>Feed temperature</td>
</tr>
<tr>
<td>9</td>
<td>$T1$</td>
<td>Heated Tank-1 temperature</td>
</tr>
<tr>
<td>10</td>
<td>$T2$</td>
<td>Heated Tank-2 temperature</td>
</tr>
<tr>
<td>11</td>
<td>$T3$</td>
<td>Heated Tank-3 temperature</td>
</tr>
<tr>
<td>12</td>
<td>$Q1$</td>
<td>Heated Tank-1 Steam Flowrate</td>
</tr>
<tr>
<td>13</td>
<td>$Q2$</td>
<td>Heated Tank-2 Steam Flowrate</td>
</tr>
<tr>
<td>14</td>
<td>$Q3$</td>
<td>Heated Tank-3 Steam Flowrate</td>
</tr>
<tr>
<td>15</td>
<td>$h3$</td>
<td>Heated Tank-3 water level</td>
</tr>
<tr>
<td>16</td>
<td>$P$</td>
<td>Steam Operating pressure</td>
</tr>
<tr>
<td>17</td>
<td>$C_{in}$</td>
<td>Feed Concentration flow rate</td>
</tr>
<tr>
<td>18</td>
<td>$C_1$</td>
<td>Heated Tank-1 Concentration</td>
</tr>
<tr>
<td>19</td>
<td>$C_2$</td>
<td>Heated Tank-2 Concentration</td>
</tr>
<tr>
<td>20</td>
<td>$C_3$</td>
<td>Heated Tank-3 Concentration</td>
</tr>
<tr>
<td>21</td>
<td>$C_{ond1}$</td>
<td>Heated Tank-1 Conductivity</td>
</tr>
<tr>
<td>22</td>
<td>$C_{ond2}$</td>
<td>Heated Tank-2 Conductivity</td>
</tr>
<tr>
<td>23</td>
<td>$C_{ond3}$</td>
<td>Heated Tank-3 Conductivity</td>
</tr>
</tbody>
</table>
PLANTWIDE CONTROL SYSTEM DESIGN

5.4.2.1 Plant Constraints

The plant design has certain limitations imposed on various flows, temperatures and energy input; it is vital all operation of the plant follows these restrictions. The following lists the details of the plant constraints and its inequality expression that needed to be implemented in this plant as well as the range of inputs that needed to be satisfied throughout the process of controller design.

The following summarise the list of constraints a

1. Feed temperature limit, $T \leq 26$. 
2. Non-concentrated Feed flow rate limits, $5 \leq F_{in} \leq 9.24$ 
3. Concentrated Feed flow rate limits, $0.335 \leq F_d \leq 0.525$ 
4. Inlet concentration, $C_{in} \leq 0.018056$. 
5. Product temperature limits, $58 \leq T_{out} \leq 62$. 
6. Product concentration, $C_{product} \leq 0.001$. 
CHAPTER 6

CONVENTIONAL CONTROL (PI) DESIGN AND IMPLEMENTATION

6.1. Overview
6.2. Implementation of PI Controllers in MATLAB
6.3. PI Controller Tuning
6.4. PI Performance analysis
6.5. Summary
6.1 Overview

The chapter includes the implementation of the dynamic model, loop pairing for the manipulated variables and controlled variables, implementing and tuning PI controllers for each process variable and performance evaluation of the controllers by introducing disturbances and set point change in process variables.

6.2 Implementation of PI controller in Matlab

When the loop is closed around the system the set point is compared to the process variable, resulting in an error. This error is sent to the controller which then decides the appropriate action the manipulated variable should take. This manipulated variable then alters the system and the corresponding process output is fed back and compared to the set point again.

![Conventional PI controller block diagram.](image)

**Figure 31: Conventional PI controller block diagram.**

6.2.1 Manipulated Variable (MV) Calculation

With the models now implemented, and dynamic responses tested, controllers must be fitted to control the seven dynamic process variables. For the seven feedback loops PI controllers are implemented; given by the follow equation:

\[
MV = MV_{ss} + K_c \left( \varepsilon + \frac{\int \varepsilon \, dt}{\tau_i} \right)
\]  

(6 - 1)
6.2.2 Control Loop Pairing (Loop Selections)

The control loop pairings that were selected for the automatic control of the heated tanks in the pilot plant can be seen in Table 9. These particular control loops leave the needle tank pump (NTP_561) and recycle flowrate as free variables that can be used as disturbance variables by manually adjusting at any point to change the flow out of the pumps.

Table 9: Control Loop Pairing.

<table>
<thead>
<tr>
<th>Process Variable (PV)</th>
<th>Manipulated Variable (MV)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Heated Tank-1 Level (h)</td>
<td>Product Pump (Fout)</td>
</tr>
<tr>
<td>Heated Tank-1 Temperature (T1)</td>
<td>Steam Valve-1 (Q1)</td>
</tr>
<tr>
<td>Heated Tank-2 Temperature (T2)</td>
<td>Steam Valve-2 (Q2)</td>
</tr>
<tr>
<td>Heated Tank-3 Temperature (Tout)</td>
<td>Steam Valve-2 (Q3)</td>
</tr>
<tr>
<td>Heated Tank-1 Concentration (C1)</td>
<td>Dye Tank Pump (Fd)</td>
</tr>
<tr>
<td>Heated Tank-2 Concentration (C2)</td>
<td>Dye Tank Pump (Fd)</td>
</tr>
<tr>
<td>Heated Tank-3 Concentration (Cout)</td>
<td>Dye Tank Pump (Fd)</td>
</tr>
</tbody>
</table>

6.2.3 Error Calculation

Where, the error is defined by $\varepsilon = SP - PV$, the manipulated variable, MV, is defined in Table 9 and the controller parameters, $K_c$ and $\tau_i$, are to be selected and tuned. The following control variable assignments were done by the following guidelines, as Perkins stated:

- “Guideline 1: Always select state variables representing inventories which are not self-regulatory” [21] [22];
Guideline 2: Select state variables which, although self-regulatory, may exceed equipment or process constraints" [23] [24];

Guideline 3: Select state variables which, although self-regulatory, may seriously interact with other inventories" [25] [26] [27].

Finally to implement these into a MATLAB script the error, \( \epsilon \), is rearranged into state space. Given there are three feedback loops, this intuitively results in three set points, SP, and errors. These feedback errors are given by:

\[
\begin{align*}
\epsilon_h &= SP_h - h \\
\epsilon_{T1} &= SP_{T1} - T_1 \\
\epsilon_{T2} &= SP_{T2} - T_2 \\
\epsilon_{Tout} &= SP_{Tout} - T_{out} \\
\epsilon_{C1} &= SP_{C1} - C_1 \\
\epsilon_{C2} &= SP_{C2} - C_2 \\
\epsilon_{Cout} &= SP_{Cout} - C_{out}
\end{align*}
\]

This becomes:

\[
\begin{align*}
\text{xdot(8)} &= \text{sph} - h \\
\text{xdot(9)} &= \text{spT1} - T_1 \\
\text{xdot(10)} &= \text{spT2} - T_2 \\
\text{xdot(11)} &= \text{spTout} - Tout \\
\text{xdot(12)} &= \text{spC1} - C_1 \\
\text{xdot(13)} &= \text{spC2} - C_2 \\
\text{xdot(14)} &= \text{spCout} - Cout
\end{align*}
\]
Note, however, the process variables have been previously defined as $x(1), x(2), x(3), x(4), x(5), x(6)$ and $x(7)$.

\[
\begin{align*}
&xdot(8) = sph - x(1); \\
&xdot(9) = spT1 - x(2); \\
&xdot(10) = spT2 - x(3); \\
&xdot(11) = spTout - x(4); \\
&xdot(12) = spC1 - x(5); \\
&xdot(13) = spC2 - x(6); \\
&xdot(14) = spCout - x(7);
\end{align*}
\]

6.2.4 Controller Equation

Now with the errors defined it is possible to rewrite the controllers as:

\[
\begin{align*}
F_{out} &= F_{out_{ss}} + K_h \left( SP_h - x(1) + \frac{x(8)}{\tau_{ih}} \right) \\
Q_1 &= Q_{1_{ss}} + K_{T1} \left( SP_{T1} - x(2) + \frac{x(9)}{\tau_{yr1}} \right) \\
Q_2 &= Q_{2_{ss}} + K_{T2} \left( SP_{T2} - x(3) + \frac{x(10)}{\tau_{yr2}} \right) \\
Q_3 &= Q_{3_{ss}} + K_{Tout} \left( SP_{Tout} - x(4) + \frac{x(11)}{\tau_{yrout}} \right) \\
F_d &= F_{d_{ss}} + K_{C1} \left( SP_{C1} - x(5) + \frac{x(12)}{\tau_{iC1}} \right) \\
F_d &= F_{d_{ss}} + K_{C2} \left( SP_{C2} - x(6) + \frac{x(13)}{\tau_{iC2}} \right) \\
F_d &= F_{d_{ss}} + K_{Cout} \left( SP_{Cout} - x(7) + \frac{x(14)}{\tau_{iCout}} \right)
\end{align*}
\]
6.2.5 Matlab Function

These controllers are now ready to be implemented into a *MATLAB* function. To solve the dynamic problem *MATLAB*’s inbuilt one step solver, *ode45*, is used; this utilizes an explicit Runge-Kutta (4,5) formula, the Dormand-Prince pair.

When employing the solver in the main *MATLAB* script the timespan, initial conditions for the differential equations and solver options must be defined. If \(x_0\) defines the differential equations initial conditions, \(t_0\) the initial solver time and \(t_f\) the final solver time the function is coded as follows:

```matlab
x0 = [hs T1s T2s Touts C1s C2s Couts 0 0 0 0 0 0 0];
t0 = 0;
tf = 50;
span = [t0 tf];
options = [];
[t, x] = ode45('FunctionPilotPlant_PI', span, x0, options);

h = x(:,1);
T1 = x(:,2);
T2 = x(:,3);
Tout = x(:,4);
C1 = x(:,5);
C2 = x(:,6);
Cout = x(:,7);
Ieh = x(:,8);
IeT1 = x(:,9);
IeT2 = x(:,10);
IeTout = x(:,11);
IeC1 = x(:,12);
IeC2 = x(:,13);
IeCout = x(:,14);
```
The last step pulls the solutions for the process variables and the integrals of the errors from the x array allowing future plotting and performance management. Now the controllers have been constructed in MATLAB, their parameters have to be determined.

6.3 PI Controller Tuning

The tuning methods used for tuning the PI controller are as follows:

- Relay Controller Tuning
- First-Order Time-Delay (FOTD) approximate Model
  - System Identification tool in MATLAB
  - Double Pulse Method (de la Barra, Jin, Kim, & Mossberg, 2008)
- ISE tuning

These values were then used to calculate the controller parameters using Ziegler-Nichols tuning guidelines [10].

6.3.1 Relay Controller Tuning

An attempt was made to use the relay tuning to achieve oscillations for the process variables. The method was dropped after it was noted that the oscillations were not good enough and the process is too time consuming.

6.3.2 First-Order Time-Delay (FOTD) approximate Model

6.3.2.1 System Identification Tool in the MATLAB

The system identification tool in the MATLAB can save a lot of time and effort if a system’s approximate model (FOTD, second order etc.) is required.
PI CONTROLLER DESIGN AND IMPLEMENTATION

In MATLAB, the following steps were carried out:

1. A step change was given to all of the process variables.
2. The process data for all of the process variables was then imported into the system identification tool one by one.
3. A FOTD for all of the process variables were then estimated by the Identification tool.

The FOTDs obtained from the tool were a very good match for some of the process variables. The results from the MATLAB system identification tool are shown in Figure 32 and Figure 33 for T1 and T2 respectively.

![CSTR1 Temperature](image1)

**Figure 32: T1 Matlab System Identification Result.**
6.3.2.2 Double Pulse Method

After identifying the problems with the relay controller, the approach for controller designing was changed. The next idea was to use different approach to get a FOTD for \( \text{Tout} \) and \( \text{Cout} \).

The method selected was ‘Identifications of first order time delay systems using two different pulse inputs’, which is an article presented in the seventeenth world congress of ‘The International Federation of Automatic Control’ in Seoul, Korea, July 6-11, 2008 [28]. The advantage of using pulses is that the process input and output will return to the initial state, thus minimizing the experiment impact on the overall process [28].
PI CONTROLLER DESIGN AND IMPLEMENTATION

In this method, the process is given a couple of pulses and the transients in the process are recorded as seen in Figure 34: Double Pulse Method for DC gain, time constant, and time delay can be calculated.

From the response, the first order system gain, time constant and time delay can be obtained using the predefined equations (6 – 2), (6 – 3), and (6 – 4) respectively [28].

![Double Pulse Method](image)

**Figure 34**: Double Pulse Method [28].

\[
T = \frac{\Delta}{\ln(y_{1a}) - \ln(y_{1b})} \quad (6 - 2)
\]

\[
K = \frac{y_{1a}}{A \left[ 1 - \left( \frac{y_{1b}}{y_{1a}} \right)^{\frac{D}{\Delta}} \right]} \quad (6 - 3)
\]

\[
L = t_a - D \quad (6 - 4)
\]
6.3.2.2.1 Tout and Cout Loop Approximation

Figure 35 and Figure 36 show the following for Tout and Cout, respectively:

- Manipulated and the Process variable response
- Values associated with the double pulse method ($y_{1a}$, $y_{1b}$, $y_{1c}$, $D$, $\delta$, $A$ and $ta$)
- FOTD model parameters ($\tau$, $K$ and $L$)
- PI controller parameters ($K$ and $\tau$)

![Figure 35: Double Pulse Method (Tout Approximate model and Controller Parameters)](image)
6.3.2.2.2 Closed Loop Response using Double Pulse Method

The controller parameters from the double pulse approximate method were holding up against set point and disturbance changes.

6.3.2.3 ISE tuning

A further attempt was made towards achieving a more tuned controller by optimising the controller parameters. ISE is a good measure for quantitating sum of the errors. Therefore, the objective function of optimisation would be minimising the integral of squared errors.
The ISE tuning measures the square of the error and integrates that error over time. Therefore, error is minimized by reducing the area under the curve of response. This minimization results in a faster and more robust response.

‘Isqnonlin’ solver in MATLAB was used to solve the optimisation. The two algorithms ‘Levenberg-Marquardt’ and ‘trust-region-reflective’ were used since they do not require any upper bound, lower bound or constraints.

The attempt was unsuccessful as the ISE error was not reduced. From our wisdom, we concluded that, since the error is too small, it may have been difficult for the algorithm to reduce it any further. The tuning method can be done again by scaling the error up and running the optimisation again, to test the hypothesis. However, this idea came at a later stage of the project when most of the analyses were complete.

### 6.3.3 PI Controller Tuning Results

Using Ziegler Nichols Tuning Rules [10] as shown in Table 10, Tuning results for PI controller can be seen in Table 11.

<table>
<thead>
<tr>
<th>Controller Type</th>
<th>$K_c$</th>
<th>$\tau_i$</th>
<th>$\tau_d$</th>
</tr>
</thead>
<tbody>
<tr>
<td>P</td>
<td>$0.5K_{cu}$</td>
<td>-</td>
<td>-</td>
</tr>
<tr>
<td>PI</td>
<td>$0.45K_{cu}$</td>
<td>$P_u/1.2$</td>
<td>-</td>
</tr>
<tr>
<td>PID</td>
<td>$0.6K_{cu}$</td>
<td>$P_u/2$</td>
<td>$P_u/8$</td>
</tr>
</tbody>
</table>
Table 11: PI Controller Tuning Results.

<table>
<thead>
<tr>
<th></th>
<th>PV</th>
<th>Kc</th>
<th>Tau</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Heated Tank 1</strong></td>
<td>T1</td>
<td>10.15</td>
<td>2.992364</td>
</tr>
<tr>
<td></td>
<td>C1</td>
<td>6.147</td>
<td>4.76</td>
</tr>
<tr>
<td><strong>Heated Tank 2</strong></td>
<td>T2</td>
<td>8.761347</td>
<td>3.162297</td>
</tr>
<tr>
<td></td>
<td>C2</td>
<td>8.051666</td>
<td>0.016466</td>
</tr>
<tr>
<td><strong>Heated Tank 3</strong></td>
<td>Tout</td>
<td>9.29</td>
<td>3.3333</td>
</tr>
<tr>
<td></td>
<td>Cout</td>
<td>7.36147</td>
<td>5.216666</td>
</tr>
<tr>
<td></td>
<td>h</td>
<td>-0.007733</td>
<td>30.64808</td>
</tr>
</tbody>
</table>

### 6.4 PI performance analysis

Two different techniques have been used in order to check the performance of PI controllers, which are as follow:

1. Set-point Tracking
2. Disturbance Rejection

#### 6.4.1 Set-point Change

The controllers were tested against a set-point change as follows:

- T1 ±5 Degree C
- T2 ±5 Degree C
- Tout ±5 Degree C
- C1 ±0.005 Kg/L
- C2 ±0.005 Kg/L
- Cout ±0.005 Kg/L
- h ±5%
1st Test: T1 +5 Degree C

The response of the system process variable (T1) and manipulated variable (Q1) is shown in Figure 37.

As can be seen in Figure 37 that a step up change of size (5), (from 34.45 C to 39.45 C), has been given in process variable of heated tank 1 (T1) which labelled by the red colour. Consequently, the manipulated variable (Q1), (labelled by the green colour) started to opened directly in order to track the desired set point.
2\textsuperscript{nd} Test: T2 – 5 Degree C

The response of the system process variable (T2) and manipulated variable (Q2) is shown in Figure 38.

As can be seen in Figure 38 that a step up change of size (5), (from 47.75 C to 42.75 C), has been given in process variable of heated tank 2 (T2) which labelled by the red colour. Consequently, the manipulated variable (Q2), (labelled by the green colour) started to close directly in order to track the desired set point.
3rd Test: Cout –0.005 Kg/L

The response of the system process variable (Cout) and manipulated variable (Fd) is shown in Figure 39.

As can be seen in Figure 39 that a step up change of size (-0.005 Kg/L), has been given in process variable of heated tank 3 (Cout) which labelled by the red colour. Consequently, the manipulated variable (Fd), (labelled by the green colour) started to close directly in order to track the desired set point.

As a result of the first technique for checking the performance of PI controller, all setpoint changes have been successfully tracked by all of the manipulated variables and each controlled variable attains its steady state.

All other responses of setpoint changes can be found in Appendix J-1.
6.4.2 Disturbance Analysis

The controllers were tested against a disturbance change as follows:

- Fin ±5%
- Frecycle ±0.5L/min
- Tin ±5 Degree C
- Cin ±5%

1st Test: Tin +5 Degree C

Figure 40: Tin Disturbance Change +5 Degree C  PI
As can be seen in Figure 40 that a step up change of size (+5 degrees C), (from 26 degrees C to 31 degrees C), has been given in the disturbance variable Tin at the suction of CSTRs bank system. Therefore, all of three process variables T1, T2 and Tout (labelled by the red colour) have been influenced and went high. Consequently, all of the relative manipulated variables Q1, Q2 and Q3 (labelled by the green colour) started to close automatically in order to reject the disturbance and keep the process variable at the desired set point.

2nd Test: Tin – 5 Degree C

Figure 41: Tin Disturbance Change -5 Degree C  PI
As can be seen in Figure 41 that a step up change of size (-5 degrees C), (from 26 degrees C to 21 degrees C), has been given in the disturbance variable \( T_{in} \) at the suction of CSTRs bank system. Therefore, all of three process variables \( T_1, T_2 \) and \( T_{out} \) (labelled by the red colour) have been influenced and went low. Consequently, all of the relative manipulated variables \( Q_1, Q_2 \) and \( Q_3 \) (labelled by the green colour) started to open automatically in order to reject the disturbance and keep the process variable at the desired set point.

As a result of the second technique for checking the performance of PI controller, all disturbance changes have been successfully rejected by all of the manipulated variables and each controlled variable attains its steady state with some errors.

All other responses of disturbance changes can be found in Appendix J-2.

### 6.5 Summary

This chapter has presented design, implementation and tuning of PI controllers for each process variable. Control performance for all designed PI controllers has been evaluated by introducing disturbances and set point change in process variables. All set point changes have been successfully tracked by all of the manipulated variables. All disturbance changes have been successfully rejected by all of the manipulated variables.
CHAPTER 7

GENERIC MODEL CONTROL (GMC) DESIGN AND IMPLEMENTATION

7.1. Overview
7.2. Implementation of GMC Controllers in MATLAB
7.3. GMC Controller Tuning
7.4. GMC Performance analysis
7.5. Summary
7.1 Overview

This chapter is associated with the design, tune and implementation of more advance controller i.e. GMC. GMC is a model-based controller which incorporates the process model to derive the manipulated variable algorithm. The algorithm will be explained briefly and the derivation of the manipulated variables will be covered.

The approach includes deriving the GMC algorithm equations and doing performance analysis in terms of set point tracking and disturbance analysis.

Finally Performance evaluation of GMC controllers was done by introducing disturbances and set point change in process variables.

7.2 Generic Model Control

7.2.1 About GMC

Real chemical processes generally behave in a nonlinear way and may change their behaviour over a period of time. Most of these processes also have a high level of interactions between the process variables. The multi variable control techniques like Model Algorithm Control (MAC) and Internal Model Control (IMC) might not be a suitable strategy while dealing with these nonlinearities. The reason is that the model used for the predictions are usually acquired either from the linearized process model or the experimentally obtained step response. These responses vary with different step sizes and they do not represent the exact model.
Generic Model Control (GMC) is a model-based control strategy that was introduced by Lee and Sullivan back in 1988. GMC uses the process model to develop a controller that follows the desired trajectory. Using the process model for developing the controller has the advantage of considering all the nonlinearities and interactions of the process. These advantages make the GMC suitable for the multivariable control systems [30].

7.2.2 GMC Algorithm

The GMC algorithm uses the process model to derive the controller’s equation. The main objective of GMC is to guide a system from its initial condition to the desired set point by manipulating its input so that the system follows the behaviour of a pre-defined reference model derived from ‘fundamental conservation and constitutive relations [22].

From previous discussion, the GMC algorithm incorporates the model based control techniques in a much generalised way.

If a process is represented by equation (4-1) given below.

\[
\frac{dy}{dt} = f(y, u, d) \tag{7 - 1}
\]

Where \( y \) represents the process output, while \( u \) and \( d \) are the input and the disturbance respectively, then equation (4-1) shows the GMC algorithm where the process variable is ‘forced’ to follow a desired or reference trajectory [30].

\[
\frac{dy}{dt}_{desired} = K_1(y_{desired} - y) + K_2 \int (y_{desired} - y) \, dt \tag{7 - 2}
\]
The term, $K_1(y_d - y)$ forces the process variable to its set-point at a speed proportional to how far away the process is. On the other hand the term $K_2 \int (y_d - y) \, dt$ accelerates the process variable towards its desired set-point. Combination of the two terms provides a satisfactory controller performance.

Since it is desired for the process variable to follow a defined trajectory, equations (4-1) and (4-2) will yield equation (4-3).

$$\frac{dy}{dt} = \left(\frac{dy}{dt}\right)_{\text{desired}} \quad (7 - 3)$$

From equation (7-3) the manipulated variable is calculated. If the model perfectly represents the process, and there are no constraints on the manipulated variable, the process will follow the desired trajectory. The integral part also compensates for any modelling errors [30].

### 7.3 GMC Implementation

The basic concept of Generic Model Control, GMC, is to find values of the manipulated inputs that force a model of the system to follow a desired reference trajectory. The method is closely related to a body of mathematical knowledge known as differential geometry involving exact linearization of nonlinear mapping between the input and output variables.

Consider a dynamic model of a process describes by a set of differential equations [5]:

$$\dot{y} = f(y, u, d, t, \theta) \quad (7 - 4)$$
In general, $f$ is a vector of nonlinear known function relationships. The second part of the algorithm is to define a reference system, $y_r$. This reference system defines a desirable rate of change of the output variables.

$$y_r = K_1(y^* - y) + K_2 \int_0^{t_k} (y^* - y) dt$$

(7 - 5)

Where $t_k$ is the current instant of time and $y^*$ is the setpoint of $y$.

This method of control results in offset free performance in addition to resulting in a quick response. Say the system is a long way from the desired setpoint, this method would quickly correct the system to travel towards the setpoint.

Now to ensure the rate of the output follows the desired reference trajectory the two equations are compared, $\dot{y} = y_r$, resulting in:

$$f(y,u,d,t,\theta) = K_1(y^* - y) + K_2 \int_0^{t_k} (y^* - y) dt$$

(7 - 6)

Once equated the expression must be rearranged to explicitly solve for the manipulated variable, MV. One of the main advantages of GMC is its ‘robust’ nature when facing disturbances and model inaccuracies.

The implementation of the algorithm is discussed in the following sub sections.

7.3.1 Reference Trajectories

The reference trajectories for the process variables ($h$, $T_1$, $T_2$, $Tout$, $C_1$, $C_2$ and $Cout$) are defined by the following equations:

$$\left(\frac{dh}{dt}\right)_{desired} = K_1 \cdot eh + K_2 \int eh \, dt$$

(7 - 7)
Where, \( e \) represents the error between the set-point and the process variable.

### 7.3.2 Manipulated Variable Equations

After defining the reference trajectory for process variables, the model equations for the same are equated to the desired trajectory equations. The model equations (G-1) through (G-7) for the system have been provided in Appendix G.

Using equation (7-3), the response of the process variable can be written as following:

\[
\left( \frac{d T_1}{dt} \right)_{desired} = K_1 e T_1 + K_2 \int e T_1 \, dt \tag{7-8}
\]

\[
\left( \frac{dT_2}{dt} \right)_{desired} = K_1 e T_2 + K_2 \int e T_2 \, dt \tag{7-9}
\]

\[
\left( \frac{dT_{out}}{dt} \right)_{desired} = K_1 e T_{out} + K_2 \int e T_{out} \, dt \tag{7-10}
\]

\[
\left( \frac{dC_1}{dt} \right)_{desired} = K_1 e C_1 + K_2 \int e C_1 \, dt \tag{7-8}
\]

\[
\left( \frac{dC_2}{dt} \right)_{desired} = K_1 e C_2 + K_2 \int e C_2 \, dt \tag{7-9}
\]

\[
\left( \frac{dC_{out}}{dt} \right)_{desired} = K_1 e C_{out} + K_2 \int e C_{out} \, dt \tag{7-10}
\]
As the loop pairing has been already performed as shown in Table 9 in the previous chapter, rearrange equations (7-11) to (716) and solve for the manipulated variables (MVs).

\[
\begin{align*}
\frac{dT_{\text{out}}}{dt}_{\text{desired}} &= \frac{1}{\rho_{\text{sw3}}.C_{p}.h.A_{3}} \left[ \rho_{\text{sw2}}C_{p} \cdot F_{2} \cdot T_{2} + \rho_{\text{sw}}C_{p} \cdot \frac{F_{d}}{3} \cdot T_{\text{in}} - \rho_{\text{sw3}}C_{p} \cdot F_{\text{out}} \cdot T_{\text{out}} + Q - UA \cdot (T_{\text{out}} - T_{\text{atm}}) \right] \\
\frac{dC_{1}}{dt}_{\text{desired}} &= \frac{1}{V_{1}} \left[ \frac{F_{d}}{3} \cdot C_{d} - F_{1} \cdot C_{1} + F_{\text{recycle}} \cdot C_{\text{out}} \right] \quad (7 - 15) \\
\frac{dC_{2}}{dt}_{\text{desired}} &= \frac{1}{V_{2}} \left[ F_{1} \cdot C_{1} + \frac{F_{d}}{3} \cdot C_{d} - F_{2} \cdot C_{2} \right] \quad (7 - 16) \\
\frac{dC_{\text{out}}}{dt}_{\text{desired}} &= \frac{1}{h.A_{3}} \left[ F_{2} \cdot C_{2} + \frac{F_{d}}{3} \cdot C_{d} - F_{\text{out}} \cdot C_{\text{out}} \right] - C_{\text{out}} \left( \rho_{\text{sw2}} \cdot F_{2} + \rho_{\text{sw}} \cdot \frac{F_{d}}{3} - \rho_{\text{sw3}} \cdot F_{\text{out}} \right) \quad (7 - 17)
\end{align*}
\]

\[
\begin{align*}
F_{\text{out}} &= \left\{ \rho_{\text{sw3}} \cdot A_{3} \ast \left[ \frac{dh}{dt}_{\text{desired}} \right] - \rho_{\text{sw2}} \cdot F_{2} - \rho_{\text{sw}} \cdot \frac{F_{d}}{3} + \rho_{\text{sw3}} \cdot F_{\text{recycle}} \right\} / - \rho_{\text{sw3}} \cdot F_{3} \quad (7 - 18)
\end{align*}
\]

\[
\begin{align*}
Q_{1} &= \rho_{\text{sw1}}C_{p}V_{1} \ast \left[ \frac{dT_{1}}{dt}_{\text{desired}} \right] - \rho_{\text{pw}}C_{p} \cdot F_{\text{in}} \cdot T_{\text{in}} - \rho_{\text{sw3}}C_{p} \cdot F_{\text{recycle}} \cdot T_{\text{out}} - \rho_{\text{sw}}C_{p} \cdot \frac{F_{d}}{3} \cdot T_{\text{in}} + \rho_{\text{sw1}}C_{p} \cdot F_{1} \cdot T_{1} + UA \cdot (T_{1} - T_{\text{atm}}) \quad (7 - 19)
\end{align*}
\]

\[
\begin{align*}
Q_{2} &= \rho_{\text{sw2}}C_{p}V_{2} \ast \left[ \frac{dT_{2}}{dt}_{\text{desired}} \right] - \rho_{\text{sw1}}C_{p} \cdot F_{1} \cdot T_{1} - \rho_{\text{sw}}C_{p} \cdot \frac{F_{d}}{3} \cdot T_{\text{in}} - \rho_{\text{sw2}}C_{p} \cdot F_{2} \cdot T_{2} + UA \cdot (T_{2} - T_{\text{atm}}) \quad (7 - 20)
\end{align*}
\]
7.3.3 GMC Parameters K1 and K2

The two controller parameters, $K_1$ and $K_2$, are determined by manipulating the general (2,1) system where:

$$
\tau = \frac{1}{\sqrt{(K_2)}} \quad \text{and} \quad \zeta = \frac{K_1}{2\sqrt{K_2}} \quad (7-25)
$$

$\zeta$ is selected to give the desired shape of response, while $\tau$ gives appropriate timing of the response in relation to the plant speed.

$$
K_1 = \frac{2\zeta}{\tau} \quad \text{and} \quad K_2 = \frac{1}{\tau^2} \quad (7-26)
$$
7.3.4 Implementation of GMC Controllers in MATLAB

To transfer GMC to MATLAB script the error and summation of errors must be defined for all loop iteration, along with the reference trajectory, \( \text{reftraj} \), and the specified equations above. An extract of the MATLAB script is given below:

\[
\begin{align*}
\text{errh} & = \text{sphg} - \text{hg}; \\
\text{errT1} & = \text{spT1g} - \text{T1g}; \\
\text{errT2} & = \text{spT2g} - \text{T2g}; \\
\text{errTout} & = \text{spToutg} - \text{Toutg}; \\
\text{errC1} & = \text{spC1g} - \text{C1g}; \\
\text{errC2} & = \text{spC2g} - \text{C2g}; \\
\text{errCout} & = \text{spCoutg} - \text{Coutg}; \\
\end{align*}
\]

\[
\begin{align*}
\text{errsumh} & = \text{errsumh} + \text{errh}; \\
\text{errsumT1} & = \text{errsumT1} + \text{errT1}; \\
\text{errsumT2} & = \text{errsumT2} + \text{errT2}; \\
\text{errsumTout} & = \text{errsumTout} + \text{errTout}; \\
\text{errsumC1} & = \text{errsumC1} + \text{errC1}; \\
\text{errsumC2} & = \text{errsumC2} + \text{errC2}; \\
\text{errsumCout} & = \text{errsumCout} + \text{errCout}; \\
\end{align*}
\]

\[
\begin{align*}
\text{reftrajh} & = k1h*\text{errh} + k2h*\text{errsumh}; \\
\text{reftrajT1} & = k1T1*\text{errT1} + k2T1*\text{errsumT1}; \\
\text{reftrajT2} & = k1T2*\text{errT2} + k2T2*\text{errsumT2}; \\
\text{reftrajTout} & = k1Tout*\text{errTout} + k2Tout*\text{errsumTout}; \\
\text{reftrajC1} & = k1C1*\text{errC1} + k2C1*\text{errsumC1}; \\
\text{reftrajC2} & = k1C2*\text{errC2} + k2C2*\text{errsumC2}; \\
\text{reftrajCout} & = k1Cout*\text{errCout} + k2Cout*\text{errsumCout}; \\
\end{align*}
\]
GMC DESIGN AND IMPLEMENTATION

% Controller Action

\[
F_{out} = \begin{pmatrix}
(psw3 \cdot A3 \cdot \text{reftrajh}) - psw2 \cdot F2g - psw3 \cdot \text{Frecyleg} - psw \cdot (Fdg/3) \\
pw \cdot \text{Cp} \cdot \text{Fing} \cdot \text{Ting} - psw3 \cdot \text{Cp} \cdot \text{Frecyleg} \cdot \text{Toutg} - psw \cdot \text{Cp} \cdot (Fdg/3) \cdot \\
pw \cdot \text{Cp} \cdot \text{V1} \cdot \text{reftrajT1} - ppw \cdot \text{Cp} \cdot \text{Fing} \cdot \text{Ting} - psw3 \cdot \text{Cp} \cdot \text{Frecyleg} \cdot \text{Toutg} - psw \cdot \text{Cp} \cdot (Fdg/3) \cdot \\
pw \cdot \text{Cp} \cdot \text{V2} \cdot \text{reftrajT2} - psw1 \cdot \text{Cp} \cdot \text{F1g} \cdot \text{T1g} - psw \cdot \text{Cp} \cdot (Fdg/3) \cdot \text{Ting} + psw2 \cdot \text{Cp} \cdot F2g \cdot T2g + (\\npsw3 \cdot \text{Cp} \cdot A3 \cdot \text{hg} \cdot \text{reftrajTout}) - psw2 \cdot \text{Cp} \cdot F2g \cdot T2g - psw \cdot \text{Cp} \cdot (Fdg/3) \cdot \text{Ting} + psw3 \cdot \text{Cp} \cdot \text{Foutg} \cdot Fdg = 3 \cdot (V1 \cdot \text{reftrajC1}) + F1g \cdot \text{C1g} - \text{Frecyleg} \cdot \text{Coutg} )/ Cing;

F_{out} = 3 \cdot (V2 \cdot \text{reftrajC2}) - F1g \cdot \text{C1g} + F2g \cdot C2g )/ Cing;

F_{out} = 3 \cdot (hg \cdot A3 \cdot \text{reftrajCout}) - F2g \cdot C2g + \text{Foutg} \cdot \text{Coutg} + psw2 \cdot F2g \cdot \text{Coutg} - psw3 \cdot \text{Foutg} \cdot \text{Coutg} )/(C
\]

% Previous PV plus Change in PV

\[
\begin{align*}
\text{hg} &= \text{hg} + dhdt; \\
T_{1g} &= T_{1g} + dT_{1dt}; \\
T_{2g} &= T_{2g} + dT_{2dt}; \\
T_{outg} &= T_{outg} + dT_{outdt}; \\
C_{1g} &= C_{1g} + dC_{1dt}; \\
C_{2g} &= C_{2g} + dC_{2dt}; \\
C_{outg} &= C_{outg} + dC_{outdt};
\end{align*}
\]

% Change in PV

\[
dhdt = \frac{1}{psw3 \cdot A3} \cdot (psw2 \cdot F2g + psw \cdot (Fdg/3) - psw3 \cdot F3g); \\
dT_{1dt} = \frac{1}{psw1 \cdot \text{Cp} \cdot \text{V1}} \cdot (ppw \cdot \text{Cp} \cdot \text{Fing} \cdot \text{Ting} + psw3 \cdot \text{Cp} \cdot \text{Frecyleg} \cdot \text{Toutg} + psw1 \cdot \text{Cp} \cdot \text{F1g} \cdot \text{T1g} + psw \cdot \text{Cp} \cdot (Fdg/3) \cdot \text{Ting} - psw2 \cdot \text{Cp} \cdot F2g \cdot \text{T2g} + psw \cdot \text{Cp} \cdot (Fdg/3) \cdot \text{Coutg}); \\
dT_{2dt} = \frac{1}{psw2 \cdot \text{Cp} \cdot \text{V2}} \cdot \text{hg} \cdot (psw2 \cdot \text{Cp} \cdot F2g \cdot \text{T2g} + psw \cdot \text{Cp} \cdot (Fdg/3)); \\
dT_{outdt} = \frac{1}{psw3 \cdot \text{Cp} \cdot A3} \cdot \text{hg} \cdot \text{Coutg} + \text{Frecyleg} \cdot \text{Coutg}); \\
dC_{1dt} = \frac{1}{V1} \cdot (Fdg/3) \cdot \text{Coutg} = \text{hg} \cdot \text{C1g} + \text{Frecyleg} \cdot \text{Coutg}); \\
dC_{2dt} = \frac{1}{V2} \cdot (F1g \cdot \text{C1g} + (Fdg/3) \cdot \text{Coutg} = \text{F2g} \cdot \text{C2g}); \\
dC_{outdt} = \frac{1}{A3} \cdot \text{hg} \cdot \text{F2g} \cdot \text{C2g} + (Fdg/3) \cdot \text{Coutg} - \text{Coutg}.)
\]
7.3.5 GMC Controller Parameters Tuning

Tuning of GMC controller is based on the selection of appropriate values of $K_1$ and $K_2$ which depends upon the selection of damping ratio to get desired the shape of response. Mathematical expressions for $K_1$ and $K_2$ are given below [30].

Where $\tau$ Time constant of the response and $\zeta$ Damping ratio of response.

These parameters are determined based on the selected trajectory. Time constant is suggested to be chosen based on the open loop time constant of the process. Damping ratio is selected to give the desired shape to the response. Shapes with less overshoot are less robust to modelling errors. Figure 42 shows a number of responses with different values of $\zeta$.

![Figure 42: Reference trajectories with different $\tau$ and $\zeta$ parameters [30].](image)
The final GMC scheme on the heated tank system consists of five GMC controllers as explained before.

Similar to tuning PI controllers, tuning of the GMC also consists of selecting two parameters. The parameters for all loops were tuned by giving the set-point change to each process variable and observing the response.

As explained before the best estimate of $\tau$ is the open loop time constant of the concerned process variable.

### 7.3.5.1 GMC Controller Tuning Results

Table 12 illustrates GMC Tuning Results.

<table>
<thead>
<tr>
<th>PV</th>
<th>Kc</th>
<th>Tau</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Heated Tank 1</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>T1</td>
<td>10.15</td>
<td>2.992364</td>
</tr>
<tr>
<td>C1</td>
<td>6.147</td>
<td>4.76</td>
</tr>
<tr>
<td><strong>Heated Tank 2</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>T2</td>
<td>8.761347</td>
<td>3.162297</td>
</tr>
<tr>
<td>C2</td>
<td>8.051666</td>
<td>0.016466</td>
</tr>
<tr>
<td><strong>Heated Tank 3</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Tout</td>
<td>9.29</td>
<td>3.3333</td>
</tr>
<tr>
<td>Cout</td>
<td>7.36147</td>
<td>5.216666</td>
</tr>
<tr>
<td>h</td>
<td>-0.0077</td>
<td>30.64808</td>
</tr>
</tbody>
</table>
7.4 Performance Analysis

Performance analysis is based on set-point tracking and disturbance rejection along with the ISE, IAE, and ITAE values for the process variables, which will be discussed in the next chapter.

7.4.1 Set-Point Tracking
Similar to PI controllers GMC controllers were tested against a set-point change as follows:

- T1 ± 5 Degree C
- T2 ± 5 Degree C
- Tout ± 5 Degree C
- C1 ± 0.005 Kg/L
- C2 ± 0.005 Kg/L
- Cout ± 0.005 Kg/L
- h ± 5%

GMC controllers have effectively tracked all set-point changes as can be seen in the following figures.

1st Test: T1 – 5 Degree C

Figure 43: T1 Setpoint Tracking -5 Degree C  GMC
GMC DESIGN AND IMPLEMENTATION

As can be seen in Figure 43 that a step down change of size (5), has been given in process variable of heated tank 1 (T1) which labelled by the red colour. Consequently, the manipulated variable (Q1), (labelled by the green colour) started to close directly in order to track the desired set point.

2nd Test: Tout – 5 Degree C

As can be seen in Figure 44 that a step down change of size (5), has been given in process variable of heated tank 3 (Tout) which labelled by the red colour. Consequently, the manipulated variable (Q3), (labelled by the green colour) started to close directly in order to track the desired set point.

All other responses of setpoint changes can be found in Appendix J-3.
7.4.2 Disturbance Analysis

GMNC controllers were tested against a disturbance change as follows:

- Fin ±5%
- Frecycle ±0.5L/min
- Tin ±5 Degree C
- Cin ±5%

1st Test: Tin +5 Degree C

As can be seen in Figure 45 that a step up change of size (+5 degrees C), (from 26 degrees C to 31 degrees C), has been given in the disturbance variable Tin at the
suction of CSTRs bank system. Therefore, all of three process variables T1, T2 and Tout (labelled by the red colour) have been influenced and went high. Consequently, all of the relative manipulated variables Q1, Q2 and Q3 (labelled by the green colour) started to close automatically in order to reject the disturbance and keep the process variable at the desired set point.

2nd Test: Tin –5 Degree C

As can be seen in Figure 46 that a step up change of size (-5 degrees C), (from 26 degrees C to 21 degrees C), has been given in the disturbance variable Tin at the suction of CSTRs bank system. Therefore, all of three process variables T1, T2 and Tout (labelled by the red colour) have been influenced and went low. Consequently, all of the relative manipulated variables Q1, Q2 and Q3 (labelled by the green colour)
started to open automatically in order to reject the disturbance and keep the process variable at the desired set point.

All other responses of disturbance changes can be found in Appendix J-4.

The other conclusion that can be seen from the response graph is that changes in the manipulated variable are not mirror image of each other i.e. the change for the positive and negative disturbance change is different as the system is a non-linear system.

7.5 Summary

Generic Model Controller (GMC) was implemented in this part of the report. The behaviour of the controller and its effect on the other system parameters was observed. Since the GMC is a model based and in real implementation it has to compensate for any model discrepancies.

The GMC so far is the best implementation on the three heated tank system given that the control values can cope up the limits of the manipulated variables changes.
CHAPTER 8

CONTROL PERFORMANCE ANALYSIS

8.1. Overview
8.2. Control Performance Measures
8.3. Statistical Performance Control (SPC)
8.4. Control Performance Comparison (ISE, IAE and ITAE)
8.5. Summary
8.1 Overview

This chapter of the report covers the performance analysis of the two controller schemes PI and GMC.

The behaviour of the controller to different step changes to process variables and disturbance variables will be discussed based on various ways of quantifying cumulative error in a system such as ISE, IAE and ITAE.

One of the statistical performance control chart will be addressed in this chapter showing the control behaviour for both of the control schemes (PI and GMC).

8.2 Control Performance Measures

There are two types of measures typically used to evaluate the performance of controllers and their “ability to follow the set point and reject disturbances”: practical performance measures and general performance measures [12].

8.2.1 Practical Performance Measures

Practical performance measures provide a method of analysis which is easy to apply to a real system [11]. These methods include the measurement and comparison of various physical attributes of the system response, including:

- **Rise Time**: the difference in time between the change in set point and the time at which the process variable first achieves that set point.
- **Overshoot**: the amount by which the process variable continues beyond the desired set point.
- **Decay Ratio**: the rate at which the amplitude of the response is reduced for each successive oscillation.
- **Settling Time**: the time taken for the process variable to return to steady state at the new set point.
- **Offset** the persisting error between the set point and the process variable after reaching steady state.

### 8.2.2 General Performance Measures

General performance measures look instead at the cumulative error that occurs in a system. These types of control methods are very precise and give exact comparisons between different control schemes, or for various sets of tuning parameters [12]. These measures are calculated by performing a set point or disturbance change to a process variable and evaluating the integral over a fixed period, determined by the time taken for the system response to settle [12].

There are various ways of quantifying cumulative error in a system [11] which are as follows:

- **Integral Error (IE)**, which is the cumulative sum of the error [12].

- **Integral Squared Error (ISE)**, which penalises large errors more than small since the magnitude of errors are squared, large errors will be significantly greater than small errors. Control systems specified to minimise ISE will tend to eliminate large errors quickly but will tolerate small persistent errors. Often this leads to fast responses, however, allows prolonged small-amplitude oscillation [12].

- **Integral Absolute Error (IAE)**, which is the sum of areas above and below the set-point. This method penalises all errors equally, regardless of direction. No weighting is added to the errors in the system response. Therefore it tends to result in slower performance than ISE, but exhibits significantly reduced persistent oscillation [12].

- **Integral Time Weighted Absolute Error (ITAE)**, which penalises persistent errors. Errors which exist after a long time are weighted much more heavily than those at the start of the response. Tuning according to this method produces responses which settle much more quickly, but may have a sluggish initial response [12].
The equations used to calculate these performance measures are shown below, equations (8—1), (8—2) and (8—3):

\[ ISE = \int e^2 dt \]  \hspace{1cm} (8—1)

\[ IAE = \int |e| dt \]  \hspace{1cm} (8—2)

\[ ITAE = \int t |e| dt \]  \hspace{1cm} (8—3)

In spite of the fact, the practical performance measures have not been used because it is difficult to distinguish between control responses and determine if the control scheme provides the ideal response which yields the greatest number of ‘passes’. Consequently, it was essential to develop a more simplified approach which is a direct, quantifiable measure of how well the response followed the ‘desired response’ [12].

By evaluating the integral squared error (ISE), control schemes are directly assessed according to their ability to eliminate large initial errors, therefore by minimising ISE; the report can identify the control scheme most suited to situations where a quick response is crucial for optimal operating conditions [12].

8.2.3 Control Charts Measures

The control chart can be defined as a graph used to exam how a process varies over time. The control chart generally has a central line that illustrates the mean, a lower line that demonstrates the lower control limit (LCL) and an upper line that shows the upper control limit. The current data are compared with previous historical data using these three lines. Then the process variation is considered either is out of control if the change unpredictable (special causes of variation) or the difference is considered as in control if the variation is consistent. Indeed, the control chart has been implemented in this case study to evaluate the performance of the controllers.
8.3 Control Performance Analysis

8.3.1 Statistical Performance Analysis

The goal of statistical process monitoring (SPM) is to detect the existence, magnitude, and time of occurrence of changes that cause a process to deviate from its desired operation.

The methodology for detecting changes is based on statistical techniques that deal with the collection, classification, analysis, and interpretation of data [29].

An attempt had been made to implement some statistical performance measures for the controllers. The two methods explored were Schewart and exponentially weighted Moving Average (EWMA) [10].

8.3.1.1 Approximating Process operation in Real time

Since the statistical process needs measured data from measurement devices kept under control and calibrated. Implementing it in a simulated environment was a challenge. In order to get data for analysis that may be equal to real world process random disturbances were given to the system in steady state. The data was collected and analysed.

8.3.1.2 Collecting and analysing process data

As mentioned earlier disturbance changes of different magnitudes and variables were given to the process for around 2000 min in MATLAB. The disturbances were given to the system such that the system stabilises itself before giving it a new
disturbance change. The data was then collected in block of 100 min for each process variable. Standard deviations of the variables were then calculated.

8.3.1.3 Schewart Chart

The upper and lower controller limits shown in Table 12 were used for the plotting of Schewart Charts for all of the seven process variables in both controllers (PI and GMC).

Table 13: Schewart chart, Upper and Lower Controller Limits

<table>
<thead>
<tr>
<th>Schewart Chart (Water Level)</th>
<th>h</th>
</tr>
</thead>
<tbody>
<tr>
<td>UCL</td>
<td>0.44</td>
</tr>
<tr>
<td>LCL</td>
<td>0.38</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Schewart Chart (Temperature)</th>
<th>T1</th>
<th>T2</th>
<th>Tout</th>
</tr>
</thead>
<tbody>
<tr>
<td>UCL</td>
<td>34.875</td>
<td>48.255</td>
<td>61.095</td>
</tr>
<tr>
<td>LCL</td>
<td>34.275</td>
<td>47.455</td>
<td>60.295</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Schewart Chart (Concentrations)</th>
<th>C1</th>
<th>C2</th>
<th>Cout</th>
</tr>
</thead>
<tbody>
<tr>
<td>UCL</td>
<td>0.00084</td>
<td>0.00117</td>
<td>0.00147</td>
</tr>
<tr>
<td>LCL</td>
<td>0.00016</td>
<td>0.00017</td>
<td>0.00047</td>
</tr>
</tbody>
</table>
CONTROL PERFORMANCE ANALYSIS

8.4 Statistical Performance Test

8.4.1 PI Schewart Charts

While the controllers are holding great for some random disturbance change, the following figures show PI Schewart Charts for all of the process variables.

Figure 47: Schewart Chart: T1 Response (Random disturbance Changes Tin)

Figure 48: Schewart Chart: T2 Response (Random disturbance Changes Tin)
Sometimes some process variables will go out of the limit performance. The reason for that can be due to the fact that the limits are calculated from simulated data and not real process data.
Figure 51: C2 Response (Random disturbance Changes Tin)

Figure 52: Cout Response (Random disturbance Changes Tin)

Figure 53: h Response (Random disturbance Changes Tin)
8.4.2 GMC Schewart Charts

The following figures show GMC Schewart Charts for all of the process variables when a random disturbances were introduced.

Figure 54: Schewart Chart: T1 Response (Random disturbance Changes).

Figure 55: Schewart Chart: T2 Response (Random disturbance Changes)
Figure 56: Schewart Chart: Tout Response (Random disturbance Changes)

Figure 57: : h Response (Random disturbance Changes).
Figure 58: C1, C2 and Cout Responses (Random disturbance Changes)
8.5 Control Performance comparison between PI and GMC

8.5.1 Set-point Tracking

The point of interest in this section however is the ISE, IAE, and ITAE values. Table 14 and Table 15 show setpoint change comparison between PI and GMC based on the ISE, IAE, and ITAE values.

Table 14: Setpoint Changes T1: ISE, IAE, and ITAE (PI and GMC)

<table>
<thead>
<tr>
<th>Setpoint Changes T1 ± 5 Degree C</th>
<th>ISE</th>
<th>PI</th>
<th>GMC</th>
<th>IAE</th>
<th>PI</th>
<th>GMC</th>
<th>ITAE</th>
</tr>
</thead>
<tbody>
<tr>
<td>T1 +5 C</td>
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<td>0</td>
<td>0</td>
<td>0.0011</td>
<td>0</td>
<td>0.0021</td>
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<td></td>
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<tr>
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<td>10.0234</td>
<td>0.3574</td>
<td>12.5828</td>
<td>0.0784</td>
<td></td>
</tr>
<tr>
<td>T2</td>
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<td>0.1342</td>
<td>0</td>
<td>0.1356</td>
<td>0</td>
<td></td>
</tr>
<tr>
<td>Tout</td>
<td>0.0006</td>
<td>0</td>
<td>0.233</td>
<td>0</td>
<td>0.2308</td>
<td>0</td>
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<tr>
<td>C1</td>
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<td>0</td>
<td>0.0087</td>
<td>0</td>
<td>0.0087</td>
<td>0.0000</td>
<td></td>
</tr>
<tr>
<td>C2</td>
<td>0</td>
<td>0</td>
<td>0.0087</td>
<td>0</td>
<td>0.0088</td>
<td>0</td>
<td></td>
</tr>
<tr>
<td>Cout</td>
<td>0</td>
<td>0</td>
<td>0.0087</td>
<td>0</td>
<td>0.0087</td>
<td>0</td>
<td></td>
</tr>
<tr>
<td>T1 -5 C</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
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<td></td>
</tr>
<tr>
<td>h</td>
<td>0</td>
<td>0</td>
<td>0.0011</td>
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<td>0.0021</td>
<td>0</td>
<td></td>
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<tr>
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<td>0.3574</td>
<td>17.5423</td>
<td>0.0784</td>
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</tr>
<tr>
<td>T2</td>
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<td>0</td>
<td>0.1342</td>
<td>0</td>
<td>0.1356</td>
<td>0</td>
<td></td>
</tr>
<tr>
<td>Tout</td>
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<td>0.233</td>
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<td>0.2307</td>
<td>0</td>
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<tr>
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<td>0</td>
<td>0.0087</td>
<td>0.0000</td>
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</tr>
<tr>
<td>C2</td>
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<td>0.0087</td>
<td>0</td>
<td>0.0088</td>
<td>0</td>
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</tr>
<tr>
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<td>0.0087</td>
<td>0</td>
<td>0.0087</td>
<td>0</td>
<td></td>
</tr>
</tbody>
</table>

Table 14 and Table 15 illustrate that, for different step sizes to the process variables, the ISE, IAE, and ITAE values for the GMC for step change in $T_1$ are slightly lower as compared to the other controller scheme. However, given the choice between
implementation of any of these schemes on the system, the GMC will be the preferred one due to its ‘beauty’ of eliminating all the interactions.

Table 15: Set-point Changes T2: ISE, IAE, and ITAE (PI and GMC)

<table>
<thead>
<tr>
<th>Setpoint Changes T2 ± 5 Degree C</th>
<th>ISE</th>
<th></th>
<th>IAE</th>
<th></th>
<th>ITAE</th>
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<tr>
<td></td>
<td>PI</td>
<td>GMC</td>
<td>PI</td>
<td>GMC</td>
<td>PI</td>
<td>GMC</td>
</tr>
<tr>
<td>T2 +5 C</td>
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<td></td>
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<tr>
<td>h</td>
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<td>0</td>
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<td>0</td>
<td>0.0021</td>
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<tr>
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<td>0.0087</td>
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</tr>
<tr>
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<tr>
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<td>0</td>
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</tr>
<tr>
<td>T1</td>
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<td>0.2843</td>
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<tr>
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</tr>
<tr>
<td>C1</td>
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</tr>
<tr>
<td>C2</td>
<td>0</td>
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<td>0.0088</td>
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</tr>
<tr>
<td>Cout</td>
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<td>0.0087</td>
<td>0.0000</td>
<td>0.0087</td>
<td>0</td>
</tr>
</tbody>
</table>

All other tables of setpoint changes comparison based on the ISE, IAE, and ITAE values can be found in Appendix K-1.
8.5.2 Disturbance Rejection

Table 16 and Table 17 show disturbance changes comparison between PI and GMC based on the ISE, IAE, and ITAE values.

Table 16: Tin Disturbance Change: ISE, IAE, and ITAE (PI and GMC)

<table>
<thead>
<tr>
<th>Disturbance Changes Tin ± 5 Degree C</th>
<th>ISE</th>
<th></th>
<th>IAE</th>
<th></th>
<th>ITAE</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>PI</td>
<td>GMC</td>
<td>PI</td>
<td>GMC</td>
<td>PI</td>
</tr>
<tr>
<td>Tin +5 C</td>
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<td></td>
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</tr>
<tr>
<td>h</td>
<td>0</td>
<td>0</td>
<td>0.0011</td>
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<td>0.0021</td>
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<tr>
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<td>10.5428</td>
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</tr>
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<td>0.0087</td>
</tr>
<tr>
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<td>0</td>
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</tr>
<tr>
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<td>0.0087</td>
<td>0.0000</td>
<td>0.0087</td>
</tr>
<tr>
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<td></td>
</tr>
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</tr>
<tr>
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<td>0.0087</td>
<td>0</td>
<td>0.0088</td>
</tr>
<tr>
<td>Cout</td>
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<td>0</td>
<td>0.0087</td>
<td>0.0000</td>
<td>0.0087</td>
</tr>
</tbody>
</table>

The values in tables add to the previous choice of control scheme selection. There is no effect of disturbances on GMC as compared to the other controller scheme.
Table 17: Fin and Frecycle Disturbance Changes: ISE, IAE, and ITAE (PI and GMC)

<table>
<thead>
<tr>
<th></th>
<th>ISE</th>
<th></th>
<th>IAE</th>
<th></th>
<th>ITAE</th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
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<td>GMC</td>
<td>PI</td>
<td>GMC</td>
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<td>GMC</td>
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<td>18.6027</td>
<td>0</td>
</tr>
<tr>
<td>T1</td>
<td>0.2418</td>
<td>0</td>
<td>4.4535</td>
<td>0</td>
<td>5.55</td>
<td>0</td>
</tr>
<tr>
<td>T2</td>
<td>0.0126</td>
<td>0</td>
<td>1.0281</td>
<td>0</td>
<td>1.2573</td>
<td>0</td>
</tr>
<tr>
<td>Tout</td>
<td>0.0021</td>
<td>0</td>
<td>0.0976</td>
<td>0</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>C1</td>
<td>0.6725</td>
<td>0</td>
<td>7.2711</td>
<td>0</td>
<td>8.515</td>
<td>0.0000</td>
</tr>
<tr>
<td>C2</td>
<td>0.0968</td>
<td>0</td>
<td>2.6745</td>
<td>0</td>
<td>3.9282</td>
<td>0</td>
</tr>
<tr>
<td>Cout</td>
<td>0.0198</td>
<td>0</td>
<td>0.5954</td>
<td>0.0000</td>
<td>0.025</td>
<td>0</td>
</tr>
</tbody>
</table>

All other tables of disturbance changes comparison based on the ISE, IAE, and ITAE values can be found in Appendix K-2.
8.6 Summary

The performance of the GMC controller was compared with the previously implemented scheme (PI) and was found to be far better. Statistical performance analysis was simulated with a random number generator (‘noise’). The GMC is the better than PI.
PART SIX: CONCLUSION AND FUTURE WORK

CHAPTER 9

CONCLUSION AND FUTURE WORK

9.1. Overview

9.2. Conclusion

9.3. Future Work
CONCLUSION AND FUTURE WORK

9.1 Overview

This chapter consists of two parts, which are as follows:

- Part one: is the conclusion, which will summarise the findings of this thesis project.
- Part two: will address the future work improvements and suggestions

9.2 Conclusion

An attempt was made to define the objectives of the report in clear. Process models have been developed using mass, energy and component balances. The system under investigation was analysed and its behaviour was studied by implementing a dynamic model in MATLAB. PI controllers were designed and tuned with different methods.

Generic Model Controller (GMC) was designed, tuned and implemented. The behaviour of the controller and its effect on the other system parameters was observed. The performance of the two controllers was then compared based on ISE, IAE and ITAE and was found that GMC is better than PI. Statistical performance analysis was simulated with a random number generator (‘noise’). In the end some statistical performance charts were implemented for controller performance analysis.

The GMC is the best implementation on the heated tank system given that the control values can cope up the limits of the manipulated variables changes.
CONCLUSION AND FUTURE WORK

9.3 Future Work

As this is simulation based project, simulation using MATLAB has been completed in this project. Therefore, the following could be as a future work, which would be a continuation of this project:

- Create fully Control schematic on spread-sheet using MEDE and deal with a real process.
- Optimisation
- Using Fin as MV to control concentration
- Design of more advanced control schemes such as DMC.
CONCLUSION AND FUTURE WORK

REFERENCES


CONCLUSION AND FUTURE WORK


CONCLUSION AND FUTURE WORK


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<tr>
<th>Appendix</th>
<th>Title</th>
</tr>
</thead>
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</tr>
<tr>
<td>9.12.</td>
<td>Appendix L: Conductivity</td>
</tr>
</tbody>
</table>
## Appendix A: Plant Tags

### Table 18: Needle Tank Parameter Tags.

<table>
<thead>
<tr>
<th>Needle Tank</th>
<th>Point ID</th>
<th>Point Parameter</th>
<th>Type</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Needle - CSTR Pump On/Off</td>
<td>NTP_OFF_561</td>
<td>NTP_OFF_561.PVFL</td>
<td>Read/Write</td>
<td>0 (OFF), 1 (ON)</td>
</tr>
<tr>
<td>Needle - CSTR Pump OP</td>
<td>NTP_REF_561</td>
<td>NTP_561.PV</td>
<td>Read/Write</td>
<td>0-60%</td>
</tr>
<tr>
<td>Needle - CSTR Pump Flow Rate</td>
<td>FT_569</td>
<td>FT_569.PV</td>
<td>Read</td>
<td></td>
</tr>
<tr>
<td>Needle - CSTR OP (Control Valve)</td>
<td>FCV_570</td>
<td>FCV_570.PV</td>
<td>Read/Write</td>
<td>100%</td>
</tr>
<tr>
<td>Needle - Ball Mill Pump Flow Rate</td>
<td>FT_573</td>
<td>FT_573.PV</td>
<td>Read</td>
<td></td>
</tr>
<tr>
<td>Needle - Storage Pump OP (Control Valve)</td>
<td>FCV_571</td>
<td>FCV_571.PV</td>
<td>Read/Write</td>
<td>0-10%</td>
</tr>
<tr>
<td>Needle - CSTR Flow Temperature</td>
<td>TT_568</td>
<td>TT_568.PV</td>
<td>Read</td>
<td></td>
</tr>
</tbody>
</table>

### Table 19: Heated Tank Parameter Tags.

<table>
<thead>
<tr>
<th>CSTR</th>
<th>Point ID</th>
<th>Point Parameter</th>
<th>Type</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Needle - CSTR OP (Control Valve)</td>
<td>FCV_570</td>
<td>FCV_570.PV</td>
<td>Read/Write</td>
<td>100%</td>
</tr>
<tr>
<td>CSTR Product Pump On/Off</td>
<td>PP_OFF_681</td>
<td>PP_ON_OFF.PVFL</td>
<td>Read/Write</td>
<td>0 (OFF), 1 (ON)</td>
</tr>
<tr>
<td>CSTR Product Pump OP</td>
<td>PP_REF_681</td>
<td>PP_681.PV</td>
<td>Read/Write</td>
<td>0-60%</td>
</tr>
<tr>
<td>CSTR Product Recycle Valve</td>
<td>FCV_690</td>
<td>FCV_690.PV</td>
<td>Read/Write</td>
<td>0-10%</td>
</tr>
<tr>
<td>CSTR Product Recycle Flow Rate</td>
<td>FT_689</td>
<td>FT_689.PV</td>
<td>Read</td>
<td></td>
</tr>
<tr>
<td>CSTR Product Valve</td>
<td>FCV_688</td>
<td>FCV_688.PV</td>
<td>Read/Write</td>
<td>0-10%</td>
</tr>
<tr>
<td>CSTR Product Pump Flow Rate</td>
<td>FT_687</td>
<td>FT_687.PV</td>
<td>Read</td>
<td></td>
</tr>
<tr>
<td>CSTR 1 Temperature</td>
<td>TT_623</td>
<td>TT_623.PV</td>
<td>Read</td>
<td></td>
</tr>
<tr>
<td>CSTR 2 Temperature</td>
<td>TT_643</td>
<td>TT_643.PV</td>
<td>Read</td>
<td></td>
</tr>
<tr>
<td>CSTR 3 Temperature</td>
<td>TT_663</td>
<td>TT_663.PV</td>
<td>Read</td>
<td></td>
</tr>
<tr>
<td>CSTR 1 Steam Supply</td>
<td>FCV_622</td>
<td>FCV_622.PV</td>
<td>Read/Write</td>
<td>0-10%</td>
</tr>
<tr>
<td>CSTR 2 Steam Supply</td>
<td>FCV_642</td>
<td>FCV_642.PV</td>
<td>Read/Write</td>
<td>0-10%</td>
</tr>
<tr>
<td>CSTR 3 Steam Supply</td>
<td>FCV_662</td>
<td>FCV_662.PV</td>
<td>Read/Write</td>
<td>0-10%</td>
</tr>
<tr>
<td>CSTR Stem Supply Flow</td>
<td>PT_669</td>
<td>PT_669.PV</td>
<td>Read</td>
<td></td>
</tr>
<tr>
<td>CSTR Agitator 1</td>
<td>AG_621</td>
<td>AG_ON_OFF.PVFL</td>
<td>Read/Write</td>
<td>0 (OFF), 1 (ON)</td>
</tr>
<tr>
<td>CSTR Agitator 2</td>
<td>AG_641</td>
<td>AG_ON_OFF.PVFL</td>
<td>Read/Write</td>
<td>0 (OFF), 1 (ON)</td>
</tr>
<tr>
<td>CSTR Agitator 3</td>
<td>AG_661</td>
<td>AG_ON_OFF.PVFL</td>
<td>Read/Write</td>
<td>0 (OFF), 1 (ON)</td>
</tr>
</tbody>
</table>
Appendix B: Calibration Approach

Table 20: Steam Flow-rate Measurement for FCV_642

<table>
<thead>
<tr>
<th>CVTR1Steam (FCV_622)</th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>10%</td>
<td>0.225</td>
<td>0.1125</td>
</tr>
<tr>
<td>20%</td>
<td>0.4</td>
<td>0.2</td>
</tr>
<tr>
<td>30%</td>
<td>0.5</td>
<td>0.25</td>
</tr>
<tr>
<td>40%</td>
<td>0.69</td>
<td>0.345</td>
</tr>
<tr>
<td>50%</td>
<td>0.85</td>
<td>0.425</td>
</tr>
<tr>
<td>60%</td>
<td>0.9</td>
<td>0.45</td>
</tr>
<tr>
<td>70%</td>
<td>1.025</td>
<td>0.5125</td>
</tr>
<tr>
<td>80%</td>
<td>1.19</td>
<td>0.595</td>
</tr>
<tr>
<td>90%</td>
<td>1.25</td>
<td>0.625</td>
</tr>
<tr>
<td>100%</td>
<td>1.375</td>
<td>0.6875</td>
</tr>
</tbody>
</table>

Figure 59: FCV_622 Calibration

Table 21: Steam Flow-rate Measurement for FCV_662

<table>
<thead>
<tr>
<th>CSTR1Steam</th>
<th>% Open</th>
<th>V(L)</th>
<th>F(L/min)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>20%</td>
<td>0.36</td>
<td>0.156522</td>
</tr>
<tr>
<td></td>
<td>40%</td>
<td>0.54</td>
<td>0.26129</td>
</tr>
<tr>
<td></td>
<td>50%</td>
<td>0.5</td>
<td>0.319149</td>
</tr>
<tr>
<td></td>
<td>60%</td>
<td>0.61</td>
<td>0.411236</td>
</tr>
<tr>
<td></td>
<td>70%</td>
<td>0.62</td>
<td>0.523944</td>
</tr>
<tr>
<td></td>
<td>80%</td>
<td>0.85</td>
<td>0.671053</td>
</tr>
<tr>
<td></td>
<td>90%</td>
<td>1.05</td>
<td>0.741176</td>
</tr>
<tr>
<td></td>
<td>100%</td>
<td>0.73</td>
<td>0.755172</td>
</tr>
</tbody>
</table>
Figure 60: FCV_622 Calibration

Table 22: Pump Flow-rate Measurement for FT_569

<table>
<thead>
<tr>
<th>Needle Tank</th>
<th>FT_569</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pump(561)NT</td>
<td>OP</td>
</tr>
<tr>
<td></td>
<td>0</td>
</tr>
<tr>
<td></td>
<td>3.5</td>
</tr>
<tr>
<td></td>
<td>20</td>
</tr>
<tr>
<td></td>
<td>40</td>
</tr>
<tr>
<td></td>
<td>100</td>
</tr>
</tbody>
</table>

Figure 61: FT_569 Calibration
Appendix C: Implementation of Control System using Experion C300

The control system in the pilot plant is facilitated by the Honeywell Experion C300 software as a distribution control system. It has a server and one or more clients; those are Experion C300 Controller interface and Experion C300 Station software. Indeed, Controller interface act as a hub between the process instruments and station software with which read or write operation of the process instruments are done by the operator using the Experion station software. Furthermore, by using the suitably designed program, it is easy to operate the plant either in the manual (with an operator) or automatic (without an operator) mode [8].

![Experion Platform Architecture](image_url)

The station software can manipulate or read process data from the various actuators of the pilot plant in real time. Also, it can store and display the history values of different actuators. Figure 11 illustrates Experion Platform Architecture.
APPENDICES

Experion system can be used for small systems as well as large systems. It has the strength and robustness to control all levels of process control, safely and more efficiently. It has Distributed Control systems (DCS) capabilities of Experion system includes Abnormal Situation Management, safety management and information management technology. Fieldbus, Profibus, Device Net, HART, LON, Control Net and Interbus etc. can be interfaced with Experion. Figure 28 shows the Basic Experion Topology for C200 Controllers.

![Basic Experion Topology for C200 Controllers](image)

The tools for supervisory control include single or multiple servers and single or multiple client computers which runs Experion’s user interface called ‘Stations’.
These servers and workstation computers form the building blocks for the plant-wide control.

Appendix C.1 Experion Server

Experion server runs on windows platform. Supervisory controls and Experion Global data infrastructure are there on this server. The server serves as the central storage for the data. The following are all central systems which are running on the server [6]:

- Data acquisition and processing
- Alarm and event management
- History collection, archiving and trending
- Reporting subsystem
- Sign-on security
- Specialist and user applications
- Running in primary (and secondary) server nod.

Appendix C.2 Distributed System Architecture

The DSA allows some terminals to reuse the same data, messages, or alarm without the necessity of storing information in a server. The different servers can use the same data. In this way, the Experion system makes all clients and servers to work in a uniform pattern. This makes the efficiency of the system to improve substantially [6].

Appendix C.3 Alarm and Event Management
Experion has a well-advanced alarm system to alert the operator and helps to find out the sources of troublesome data or event and its management. It allows the system to capture and store the critical process information in the event of failure or hazards. Figure 29 displays the alarm list in the station software.

Appendix C.4 Stations

The ‘Station’ is the HMI system for the Experion server. It also provides various kinds of ‘Station’ software. The station software has an overview page and different subpages such as the storage tanks, the ball mill, the cyclone stage (including cyclone underflow and lamella tank), the needle and nonlinear tank, CSTR bank etc.
Appendix C.5 Data Exchange

Experion can be fully integrated with the existing third-party interfaces such as Excel. There are many ways by which data from the server can exchange to other databases or programs [8].

Appendix C.5.1 Open Database Connectivity (ODBC) Driver

This driver software permits other ODBC compliant programs such as excel, office, visual basic, crystal reports etc.to communicate (read/write) data from the server [6].

Appendix C.5.2 Open Database Connectivity (ODBC) Data Exchange

This functionality allows other ODBC compliant database to exchange data. This can be done periodically or any other way required by the particular application [8].

Appendix C.5.3 Object Linking and Embedding (OLE) for Process control

It has some interfaces namely Experion OPC client interface, Experion OPC advanced client, Experion display data client etc. [6].

Appendix C.5.4 Microsoft Excel Data Exchange (MSEDE)

MEDE has a wizard which enables the excel sheet to read/write point value or historical value from the server.
Appendix C.5.5 Experion Application Programming Interface (API)

Programmers can create an application on Experion server. They can use C/C++ for this. The Experion API enables this application. The API library has many resources which may be helpful to the application developers [6].

Appendix C.5.6 Network API

Using these API programmers can develop client/server applications in visual c++/visual necessary Experion also support other third-party interfaces such as Excel and OPC to exchange data between Experion server and client. In the pilot plant automation, excel is the tool supplied by Microsoft used as an HMI for the plant monitoring and control system [8].

Appendix D: Conductivity Sensor, E+H Indumax CLS50D Technical Information

Table 23: Indumax CLS50D Specifications [9].

<table>
<thead>
<tr>
<th>Measuring Principle</th>
<th>Inductive</th>
</tr>
</thead>
<tbody>
<tr>
<td>Application</td>
<td>Wastewater, process</td>
</tr>
<tr>
<td>Measurement range</td>
<td>2µS/cm - 2000 mS/cm</td>
</tr>
<tr>
<td>Measuring principle</td>
<td>Inductive conductivity measurement</td>
</tr>
<tr>
<td>Material</td>
<td>PEEK (Polyether ether ketone)</td>
</tr>
<tr>
<td>Dimension</td>
<td>Electrode:</td>
</tr>
<tr>
<td></td>
<td>outside diameter: approx. 47 mm</td>
</tr>
<tr>
<td></td>
<td>inside diameter: approx. 15 mm</td>
</tr>
<tr>
<td>Process temperature</td>
<td>max. 125°C (PEEK)</td>
</tr>
<tr>
<td>Process pressure</td>
<td>max. 20 bar (PEEK)</td>
</tr>
<tr>
<td>Ingress protection</td>
<td>IP67</td>
</tr>
</tbody>
</table>
Appendix E: Direct Synthesis Controller Design

Using the approach outlined in Ogunnaikke [40].

$G_p$ is the first order approximation of the process

$$G_p = \frac{K_p}{\tau_p s + 1} \quad (E - 1)$$

$q$ is the desired first order response

$$q(s) = \frac{K_d}{\tau_d s + 1} \quad (E - 2)$$

set $K_d = 1$ as no steady state offset is desired

$G_c$ is the controller algorithm, calculated as follows:

$$G_c = \frac{1}{g_p} \left( \frac{q}{1 - q} \right) \quad (E - 3)$$

$$G_c = \frac{1}{K_p} \left( \frac{1}{\tau_d s + 1} \right) \left( \frac{1}{1 - \frac{1}{\tau_d s + 1}} \right) \quad (E - 4)$$

$$G_c = \frac{\tau_p s + 1}{K_p} \left( \frac{1}{\tau_d s + 1} \right) \quad (E - 5)$$

$$G_c = \frac{\tau_p s + 1}{K_p} \left( \frac{1}{\tau_d s} \right) \quad (E - 6)$$

$$G_c = \frac{\tau_p s + 1}{K_p \tau_d s} \quad (E - 7)$$
APPENDICES

\[ G_c = \frac{\tau_p}{K_p \tau_d} s + \frac{1}{K_p \tau_d} s \]  \hspace{1cm} (E - 8)

\[ G_c = \frac{\tau_p}{K_p \tau_d} + \left( \frac{1}{K_p \tau_d} \times \frac{1}{s} \right) \]  \hspace{1cm} (E - 9)

\[ G_c = \frac{1}{K_p \tau_d} + \left( \tau_p \times \frac{1}{s} \right) \]  \hspace{1cm} (E - 10)

\[ G_c = \frac{\tau_p}{K_p \tau_d} + \left( 1 \times \frac{1}{\tau_p s} \right) \]  \hspace{1cm} (E - 11)

Rearrange to be in the form of a standard PI controller.

Controller gain and integral time can be found as follows:

\[ G_c = K_c \times \left( 1 \times \frac{1}{\tau_i s} \right) \]  \hspace{1cm} (E - 12)

\[ K_c = \left( \frac{\tau_p}{K_p \tau_d} \right) \]  \hspace{1cm} (E - 13)

\[ \tau_i = \frac{1}{\tau_s} \]  \hspace{1cm} (E - 14)
### Appendix F: Steady State Parameters Values

**Table 24: Variables steady state values**

<table>
<thead>
<tr>
<th>S. No.</th>
<th>Variable</th>
<th>Description</th>
<th>Value</th>
<th>Units</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>$F_{in}$</td>
<td>Feed flow rate</td>
<td>5.963</td>
<td>$\frac{L}{min}$</td>
</tr>
<tr>
<td>2</td>
<td>$F_{out}$</td>
<td>Product flow rate</td>
<td>6.298</td>
<td>$\frac{L}{min}$</td>
</tr>
<tr>
<td>3</td>
<td>$F_{recycle}$</td>
<td>Circulating (Recycle) flow rate</td>
<td>0.087</td>
<td>$\frac{L}{min}$</td>
</tr>
<tr>
<td>4</td>
<td>$F_{d}$</td>
<td>Dye Tank Flowrate</td>
<td>0.335</td>
<td>$\frac{L}{min}$</td>
</tr>
<tr>
<td>5</td>
<td>$F_{1}$</td>
<td>$F_{in} + F_{recycle} + F_{d}$</td>
<td>-</td>
<td>$\frac{L}{min}$</td>
</tr>
<tr>
<td>6</td>
<td>$F_{2}$</td>
<td>$F_{1} + F_{d}$</td>
<td>-</td>
<td>$\frac{L}{min}$</td>
</tr>
<tr>
<td>7</td>
<td>$F_{3}$</td>
<td>$F_{out} + F_{recycle}$</td>
<td>-</td>
<td>$\frac{L}{min}$</td>
</tr>
<tr>
<td>8</td>
<td>$T_{in}$</td>
<td>Feed temperature</td>
<td>26.0</td>
<td>°C</td>
</tr>
<tr>
<td>9</td>
<td>$T_{1}$</td>
<td>Heated Tank-1 temperature</td>
<td>34.0</td>
<td>°C</td>
</tr>
<tr>
<td>10</td>
<td>$T_{2}$</td>
<td>Heated Tank-2 temperature</td>
<td>44.0</td>
<td>°C</td>
</tr>
<tr>
<td>11</td>
<td>$T_{3}$</td>
<td>Heated Tank-3 temperature</td>
<td>52.0</td>
<td>°C</td>
</tr>
<tr>
<td>12</td>
<td>$Q_{1}$</td>
<td>Heated Tank-1 Steam Flowrate</td>
<td>237.08</td>
<td>$\frac{KJ}{L}$</td>
</tr>
<tr>
<td>13</td>
<td>$Q_{2}$</td>
<td>Heated Tank-2 Steam Flowrate</td>
<td>432.02</td>
<td>$\frac{KJ}{L}$</td>
</tr>
<tr>
<td>14</td>
<td>$Q_{3}$</td>
<td>Heated Tank-3 Steam Flowrate</td>
<td>453.30</td>
<td>$\frac{KJ}{L}$</td>
</tr>
<tr>
<td>15</td>
<td>$h_{3}$</td>
<td>Heated Tank-3 water level</td>
<td>0.41</td>
<td>meter</td>
</tr>
<tr>
<td>16</td>
<td>$P$</td>
<td>Steam Operating pressure</td>
<td>550.0</td>
<td>Kpa</td>
</tr>
<tr>
<td>17</td>
<td>$C_{in}$</td>
<td>Feed Concentration flow rate</td>
<td>0.018056</td>
<td>$\frac{Kg}{L}$</td>
</tr>
<tr>
<td>18</td>
<td>$C_{1}$</td>
<td>Heated Tank-1 Concentration</td>
<td>0.00034</td>
<td>$\frac{Kg}{L}$</td>
</tr>
<tr>
<td>19</td>
<td>$C_{2}$</td>
<td>Heated Tank-2 Concentration</td>
<td>0.00067</td>
<td>$\frac{Kg}{L}$</td>
</tr>
<tr>
<td>20</td>
<td>$C_{3}$</td>
<td>Heated Tank-3 Concentration</td>
<td>0.00097</td>
<td>$\frac{Kg}{L}$</td>
</tr>
</tbody>
</table>
Table 25: System parameters steady state values.

<table>
<thead>
<tr>
<th>S. No.</th>
<th>Parameter</th>
<th>Description</th>
<th>Value</th>
<th>Units</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>A</td>
<td>Cross-sectional area of CSTR3</td>
<td>0.1225</td>
<td>meter</td>
</tr>
<tr>
<td>2</td>
<td>V1</td>
<td>Heated Tank-1 Volume</td>
<td>55.125</td>
<td>L</td>
</tr>
<tr>
<td>3</td>
<td>V2</td>
<td>Heated Tank-2 Volume</td>
<td>55.125</td>
<td>L</td>
</tr>
<tr>
<td>4</td>
<td>$\rho_{pw}$</td>
<td>Feed Pure water density</td>
<td>1.0</td>
<td>Kg/L</td>
</tr>
<tr>
<td>5</td>
<td>$\rho_{sw}$</td>
<td>Feed salty water density from dye tank</td>
<td>1.0181</td>
<td>Kg/L</td>
</tr>
<tr>
<td>6</td>
<td>$\rho_{sw1}$</td>
<td>Salty water density in Heated Tank-1</td>
<td>-</td>
<td>Kg/L</td>
</tr>
<tr>
<td>7</td>
<td>$\rho_{sw2}$</td>
<td>Salty water density in Heated Tank-2</td>
<td>-</td>
<td>Kg/L</td>
</tr>
<tr>
<td>8</td>
<td>$\rho_{sw3}$</td>
<td>Salty water density in Heated Tank-3</td>
<td>-</td>
<td>Kg/L</td>
</tr>
<tr>
<td>9</td>
<td>$C_p$</td>
<td>Heat capacity of the liquid</td>
<td>4.18</td>
<td>kJ.(kg.°C$^{-1}$)</td>
</tr>
<tr>
<td>10</td>
<td>$\lambda_s$</td>
<td>Latent heat of steam</td>
<td>2107.4</td>
<td>kW.(kg.m$^{-1}$)$^{-1}$</td>
</tr>
<tr>
<td>11</td>
<td>U</td>
<td>Overall heat transfer coefficient</td>
<td>3.376</td>
<td>kW.K$^{-1}$</td>
</tr>
</tbody>
</table>
Appendix G

G.1 Process Model Equations

G.1.1 Heated Tank-3 process liquid mass balance

\[ \frac{dh}{dt} = \frac{1}{\rho_{sw3} \cdot A_3} \left[ \rho_{sw2} \cdot F_2 + \rho_{sw} \cdot \frac{F_d}{3} - \rho_{sw3} \cdot F_3 \right] \quad (G - 1) \]

G.1.2 Process liquid energy balance

G.1.2.1 Heated Tank-1 energy balance

\[ \frac{dT_1}{dt} = \frac{1}{\rho_{sw1} C_p \cdot V_1} \left[ \rho_{pw} C_p \cdot F_{in} T_{in} + \rho_{sw3} C_p \cdot F_{recycle} T_{out} + \rho_{sw} C_p \cdot \frac{F_d}{3} T_{in} \right. \\
\left. \quad - \rho_{sw1} C_p \cdot F_1 T_1 + Q - UA \cdot (T_1 - T_{atm}) \right] \quad (G - 2) \]

G.1.2.2 Heated Tank-2 energy balance

\[ \frac{dT_2}{dt} = \frac{1}{\rho_{sw2} C_p \cdot V_2} \left[ \rho_{sw1} C_p \cdot F_1 T_1 + \rho_{sw} C_p \cdot \frac{F_d}{3} T_{in} - \rho_{sw2} C_p \cdot F_2 T_2 + Q \right. \\
\left. \quad - UA \cdot (T_2 - T_{atm}) \right] \quad (G - 3) \]

G.1.2.3 Heated Tank-3 energy balance

\[ \frac{dT_{out}}{dt} = \frac{1}{\rho_{sw3} C_p \cdot h \cdot A_3} \left[ \rho_{sw2} C_p \cdot F_2 T_2 + \rho_{sw} C_p \cdot \frac{F_d}{3} T_{in} \right. \\
\left. \quad - \rho_{sw3} C_p \cdot F_{out} T_{out} + Q - UA \cdot (T_{out} - T_{atm}) \right. \\
\left. \quad - T_{out} \left( \rho_{sw2} \cdot F_2 + \rho_{sw} \cdot \frac{F_d}{3} - \rho_{sw3} \cdot F_3 \right) \right] \quad (G - 4) \]
G.1.3 Process liquid component balance

G.1.3.1 Heated Tank-1 component balance

\[
\frac{dC_1}{dt} = \frac{1}{V_1} \left[ \frac{F_d}{3} \cdot C_d - F_1 \cdot C_1 + F_{recycle} \cdot C_{out} \right] \tag{G - 5}
\]

G.1.3.2 Heated Tank-2 component balance

\[
\frac{dC_2}{dt} = \frac{1}{V_2} \left[ F_1 \cdot C_1 + \frac{F_d}{3} \cdot C_d - F_2 \cdot C_2 \right] \tag{G - 6}
\]

G.1.3.3 Heated Tank-3 component balance

\[
\frac{dC_{out}}{dt} = \frac{1}{h \cdot A_3} \left[ F_2 \cdot C_2 + \frac{F_d}{3} \cdot C_d - F_{out} \cdot C_{out} \right. \\
\left. - C_{out} \left( \rho_{sw2} \cdot F_2 + \rho_{sw} \cdot \frac{F_d}{3} - \rho_{sw3} \cdot F_{out} \right) \right] \tag{G - 7}
\]

G.1.4 Process liquid flowrates

\[
F_1 = F_1 + \frac{F_d}{3} + F_{recycle} \tag{G - 8}
\]

\[
F_2 = F_1 + \frac{F_d}{3} \tag{G - 9}
\]

\[
F_3 = F_{out} + F_{recycle} \tag{G - 10}
\]
Appendix H: Dynamic Response Computation Using \textit{ode45}

\begin{verbatim}
% % Function File
% % Heated Tanks Simulation Project - Murdoch University
% % Open loop and Dynamic Run
% % Written by: Dawood Al-Kahali - 2017/2018
% % Supervisor: Professor Parisa A. Bahri

function xdot=HeatedTanks_DynamRun_ODE(t, x)

% Global Variables
global ppw psw psw1 psw2 psw3 Cp Ua Ua2 Ua3 A3 V1 V2
global Fin F1 F2 Fout Frecycle F3 Fd Q1 Q2 Q3 Tin T1 T2 Tout h Cin
global Fins F1s F2s Fouts Frecycles F3s Fds Qs Q2s Q3s Tins T1s T2s Touts Tatm hs Tsts Cin
global StepFin StepFout StepFrecycle StepFd StepQ1 StepQ2 StepQ3
StepCin hSP_Step UpTime DownTime
global C1 C2 C1s C2s Couts
global StepC1 StepC2 StepCout
global Cond1 Cond2 Cond3 Cond1s Cond2s Cond3s StepCond1 StepCond2 StepCond3

% Global Variables
Cond1s= (x(5)*1000000)/0.65;
Cond2s= (x(6)*1000000)/0.65;
Cond3s= (x(7)*1000000)/0.65;

% If condition
if t <= 20
    Fin  = Fins;
    Fout= Fouts;
    Frecycle= Frecycles;
    F3= F3s;
    F1= F1s;
    F2= F2s;
    Fd  = Fds;
    Tin = Tins;
    Q1 = Qs;
    Q2 = Q2s;
    Q3 = Q3s;
    Cin = Cins;
    h = hs;
    T1 =T1s;
    T2 =T2s;
    Tout =Touts;
    C1 =C1s;
    C2 =C2s;
    Cout =Couts;
    Cond1 =Cond1s;
    Cond2 =Cond2s;
    Cond3 =Cond3s;
else
    Fin  = Fins + StepFin;
    Fout= Fouts + StepFout;
    Frecycle= Frecycles + StepFrecycle;
    F3 = F3s + StepFout + StepFrecycle;
    F1 = F1s + StepFin + StepFd/3 + StepFrecycle;
end
\end{verbatim}
\[ F_2 = F_{2s} + \text{StepFin} + \text{StepFrecycle} + \text{StepFd}/3 + \text{StepFd}/3; \]
\[ F_d = F_{ds} + \text{StepFd}; \]
\[ T_{in} = T_{ins} + \text{StepTin}; \]
\[ Q_1 = Q_{s} + \text{StepQ1}; \]
\[ Q_2 = Q_{2s} + \text{StepQ2}; \]
\[ Q_3 = Q_{3s} + \text{StepQ3}; \]
\[ C_{in} = C_{ins} + \text{StepCin}; \]
\[ h = h_{s} + \text{Steph}; \]
\[ T_1 = T_{1s} + \text{StepT1}; \]
\[ T_2 = T_{2s} + \text{StepT2}; \]
\[ T_{out} = T_{outs} + \text{StepTout}; \]
\[ C_1 = C_{1s} + \text{StepC1}; \]
\[ C_2 = C_{2s} + \text{StepC2}; \]
\[ C_{out} = C_{outs} + \text{StepCout}; \]
\[ \text{Cond1} = \text{Cond1s} + \text{StepCond1}; \]
\[ \text{Cond2} = \text{Cond2s} + \text{StepCond2}; \]
\[ \text{Cond3} = \text{Cond3s} + \text{StepCond3}; \]

end

\% Model Equations - 1 : " LEVEL 
\[ \text{xdot}(1) = (1/psw3*A3)*(psw2*F2 + psw*(Fd/3) - psw3*F3); \]

\% Model Equations - 2 : " TEMPERATURE 
\[ \text{xdot}(2) = (1/psw1*Cp*V1) * (ppw*Cp*Fin*Tin + psw3*Cp*Frecycle*x(4) + psw*Cp*(F2/3)*Tin - psw1*Cp*F1*x(2) + (Q1)); \]
\[ \text{xdot}(3) = (1/psw2*Cp*V2) * (psw1*Cp*F1*x(2) + psw*Cp*(Fd/3)*Tin - psw2*Cp*F2*x(3) + (Q2)); \]
\[ \text{xdot}(4) = (1/psw3*Cp*A3*x(1)) * (psw2*Cp*F2*x(3) + psw*Cp*(Fd/3)*Tin - psw3*Cp*Fout*x(4) + (Q3)); \]

\% Model Equations - 3 : " CONCENTRATION 
\[ \text{xdot}(5) = (1/V1) * ((Fd/3)*Cin - F1*x(5) + Frecycle*x(7)); \]
\[ \text{xdot}(6) = (1/V2) * (F1*x(5) + (Fd/3)*Cin - F2*x(6)); \]
\[ \text{xdot}(7) = (1/(A3*x(1)))* (F2*x(6) + (Fd/3)*Cin - Fout*x(7)); \]

\[ \text{xdot} = [\text{xdot}(1) \text{ xdot}(2) \text{ xdot}(3) \text{ xdot}(4) \text{ xdot}(5) \text{ xdot}(6) \text{ xdot}(7)]; \]

end

\% Main File
clc;
clear;

\% Global Variables
\% pps psw ps1 ps2 ps3 Cp Ua Ua2 Ua3 A3 V1 V2
\% Fin F1 F2 Fout Frecycle F3 Fd Q1 Q2 Q3 Tin T1 T2 Tout h
\% Cin
\% Fins F1s F2s Fouts Frecycles F3s Fds Qs Q2s Q3s Tins T1s T2s Touts
\% hs TsTs Tats Cins
\% StepFin StepFout StepFrecycle StepFd StepTin StepQ1 StepQ2 StepQ3
\% StepCin StepUpTime StepDownTime
\% C1 C2 Cout C1s C2s Cout
\% Cond1s Cond2s Cond3s Cond1 Cond2 Cond3 StepCond1 StepCond2
\% StepCond3

\% Constants
\% V1=0.055125; \% [m^3] Volume of CSTR1 tank
\% V2=0.055125; \% [m^3] Volume of CSTR2 tank
\% V1=55.125; \% [L] Volume of CSTR1 tank
% V2=55.125; % [L] Volume of CSTR2 tank
A3=0.1225; % [m^2] Area of the CSTR3 tank

% Densities
ppw=1; % kg/L (Pure Water density)
psw=1.01806; % kg/L (Salty Water density)
psw1=1.01200; % kg/L (Salty Water density) in CSTR1
psw2=1.01000; % kg/L (Salty Water density) in CSTR2
psw3=1.01400; % kg/L (Salty Water density) in CSTR3
Cp = 4.189; % kJ/(Kg*C) % J/kgK (Heat Capacity)

%==========================================================================
% Initial Conditions and Steady State Values
hs=0.41; % height of the tank at steady state
Fins = 5.963; % L/min Steady state Cold Stream water InFlowrate
Fouts= 6.27370005;
Frecycles =0.087; % L/min Steady state Recycle Flowrate
Fds = 0.335; % L/min Steady state Dye Flowrate
Tins=26; % Degree C, Steady state Inflow Temperature
T1s=34.448; % Degree C, Steady state CSTR1 Temperature
T2s=47.814; % Degree C, Steady state CSTR2 Temperature
Touts=61.09; % Degree C, Steady state Product stream Temperature
Tsts=151.8; % Degree C, Steady State Steam Temperature

% Tin=Tins; % Degree C, Inflow Temperature
Tatm=25;

% Cins = 18.0555; % g/L
Cins = 0.018056; % Kg/L
% Cin=Cins;
% C1s=0.000341;
% C2s=0.000665;
% Cout5=0.00097;

Qsteam1=0.1125; % Steady State Steam Flowrate1 Kj/min in CSTR1
Qsteam2=0.205; % Steady State Steam Flowrate2 Kj/min in CSTR2
Qsteam3=0.2151;
Lemdas=2107.4; % Latent Heat

Qs = Qsteam1 * Lemdas;
Q2s= Qsteam2 * Lemdas;
Q3s= Qsteam3 * Lemdas;

Ua=3.376;
Ua2=3.376;
Ua3=3.376;
F3s= Fouts + Frecycles;
F1s = Fins + (Fds/3) + Frecycles ;
F2s = F1s + (Fds/3);

%==========================================================================
% % Step Conditions%
% % Step and Disturbance changes
% % Flowsrates Step Change
StepFin=0; % Step change of Fin by -20 L/min = -0.020 m^3/min
StepFout =0; %0.00004; %Product flow rate
% StepFrecycle=0.1*Frecycles;
StepFrecycle=0;
StepFd=0;
% Temperature Step Change
StepTin=5;
StepT1=0;
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StepT2 = 0;
StepTout = 0;
% Steam Flowrates Step Change
StepQ1 = 0;
StepQ2 = 0;
StepQ3 = 0;
Steph = 0;
% Concentration Step Change
% StepCin = 0.05 * Cin;
StepCin = 0;
StepC1 = 0;
StepC2 = 0;
StepCout = 0;
%==========================================================================
% ODE Conditions
% t0 = 0;                  % Initial Time
% tf = TotalRunTime;
tf = 100;                  % Final Time
tspan = [t0 tf];           % Time Span
x0 = [hs T1s T2s Touts 0.000341 0.00065635 0.000977]';
options = [] ;             % Not Used
[t, x] = ode45('HeatedTanks_DynaminRun_ODE', tspan, x0, options);  % ODE Solver
%==========================================================================
% If condition
%     if t(i) < StepUpTime || t(i) > StepDownTime
st = max(size(t));
for i = 1 : st
    if t(i) <= 20
        Fin(i) = Fins;
        Fout(i) = Fouts;
        Frecycle(i) = Frecycles;
        F3(i) = F3s;
        F1(i) = F1s;
        F2(i) = F2s;
        Fd(i) = Fds;
        Tin(i) = Tins;
        Q1(i) = Qs;
        Q2(i) = Q2s;
        Q3(i) = Q3s;
        Cin(i) = Cin;
        h(i) = hs;
        T1(i) = T1s;
        T2(i) = T2s;
        Tout(i) = Touts;
        C1(i) = C1s;
        C2(i) = C2s;
        Cout(i) = Cout;
        Cond1(i) = Cond1s;
        Cond2(i) = Cond2s;
        Cond3(i) = Cond3s;
    else
        Fin(i) = Fins + StepFin;
        Fout(i) = Fouts + StepFout;
        Frecycle(i) = Frecycles + StepFrecycle;
        F3(i) = F3s + StepF3 + StepFrecycle;
        F1(i) = F1s + StepFin + StepFd/3 + StepFrecycle;
        F2(i) = F2s + StepFin + StepFrecycle + StepFd/3 + StepFd/3;
        Fd(i) = Fds + StepFd;
        Tin(i) = Tins + StepTin;
        Q1(i) = Qs + StepQ1;
    end
end
\[Q2(i) = Q2s + \text{StepQ2};\]
\[Q3(i) = Q3s + \text{StepQ3};\]
\[Cin(i) = Cins + \text{StepCin};\]
\[h(i) = hs + \text{Steph};\]
\[T1(i) = T1s + \text{StepT1};\]
\[T2(i) = T2s + \text{StepT2};\]
\[Tout(i) = Touts + \text{StepTout};\]
\[C1(i) = C1s + \text{StepC1};\]
\[C2(i) = C2s + \text{StepC2};\]
\[Cout(i) = Cout + \text{StepCout};\]
\[\text{Cond1}(i) = \text{Cond1s} + \text{StepCond1};\]
\[\text{Cond2}(i) = \text{Cond2s} + \text{StepCond2};\]
\[\text{Cond3}(i) = \text{Cond3s} + \text{StepCond3};\]
\end

%==========================================================================
% Extract ODE Solution
% Prepare for plotting
h=x(:,1);
T1=x(:,2);
T2=x(:,3);
Tout=x(:,4);
C1=x(:,5);
C2=x(:,6);
Cout=x(:,7);

%==========================================================================
figure(3)
subplot(4,4,1), plot(t, Fin)
% title('Disturbance Variables')
title('Inlet Flowrate')
xlabel('t, min')
ylabel('Fin, (L/m)')
axis([0 40 346 350])
subplot(4,4,5), plot(t, Frecycle)
title('Recycle Flowrate')
xlabel('t, min')
ylabel('Frecycle, (L/m)')
axis([0 40 0.35 0.371])
subplot(4,4,9), plot(t, Tin)
title('Feed temp')
xlabel('t, min')
ylabel('Tin, (Degree C)')
subplot(4,4,13), plot(t, Cin, 'r')
title('Inlet Concentration')
xlabel('t, min')
ylabel('Cin (Kg/L)')
subplot(4,4,2), plot(t, F1)
% title('Associated Flowrates')
title('F1 = Fin + Frecycle + Fd/3')
xlabel('t, min')
ylabel('F1, (L/m)')
subplot(4,4,6), plot(t, F2)
title('F2 = F1 + Fd/3')
xlabel('t, min')
ylabel('F2, (L/m)')
subplot(4,4,10), plot(t, F3)
title('Drain Flowrate (F3 = Fout + Frecycle)')
xlabel('t, min')
ylabel('F3, (L/m)')

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Appendix I: MATLAB System Identification Result

% subplot(4,4,3)  % subplot(4,4,16), plot(t, h)  
% title('CSTR3 Water Level')  % axis([t0 tf min(h)-0.01 max(h)+0.01])  % xlabel('t, min')  % ylabel('h, (meter)')
% subplot(4,4,1), plot(t, T1)  % title('CSTR1 Temperature')  % axis([t0 tf min(T1)-5 max(T1)+5])  % xlabel('t, min')  % ylabel('T1, (Degree C)')
% subplot(4,4,7), plot(t, T2)  % title('CSTR2 Temperature')  % axis([t0 tf min(T2)-5 max(T2)+5])  % xlabel('t, min')  % ylabel('T2, (Degree C)')
% subplot(4,4,11), plot(t, Tout)  % title('CSTR3 Temperature')  % axis([t0 tf min(Tout)-5 max(Tout)+5])  % xlabel('t, min')  % ylabel('Tout, (Degree C)')
% subplot(4,4,4), plot(t, C1)  % title('CSTR1 Concentration')  % axis([t0 tf min(C1)-0.0005 max(C1)+0.0005])  % xlabel('t, min')  % ylabel('C1 (Kg/L)')
% subplot(4,4,8), plot(t, C2)  % title('CSTR2 Concentration')  % axis([t0 tf min(C2)-0.0005 max(C2)+0.0005])  % xlabel('t, min')  % ylabel('C2 (Kg/L)')
% subplot(4,4,12), plot(t, Cout)  % title('CSTR3 Concentration')  % axis([t0 tf min(Cout)-0.0005 max(Cout)+0.0005])  % xlabel('t, min')  % ylabel('Cout (Kg/L)')
Figure 65: C1 MATLAB System Identification Result

Figure 66: C2 MATLAB System Identification Result
Appendix J: Performance Analysis

Appendix J-1: PI performance analysis: Set point changes

Figure 67: T1 Setpoint Tracking -5 Degree C PI
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Figure 68: T2 Setpoint Tracking +5 Degree C  PI

Figure 69: Tout Setpoint Tracking +5 Degree C  PI
Appendices

Figure 70: Tout Setpoint Tracking -5 Degree C  PI

Figure 71: C1 Setpoint Tracking +1 %  PI
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Figure 72: C1 Setpoint Tracking -1 % PI

Figure 73: C2 Setpoint Tracking +1 % PI
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Figure 74: C2 Setpoint Tracking -1 % PI

Figure 75: Cout Setpoint Tracking +1 % PI
Appendix J-2: PI performance analysis: Disturbance Rejection

Figure 76: Fin disturbance change +5% PI
Figure 77: Frecycle disturbance change +0.05 L/min  PI
Figure 78: Freecycle disturbance change -0.05 L/min PI
Figure 79: Cin disturbance change +0.005 Kg/L PI
Figure 80: Cin disturbance change -0.005 Kg/L PI
Appendix J-3: GMC Performance Analysis: Set point Changes

Figure 81: T1 Setpoint Tracking +5 Degree C  GMC

Figure 82: T2 Setpoint Tracking -5 Degree C  GMC
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Figure 83: Tout Setpoint Tracking +5 Degree C  GMC

Figure 84: Tout Setpoint Tracking -5 Degree C  GMC
Appendix J-4: GMC Performance Analysis: Disturbance Rejection

Figure 85: Freecycle disturbance change +0.05 L/min GMC
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Figure 86: Fin disturbance change +5% GMC

Appendix K: ISE, IAE, and ITAE (PI and GMC)
### Appendix K-1: Set-point Changes; ISE, IAE, and ITAE (PI and GMC)

Table 26: Tout Set-point Change; ISE, IAE, and ITAE (PI and GMC)

<table>
<thead>
<tr>
<th>Tout ±5°C</th>
<th>ISE</th>
<th>GMC</th>
<th>ISE</th>
<th>GMC</th>
<th>ISE</th>
<th>GMC</th>
</tr>
</thead>
<tbody>
<tr>
<td>h</td>
<td>0</td>
<td>0</td>
<td>0.0011</td>
<td>0</td>
<td>0.0021</td>
<td>0</td>
</tr>
<tr>
<td>T1</td>
<td>0.0019</td>
<td>0</td>
<td>0.4243</td>
<td>0</td>
<td>0.4656</td>
<td>0</td>
</tr>
<tr>
<td>T2</td>
<td>0.0005</td>
<td>0</td>
<td>0.2216</td>
<td>0</td>
<td>0.2472</td>
<td>0</td>
</tr>
<tr>
<td>Tout</td>
<td>0.0013</td>
<td>0.0379</td>
<td>0.1086</td>
<td>0.7611</td>
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Table 27: C1 Set-point Change; ISE, IAE, and ITAE (PI and GMC)

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### Table 28: C2 Set-point Change ; ISE, IAE, and ITAE (PI and GMC).

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### Table 29: Cout Set-point Change ; ISE, IAE, and ITAE (PI and GMC).

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### Table 30: Cin Disturbance Changes; ISE, IAE, and ITAE (PI and GMC)

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Appendix L: Calculated Conductivity

Conductivity in each heated tank has been calculated based on the measured concentrations using the following:

1. When Concentrations were at steady state.

![Graphs showing CSTR1, CSTR2, CSTR3 Concentration and Conductivity over time at steady state.]

Figure 87: Calculated Conductivities when Concentrations at Steady State.
2. When disturbance changed has been introduced (C_{in} + 0.005 Kg/L)

Figure 88: Conductivities when disturbance change (C_{in} +0.005 Kg/L)
3. When disturbance changed has been introduced (C_{in} -0.005 \text{ Kg/L})

![Figure 89: Conductivities when disturbance change (C_{in} -0.005 \text{ Kg/L})](image)
Figure 90: Heated Tanks + Dye Tank (Real Photo).