Linear Model Predictive Control of a Debutanizer Column (Simulation Work)

by

Mirralaxmi R Chandra Mohan
15109

Dissertation submitted in partial fulfilment of
the requirements for the
Bachelor of Engineering (Hons)
(Chemical Engineering)

MAY 2014

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CERTIFICATION OF APPROVAL

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A project dissertation submitted to the
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(CHEMICAL ENGINEERING)

Approved by,

____

(Mr. Nasser B M Ramli)

UNIVERSITI TEKNOLOGI PETRONAS
TRONOH, PERAK
May 2014

CERTIFICATION OF ORIGINALITY

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

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MIRRALAXMI R CHANDRA MOHAN

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CHEMICAL ENGINEERING

ABSTRACT

This dissertation has been prepared in order to fulfill the partial requirement of a final year chemical engineering student in Universiti Teknologi PETRONAS (UTP). It is submitted to the university as a requirement of the Final Year Project II.

Linear model predictive control studies have been an interesting field of research in the scope of chemical engineering. This research has been done in order to benefit the operations of PETRONAS Penapisan Terengganu Sdn. Bhd., PP(T)SB. The C-110 Debutanizer column at the plant is not achieving its desired output which is 30% propane and 70% butane. Therefore this study is conducted in order to obtain the optimum tuning parameters for Model Predictive Control, MPC controllers that can help achieve the desired output at the plant operation.

The C-110 Debutanizer column has been simulated using HYSYS^{TM} software according to real plant data. By conducting an open loop step test, relevant data were collected in order to proceed to MATLAB programming. By using the IDENT System Identification tool available in MATLAB, simple programming and coding were done to obtain necessary input information for the MPC controllers. MPC controllers with different tuning settings were tested in the HYSYS^{TM} environment and the best tuning method is suggested to PP(T)SB to enhance their plant operation.

Chapter 1 provides a short introduction on the background of the project. The problem statements of this project will be well discussed in accordance to its objectives. Chapter 2 gives a detailed literature review on the mentioned topic of research. In this chapter, the concept and basic understanding of the project is shown.

In Chapter 3, the research methodology and project activities are mentioned. The milestones of this project are also presented. Chapter 4 shows the results gained in this project. The relevancy of this project to its objectives and its probable future works are also discussed. Chapter 5 discusses the conclusion and recommendations for this project.

The findings of this project will help the operations at PETRONAS Penapisan Terengganu Sdn. Bhd., PP(T)SB in order to achieve the desired output of the C-110 Debutanizer column.

ACKNOWLEDGEMENT

First and foremost, I would like to thank my supervisor, Mr. Nasser B M Ramli for all his support towards me in completing this project. Without his coaching and supervision, I would not have made it this far. Special thanks to PETRONAS Penapisan Terengganu Sdn. Bhd., PP(T)SB for providing all the necessary information related to my project.

I would also like to thank my university, Universiti Teknologi PETRONAS, UTP for giving me an opportunity to excel as a Final Year student.

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TABLE OF CONTENTS

LIST OF FIGURES

LIST OF TABLES

CHAPTER 1

PROJECT BACKGROUND

1.1 Background Study

This study is being conducted to benefit the operations at PETRONAS Penapisan (Terengganu) Sdn. Bhd., PP(T)SB, one of Malaysia's crude oil refinery. A picture of the refinery is shown in Figure 1. This refinery produces almost 49 000 barrels of Malaysian light, sweet crude oil on a daily basis. After a recent addition of a Condensate Splitter Unit (KR-2A), the plant also produces 74 300 barrels of naphtha condensates per day. This product is used as the main feed stream in aromatics plants located nearby. This plant receives feedstock mainly from Bintulu and Terengganu with contains a low amount of sulphur.

The main focus of this project will be the Debutanizer column in Crude Distillation Unit (CDU) of Kerteh Refinery-1 (KR-1). In order to achieve high purity of product formation, a deep study will be conducted in order to maximize the product (propane) and minimize the byproduct (C4+) at this column. The flow diagram of the Crude Distillation Unit (CDU) is shown in Figure 2. The outlet of the Debutanizer is Liquefied Petroleum Gas (LPG) and Light Naphtha.

Figure 1.1: PETRONAS Penapisan (Terengganu) Sdn. Bhd

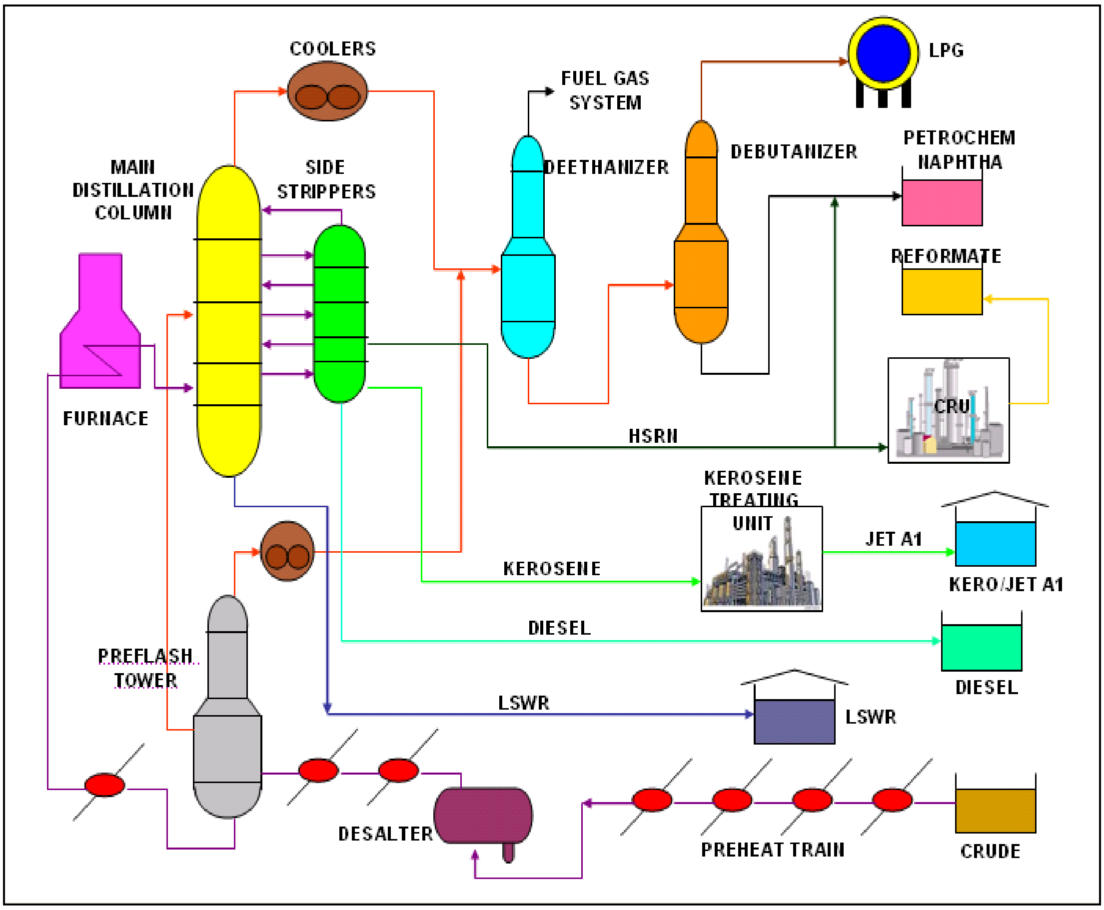

Figure 1.2: Process Flow Diagram (PFD) of Crude Distillation Unit (CDU)

1.2 Problem Statement

As explained above, the Debutanizer column is crucial in the formation of LPG. In this scenario, LPG contains 30% of propane (C3+) and 70% of butane (C4+). However the operators of the plant always face problems as the desired product composition is hard to be obtained. The outlet of the debutanizer has lower amount of propane which forces the plant operators to import propane (C3+) from adjacent gas processing plant, PETRONAS Gas Berhad (PGB) in Kerteh. This step creates additional cost for the management. Therefore it is crucial to conduct a study in order to improve the product composition of the debutanizer.

1.3 Objective

The main focus of this project is to come up with a control strategy that can help to maximize the product output at the Debutanizer column. The desired output is 30% of propane (C3+) and 70% of butane (C4%). In order to achieve this, steady state and dynamic modelling is to be done using the HYSYS^{TM} software. Once the dynamic model is ready, a throughout analysis will be conducted as to fulfill the objectives stated below.

- To obtain maximum yield of LPG from the Crude Distillation Unit (CDU)
- To gain optimum performance of the Debutanizer column with desired output composition
- To analyze the process variables of each controller by using different tuning relations.

1.4 Scope of Study

The scope of this study revolves around the C-110 Debutanizer column. The most important parameter that needs to be studied is the outlet composition of this Debutanizer. From the element of needed simulation software, in this project, the HYSYS^{TM} and MATLAB software will be used. Real plant data will be used to conduct this project.

1.5 Relevancy and feasibility of project

This project will be relevant to the plant operation at PETRONAS Penapisan (Terengganu) Sdn. Bhd., PP(T)SB as to maximize the outlet composition of the C-110 Debutanizer column. The project findings will be helpful in tuning the controllers around the Debutanizer column so that the desired output can be achieved.

This project is feasible to be completed within the timeframe of 2 semesters given. The first semester focuses more on data collection and dynamic simulation with HYSYS^{TM} while the second semester focuses more on MATLAB programming.

CHAPTER 2

LITERATURE REVIEW

Important theories are explained in this chapter to create a better understanding on the project as a whole.

2.1 Model Predictive Control – MPC

According to Lawrynczuk (2007), MPC is an advanced control technique that performs better than the existing PID controllers. It takes into account of a system's input, output and controller constraints which results in a better control system. The computerized control algorithm is capable of controlling future behavior of an observed system

2.2 Steady State and Dynamic State Modelling

A linear system can be associated to a steady state model. A steady state model has variables that do not change with respect to time (A. H., Abdul Malik, 2009). Therefore, analyzing a steady state model is less complex than a dynamic model. However in real plant situations, steady state operations are not feasible due to many factors such as disturbances from environment, loss of heat to the environment, heat exchanger fouling and so on. This is why in a real plant situation; normally we use dynamic state modelling. Software like HYSYS^{TM} helps us to build a virtual plant environment that makes it easier to predict, control and analyze a real plant environment. At first, a steady state model is build and made to converge. Only then it is switched to a dynamic mode that allows various control strategies to be applied to the system. How the system reacts to various control strategies will be studied by an engineer.

2.3 Debutanizer column

A Debutanizer column is basically a distillation column that has a reboiler and a condenser. It is used to separate components of crude oil. In this case, the debutanizer separates propane (C3+) from butane (C4+). The highly vaporized component will exit from the top outlet of the column while the heavier hydrocarbon will be collected at the bottom of the column (Training Manual For Crude Distillation Unit of PP(T)SB).

2.4 PID Controller

PID controllers are an essential part of distributed control system (K. J., Astrom, 2002). It is a combination of 3 different actions which are proportional, integral and derivative action. Almost 95% of controllers in the industries are PID controllers. After the evolution of microprocessors, PID controllers are now able to perform additional tasks like automatic tuning, gain scheduling, and continuous adaptation.

2.5 Liquefied Petroleum Gas (LPG)

Liquefied Petroleum Gas (LPG) is an important energy source in Malaysia (GAS MALAYSIA). It is often used for cooking purposes. LPG is a good alternative as compared to wood and fossil fuels due to its clean burning properties. Therefore using LPG is more environment friendly.

2.6 Related researches done previously

Many researches have been conducted in the field of Model Predictive Control which involves simulation of a distillation column. Kanthasamy (2009) has conducted a study to control binary distillation column by developing 2 nonlinear model predictive control (NMPC) systems using Hammerstein model and nonlinear autoregressive model with exogenous input (NARX). A pilot plant distillation column separating methanol and water was used for this project. Using MATLAB simulation, closed loop control studies were done to verify the behavior of the NMPC techniques in disturbance rejection and set-point tracking. Results from the MATLAB programming shows that the performance of Hammerstein NMPC was superior to NARX NMPC in controlling the distillation column.

In another paper presented by Mishra et al., (2010), the effects of tuning parameter of a binary distillation column model were studied by using MPC. Wood and Berry 2x2 function was used for the simulation. The response of the model with and without disturbance was studied. The techniques of removing the ringing effect in MPC manipulated variables were also studied. It is found that at certain tuning parameter the performance of control system is better than any classical control system. Removal of ringing effect can be done by increasing the manipulated input weights. Figure 2.1 and 2.2 shows the results of this project.

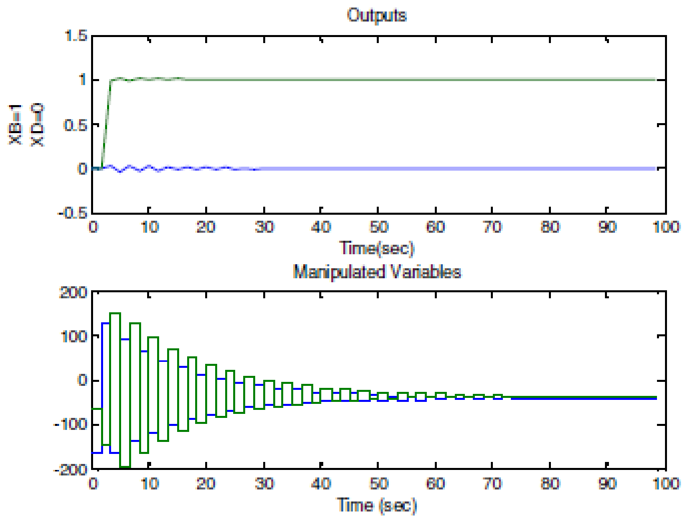

Figure 2.1: Response of Wood Berry without Disturbance

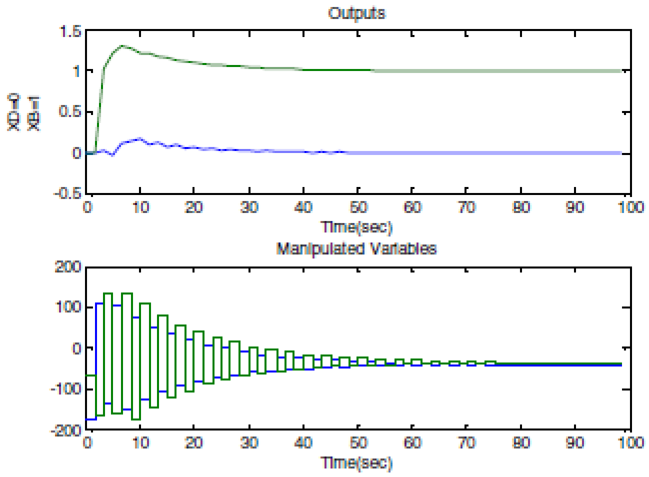

Figure 2.2: Response of Wood Berry with Disturbance

Besides that, Hesam et al., (2010) worked on a distillation column simulated in HYSYS and the data obtained were studied in MATLAB for identification and control purpose. Linear model structure based on ARX (Autoregressive with external input) and nonlinear model structure based on neural network were used in this study. Two linear and nonlinear model predictive controllers are applied for control goals. General predict control (GPC) and nonlinear predict control (NPC) were compared.

Real-time identification based recursive parameter estimation is used. The results show that adaptive GPC based recursive parameter estimation is successful and has excellent capabilities in real-time identification and control.

Slightly similar to the current project, Ashraf et al., (2010) worked on on-line tuning strategy for linear Model Predictive Control (MPC) algorithms. The tuning strategy is based on the linear approximation between the closed-loop predicted output and the MPC tuning parameters. By direct utilization of the sensitivity expressions for the closed-loop response with respect to the MPC tuning parameters, new values of the tuning parameters was found to steer the MPC feedback response inside predefined time-domain performance specifications. Effectiveness of the proposed strategy is tested on a linear model for a three-product distillation column and a non-linear model for a CSTR. The obtained results showed successful implementation of the tuning algorithm despite the presence of model-plant mismatch, non-linearity and instability in some of the cases. The simulation also revealed that, in some cases, the constraints on the manipulated variable and its moves may prevent the tuning algorithm from improving the closed-loop performance.

However, this current project will have its own unique features as compared to all the previous studies done in the field of MPC related to distillation column. In this project, open loop step test data obtained from HYSYS simulation will be interpreted using MATLAB programming and linear approximation. The system identification tool, IDENT, available in MATLAB will be used to further process the findings in order to obtained relevant input data for MPC controllers that will specifically benefit the operations of PETRONAS Penapisan Terengganu Sdn. Bhd., PP(T)SB.

CHAPTER 3

METHODOLOGY

3.1 Research Methodology and Project Activities

This project is divided into Final Year Project 1 and Final Year Project 2. Therefore this project will be conducted continuously for two semesters. Explained below are the research methodology and project activities of this project.

- Real plant data was collected from PETRONAS Penapisan Terenganu Sdn. Bhd. All important properties like operating temperature, pressure, feed mole fraction and heat duty were obtained from the plant's process engineer.
- Using the data collected, a steady state model was constructed using the HYSYS^{TM} software.
- From the built steady state model, a dynamic model was developed using the HYSYS^{TM} software.
- While the simulation is running in dynamic mode, a step test was conducted using the open loop method.
- The data from the step test including the process variable and manipulated variable were collected to be further processed with MATLAB.
- Using the IDENT, system identification tool available in MATLAB, the data from the open loop step test were analyzed. The transfer functions for each controller was determined.
- MATLAB programming was done to obtain overall transfer function of all the controllers using the 'no constrain method' and 'quadratic programming method'.
- The response of the controllers with respect to its process variables and manipulated variables were studied.
- The process gain, time constant and time delay for each controller were obtained to be used as inputs for the MPC controllers.
- In HYSYS^{TM}, the PID controllers in the dynamic mode will be replaced with MPC controllers according to the data collected from MATLAB programming.

- MPC controller response will be studied and analyzed with respect to the ‘no constrain method’ and ‘quadratic programming method’.
- How the controller response varies between the 2 methods will be studied. Controller settings that will help optimize the Debutanizer column will be then suggested to be used in the real plant environment to further enhance production yield.

3.2 Key Milestones

The main goal of this project is to suggest to the plant operators on which controller settings will be helpful in optimizing the product output of the Debutanizer column. The optimum settings for the MPC controllers will ensure better performance of the Debutanizer column. This can be done by simulating a dynamic state model that can well represent the actual plant situation at PETRONAS Penapisan Terenganu Sdn. Bhd. which will be able to tell us which operating condition can yield the most optimum product output at the Debutanizer column.

A reliable and precise dynamic model will be able to help the process engineer to fine tune the controller settings at the plant depending on the modelling results in order to improve product composition at the Debutanizer column.

Once these modelling results can be applied at the real plant, the extra costing of importing C3+ from a nearby plant can be reduced.

3.3 Overall Gantt Chart

NO	DETAILS WEEK	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
1	Project work continues															
2	Submission of Progress Report								●							
3	Project work continues															
4	Pre - EDX											●				
5	Submission of Draft Report												●			
6	Submission of Dissertation (soft bound)													●		
7	Submission of Technical Paper													●		
8	Oral Presentation														●	
9	Submission of Project Dissertation (hard bound)															●

● Suggested milestone

Process

3.4 Specific Gantt Chart

NO	DETAILS	1	2	3	4	5	6	7	8	9	10	11	12	13	14
1	Conduct open loop step test in HYSYS™														
2	Analyse data with System Identification Tool, IDENT, MATLAB														
3	MATLAB programming and coding														
4	Analyse and interpret the findings from MATLAB														
5	Use findings for MPC controllers in HYSYS™														
6	Observe how the dynamic system responds in HYSYS™														
7	Documentation of findings														

Process

CHAPTER 4

RESULTS AND DISCUSSION

4.1 Data Collection

Table 4.1: Debutanizer column plant data

Number of tray of the column	35
Feed tray - stage number	23
Type of tray used	Valve
Column diameter	1.3m
Column length	0.61m
Type of condenser	Partial
Feed mass flowrate	44106 kg/hr
Feed temperature	113 ^{0}C
Feed pressure	823.8 kPa
Overhead vapor mass flowrate	11286 kg/hr
Overhead liquid mass flowrate	5040 kg/hr
Pressure Condenser	823.8 kPa
Pressure Reboiler	853.2 kPa

Table 4.2: Composition at Feed

Composition	Mass Fraction
Propane	0.037
i-Butane	0.093
n-Butane	0.062

i-Pentane	0.082
n-Pentane	0.110
Hypo50_13*	0.017
Hypo60_13*	0.191
Hypo70_13*	0.245
Hypo80_13*	0.063
Hypo90_13*	0.070
Hypo100_13*	0.029
Hypo110_13*	0.003
Hypo120_13*	0.001

Table 4.3: Properties of the hypothetical components

Component	Boiling Temp (^{0}C)	Critical P (kPa)	Critical T (^{0}C)	Critical Volume (m3/kgmole)	Molecular Weight	SG	Viscosity 50C (cSt)	TVP (kPa)
Hypo40_13*	38	3363	201.7	0.3171	71.34	642.2	0	0
Hypo50_13*	45	4545	221	0.2483	70.13	760.3	0.21	68.45
Hypo60_13*	55	3162	221	0.3475	85.98	666	0.21	47.81
Hypo70_13*	65	3053	232.2	0.3658	85.69	681.8	0.21	44.31
Hypo80_13*	75	3957	261.7	0.303	83.83	774.9	0.21	26.61
Hypo90_13*	85	2907	255.9	0.3983	99.02	704.7	0.21	17.2
Hypo100_13*	95	3141	274.1	0.3813	98.44	736.8	0.21	14.63
Hypo110_13*	105	3262	290	0.377	105	758.2	0.2114	8.582
Hypo120_13*	115	2739	293.4	0.4474	111.7	737.2	0.2213	6.168

4.2 HYSYS™ Modelling Results

Figure 4.1, 4.2 and 4.3 shows the results obtained from HYSYS™ modelling.

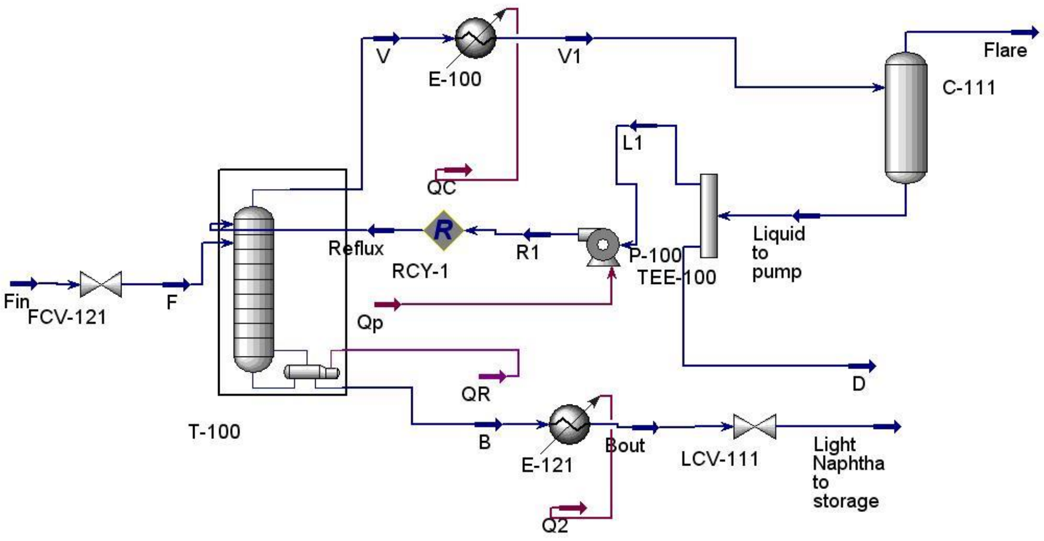

Figure 4.1: Steady State model built in HYSYS™ software

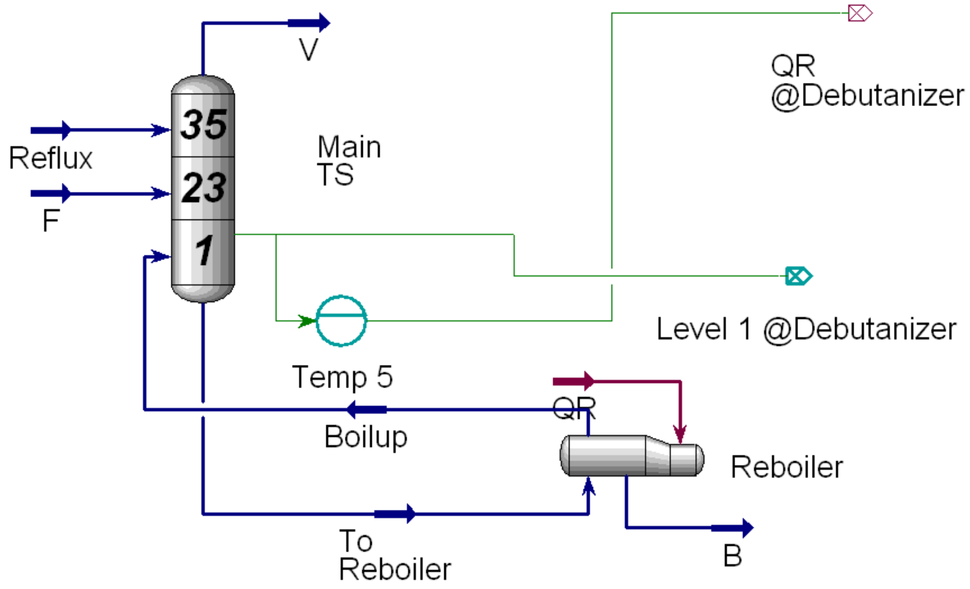

Figure 4.2: Distillation Column built in HYSYS™ software

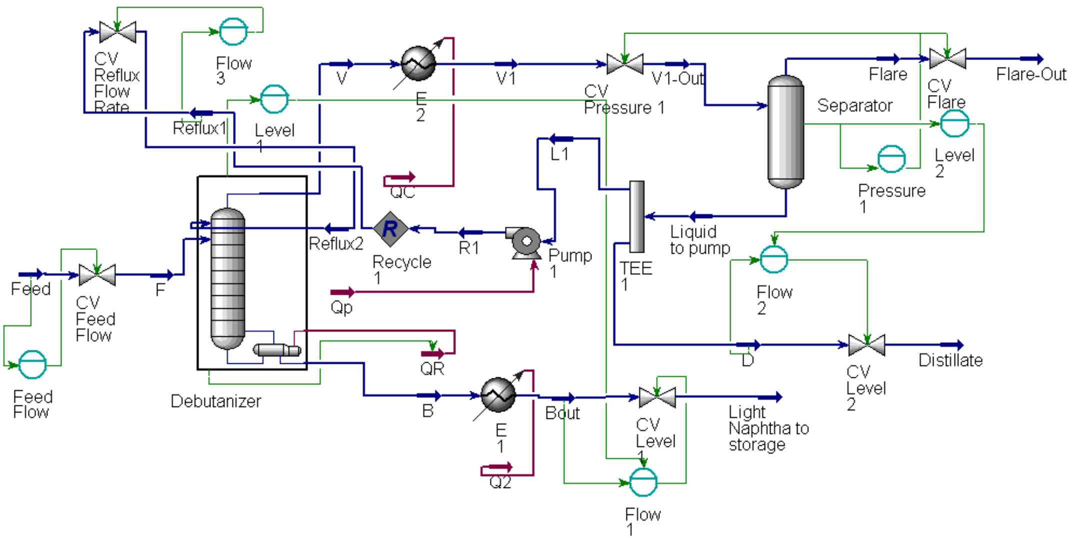

Figure 4.3: Dynamic model built in HYSYS^{TM} software

Table 4.4 shows the properties of Debutanizer streams at steady state while Table 4.5 shows the mole fractions of the specific streams at steady state. Figure 4.4 shows Debutanizer Pressure versus Tray position from Top at Steady state condition.

Table 4.4: Properties of Debutanizer Streams at Steady state

Name	**F**	**V1**	**B**
Vapour Fraction	0.329955937	3.45845952088873e-323	0.25638467
Temperature [C]	107.5022199	-65.18836681	109.270815
Pressure [kPa]	716.4107104	667.9100896	725.611079
Molar Flow [kgmole/h]	614.5723601	667.2681839	-680.146911
Mass Flow [kg/h]	46415.39482	29510.48412	-51368.2124
Liquid Volume Flow [m3/h]	71.6775569	58.16307157	-79.32579
Heat Flow [kJ/h]	-94096266.2	-86582023.84	104924061
Name	**V**	**Feed**	**Bout**
Vapour Fraction	1	0	0.36306013
Temperature [C]	15.29587791	136.4096062	112.999959
Pressure [kPa]	711.6783631	1913	784.552667
Molar Flow [kgmole/h]	667.2681822	614.5723601	-680.127157
Mass Flow [kg/h]	29510.48332	46415.39421	-51366.4022

Liquid Volume Flow [m3/h]	58.1630706	71.67755638	-79.3232117
Heat Flow [kJ/h]	-70582244.6	-94096256.19	102979001

Table 4.5: Mole fractions of Debutanizer Streams at Steady state

Name	**F**	**V1**	**B**
Comp Mole Frac (Propane)	6.32E-02	9.93E-01	6.32E-02
Comp Mole Frac (i-Butane)	0.120480088	3.75E-03	0.12047553
Comp Mole Frac (n-Butane)	8.03E-02	1.49E-03	8.03E-02
Comp Mole Frac (i-Pentane)	8.56E-02	6.44E-04	8.56E-02
Comp Mole Frac (n-Pentane)	0.114799082	5.94E-04	0.11479934
Comp Mole Frac (Hypo50_13*)	1.83E-02	8.62E-05	1.83E-02
Comp Mole Frac (Hypo60_13*)	0.167272327	2.67E-04	0.1672749
Comp Mole Frac (Hypo70_13*)	0.21529015	1.54E-04	0.21529465
Comp Mole Frac (Hypo80_13*)	5.66E-02	1.67E-05	5.66E-02
Comp Mole Frac (Hypo90_13*)	5.32E-02	2.07E-06	5.32E-02
Comp Mole Frac (Hypo100_13*)	2.22E-02	5.52E-08	2.22E-02
Comp Mole Frac (Hypo110_13*)	2.15E-03	2.56E-10	2.15E-03
Comp Mole Frac (Hypo120_13*)	6.74E-04	2.19E-12	6.74E-04
Name	**V**	**Feed**	**Bout**
Comp Mole Frac (Propane)	0.992997771	6.32E-02	6.32E-02
Comp Mole Frac (i-Butane)	3.75E-03	0.12048016	0.12048014
Comp Mole Frac (n-Butane)	1.49E-03	8.03E-02	8.03E-02
Comp Mole Frac (i-Pentane)	6.44E-04	8.56E-02	8.56E-02
Comp Mole Frac (n-Pentane)	5.94E-04	0.114799081	0.11479908
Comp Mole Frac (Hypo50_13*)	8.62E-05	1.83E-02	1.83E-02
Comp Mole Frac (Hypo60_13*)	2.67E-04	0.167272287	0.1672723
Comp Mole Frac (Hypo70_13*)	1.54E-04	0.215290078	0.2152901
Comp Mole Frac (Hypo80_13*)	1.67E-05	5.66E-02	5.66E-02
Comp Mole Frac (Hypo90_13*)	2.07E-06	5.32E-02	5.32E-02
Comp Mole Frac (Hypo100_13*)	5.52E-08	2.22E-02	2.22E-02
Comp Mole Frac (Hypo110_13*)	2.56E-10	2.15E-03	2.15E-03
Comp Mole Frac (Hypo120_13*)	2.19E-12	6.74E-04	6.74E-04

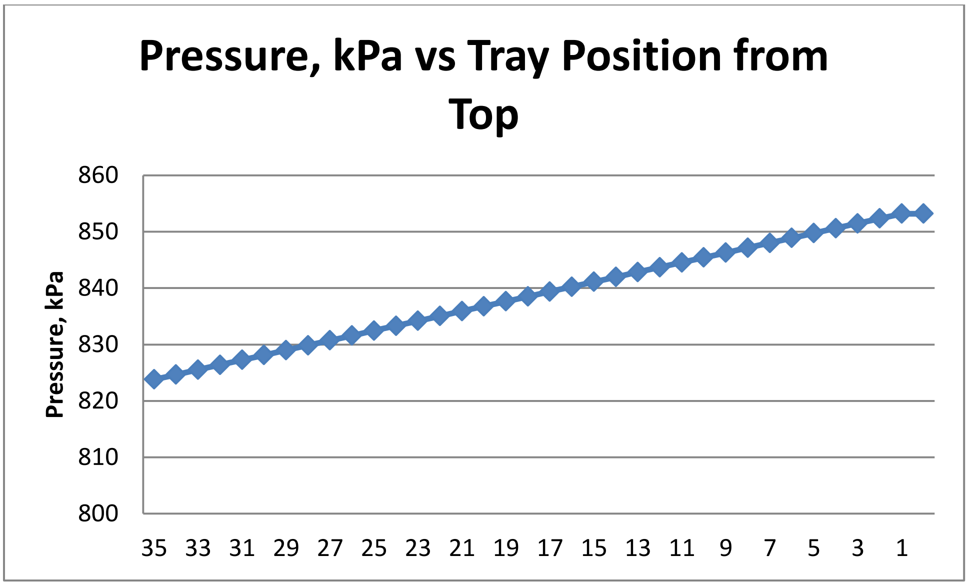

Figure 4.4: Debutanizer Pressure versus Tray position from Top at Steady state.

Figure 4.5 to 4.8 displays the properties of the Debutanizer at dynamic state. Since the condition is at dynamic state, the values are not constant.

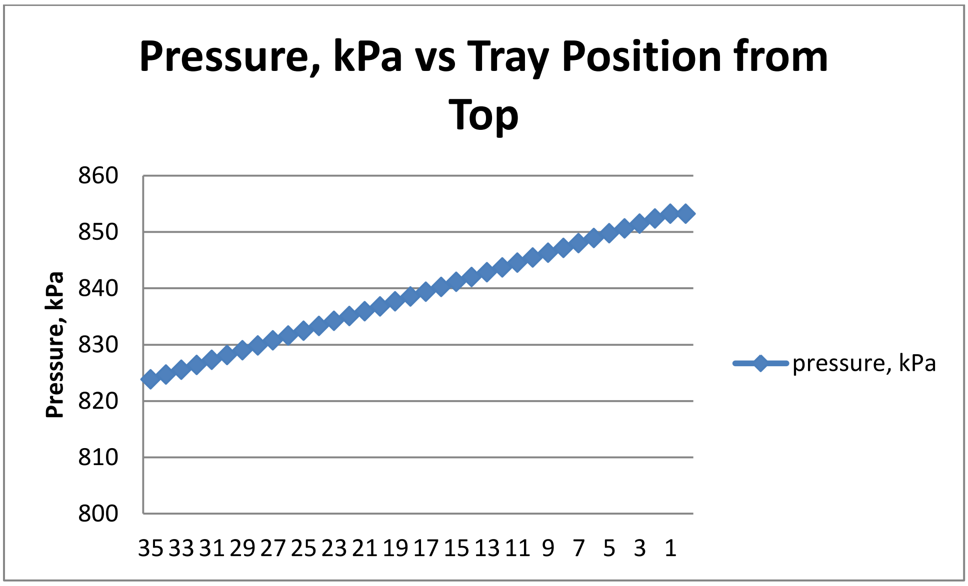

Figure 4.5: Debutanizer Pressure versus Tray Position from Top at Dynamic state.

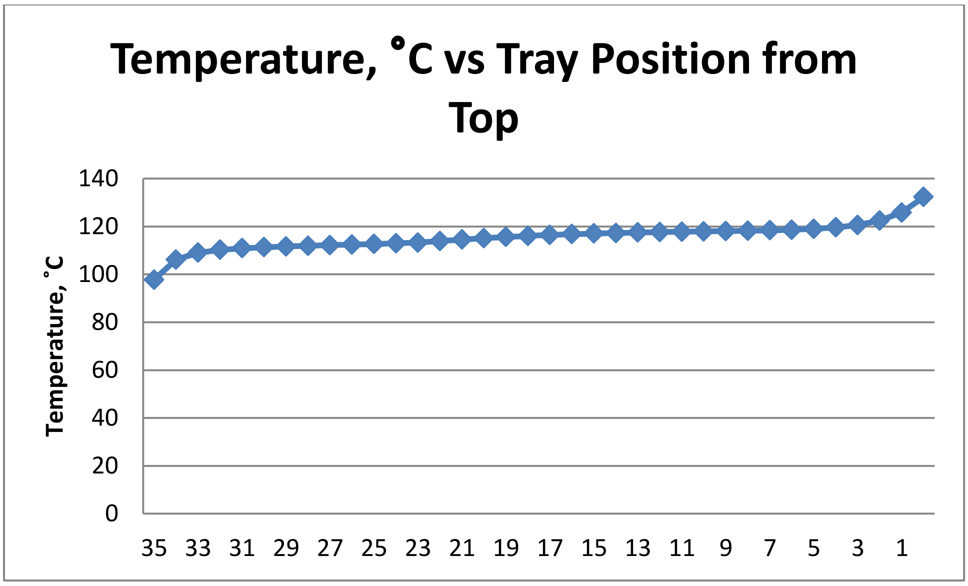

Figure 4.6: Debutanizer Temperature, ˚C versus Tray Position from Top at Dynamic state.

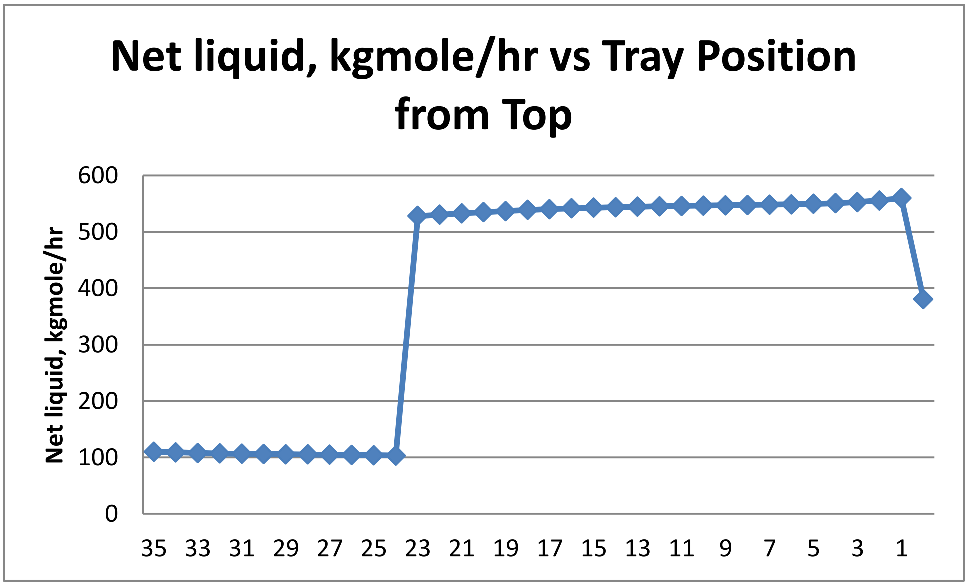

Figure 4.7: Debutanizer Net liquid, kgmole/hr versus Tray Position from Top at Dynamic state.

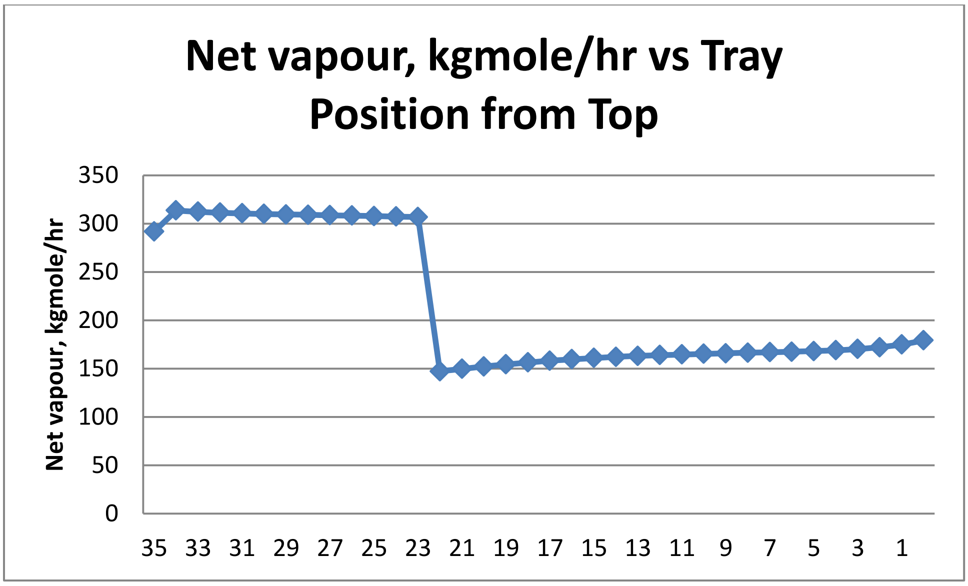

Figure 4.8: Debutanizer Net vapour, kgmole/hr versus Tray Position from Top at Dynamic state.

4.2.1 Open loop Step test results in HYSYS™

An open loop step test was conducted in HYSYS™ by changing the opening percentage (OP) of the Feed Flow controller from 100% to 50% repeatedly for a few times. The corresponding responses of all the other 7 controllers were observed. The 7 controllers are Level 1 controller, Level 2 controller, Flow 1 controller, Flow 2 controller, Flow 3 controller, Pressure 1 controller and Temp 5 controller. Pressure 1 controller did not give any significant response to the applied step test. Therefore, it has been omitted. Figure 4.9 shows the details of the Feed Flow controller when the step test was conducted. The responses of the other 6 controllers are presented below from Figure 4.10 to Figure 4.15. The shown responses are the behavior of the process variable (PV) at each controller. The Flow controllers show process variables in flowrate (m^3/hr). The Level controllers show process variables in percentage (%). Meanwhile the Temperature controller shows process variable in degree Celsius (˚C).

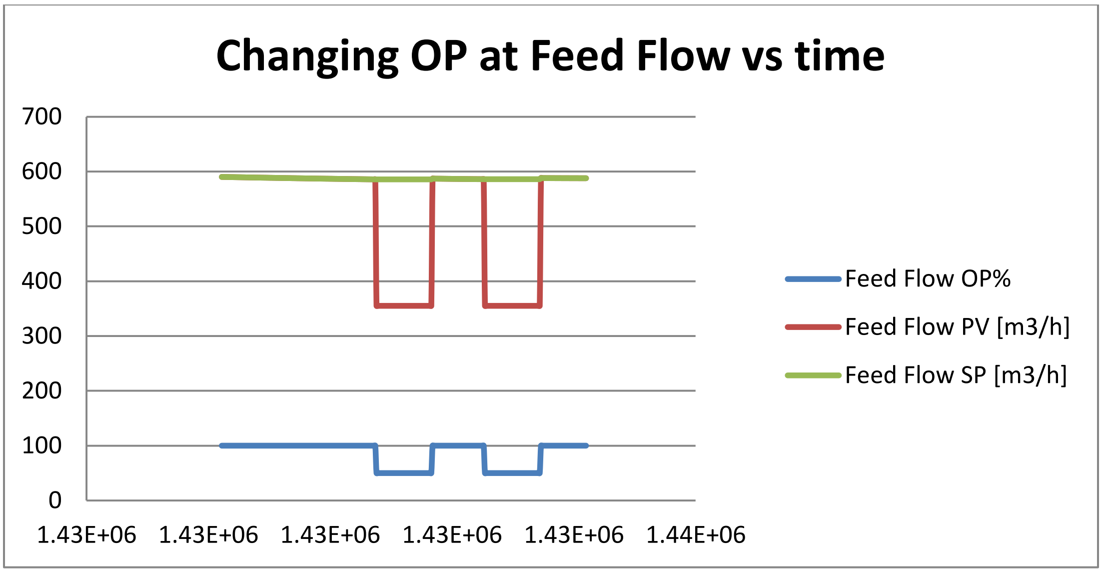

Figure 4.9: Details of Feed Flow controller during step test.

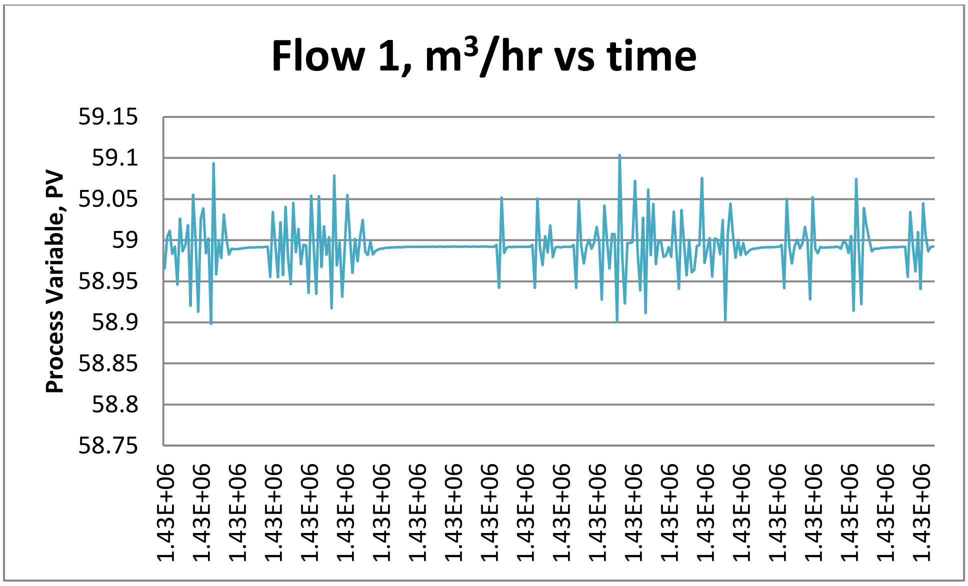

Figure 4.10: Response of Flow 1 controller during step test.

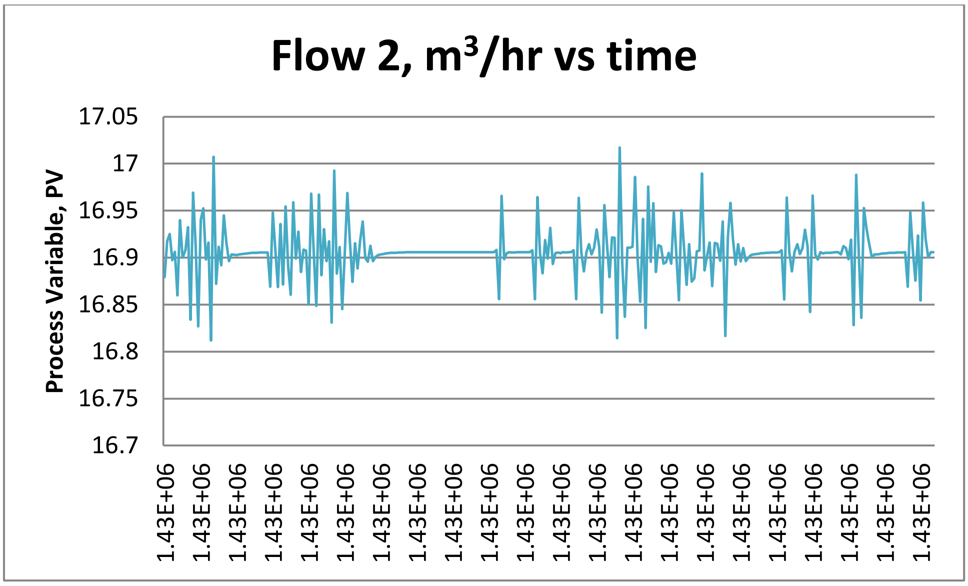

Figure 4.11: Response of Flow 2 controller during step test.

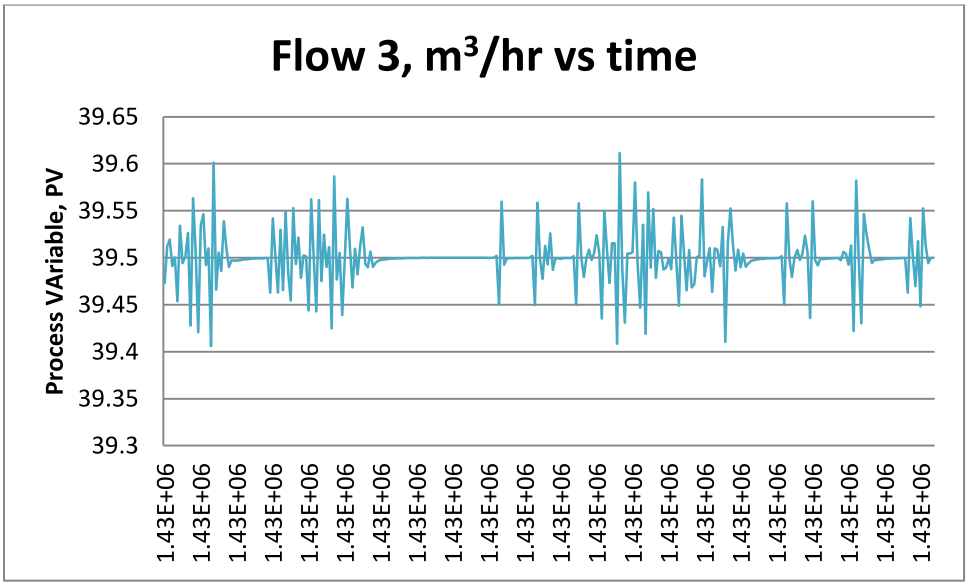

Figure 4.12: Response of Flow 3 controller during step test.

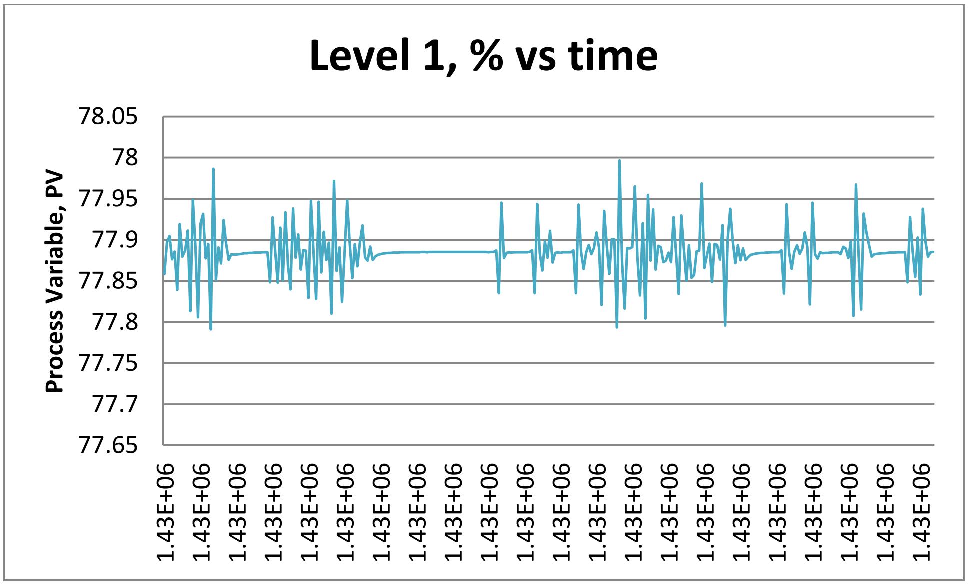

Figure 4.13: Response of Level 1 controller during step test.

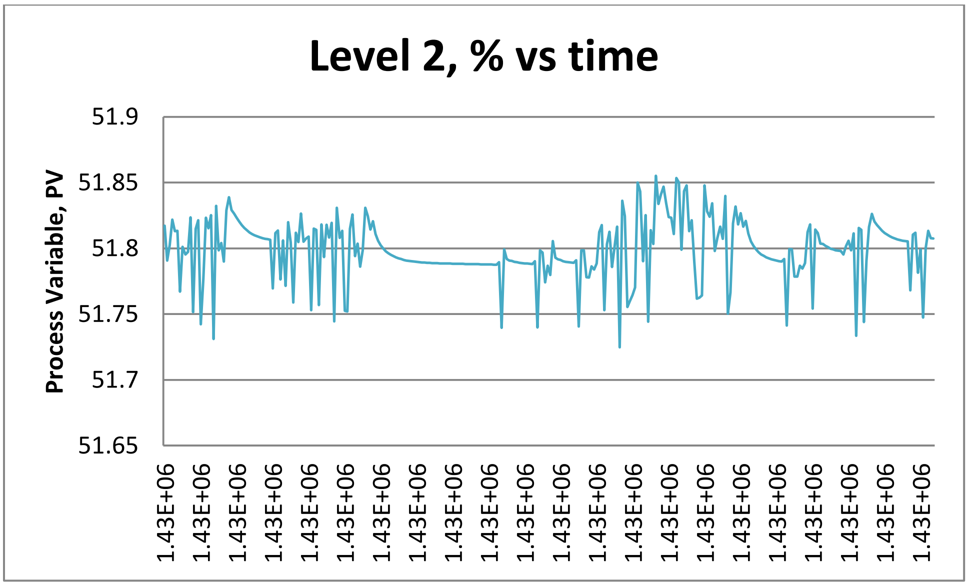

Figure 4.14: Response of Level 2 controller during step test.

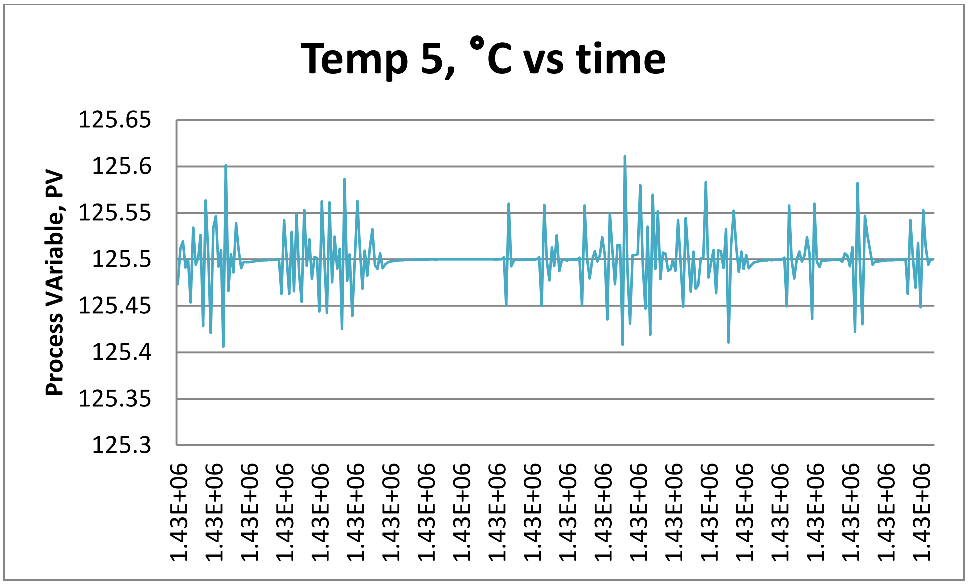

Figure 4.15: Response of Temp 5 controller during step test.

When the step test was conducted, deviations were observed at the composition of the top and bottom streams of the Debutanizer column. The mole fractions of major components showed variations. The major components are propane, n-pentane, n-butane, i-pentane and i-butane. The top stream exiting the Debutanizer is labelled as Distillate while the bottom stream is labelled as Light Naphtha to Storage. Figure 4.16 to Figure 4.25 shows the mole fraction fluctuations with respect to time, of the top and bottom streams exiting the Debutanizer.

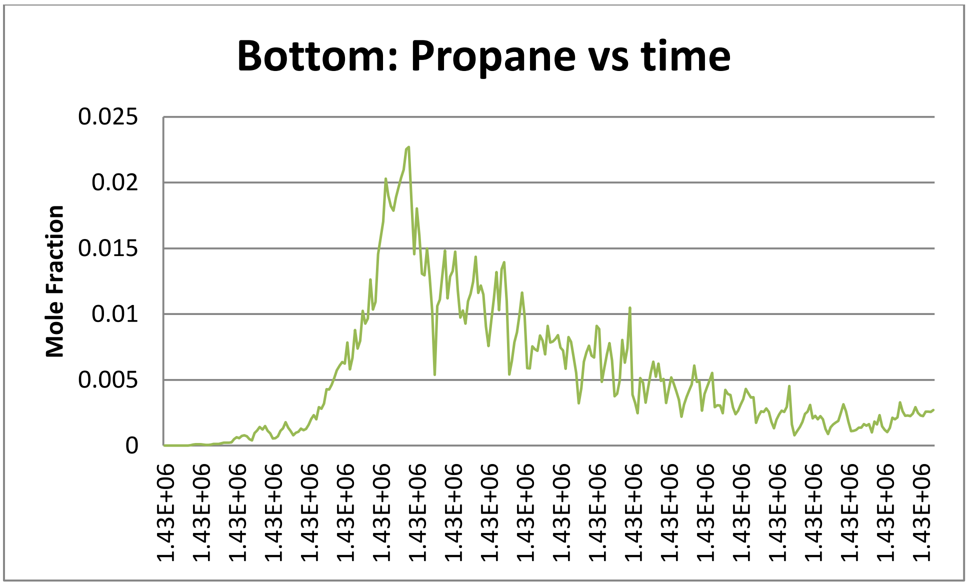

Figure 4.16: Mole fraction of propane at the bottom exit of Debutanizer.

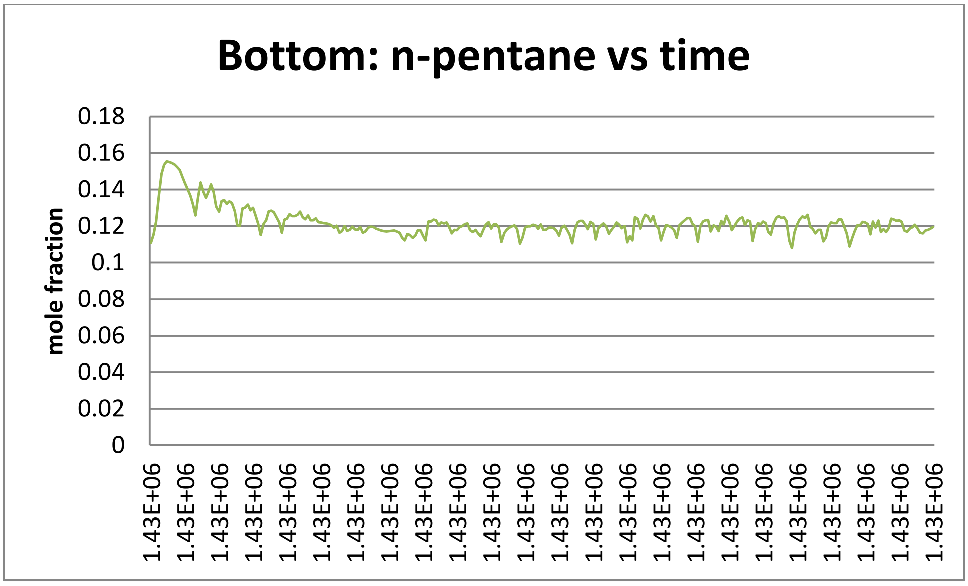

Figure 4.17: Mole fraction of n-pentane at the bottom exit of Debutanizer.

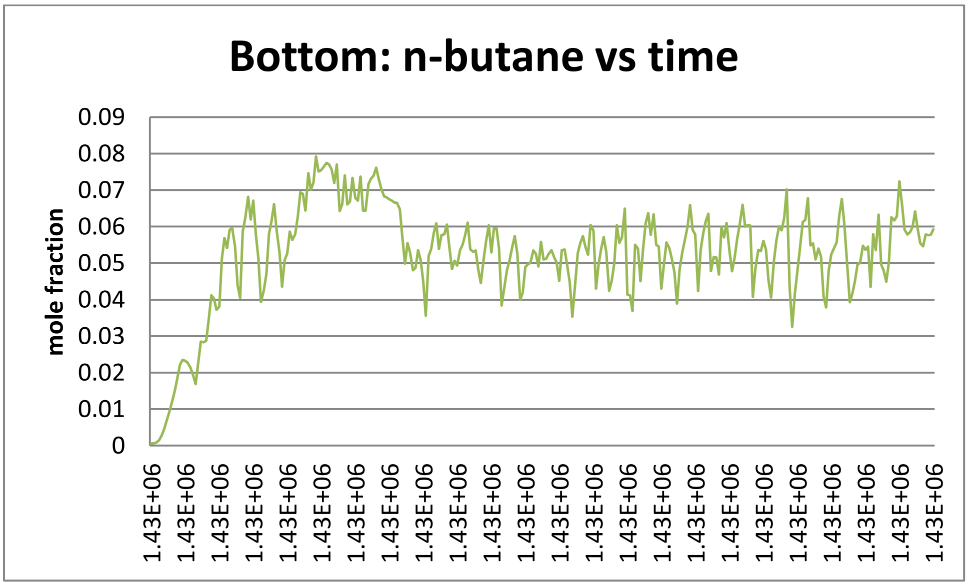

Figure 4.18: Mole fraction of n-butane at the bottom exit of Debutanizer.

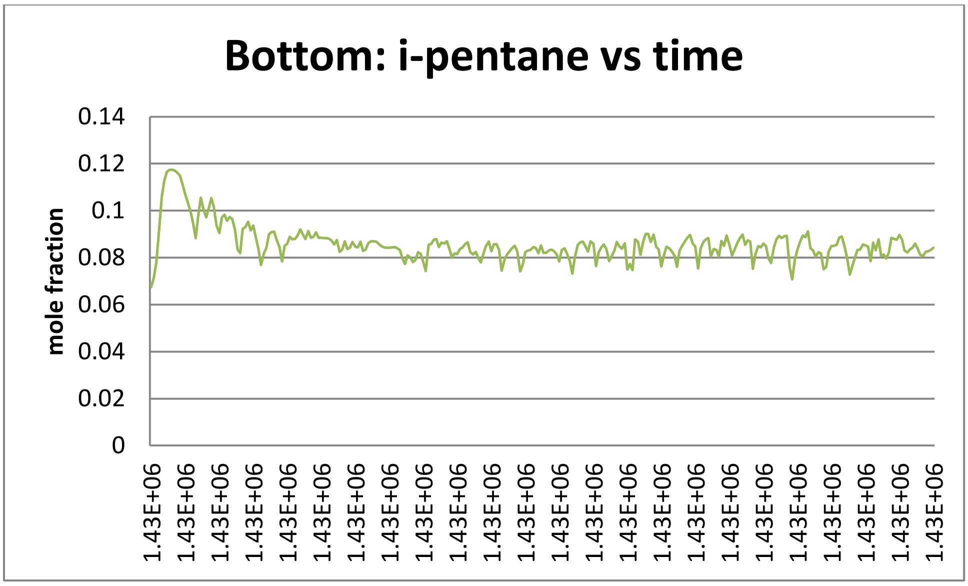

Figure 4.19: Mole fraction of i-pentane at the bottom exit of Debutanizer.

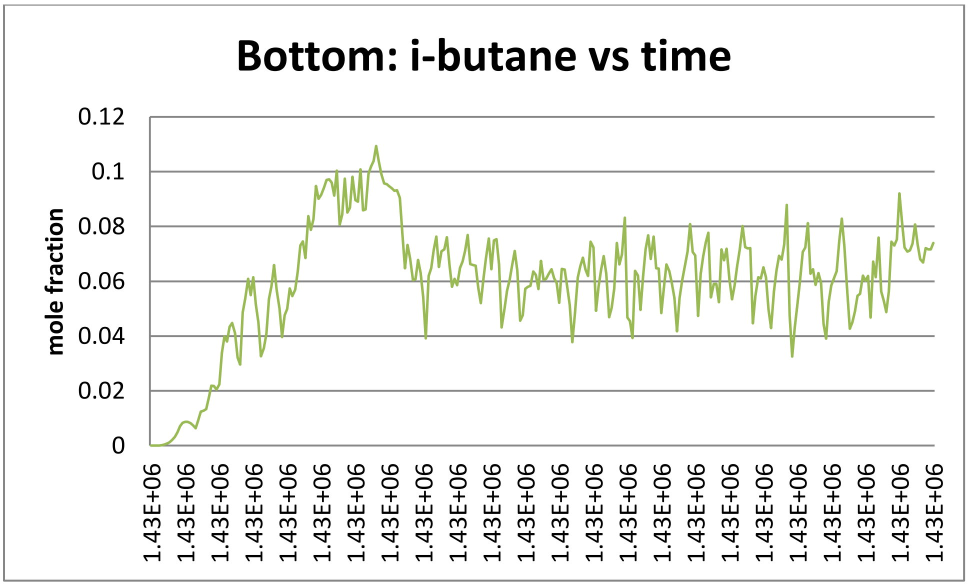

Figure 4.20: Mole fraction of i-butane at the bottom exit of Debutanizer.

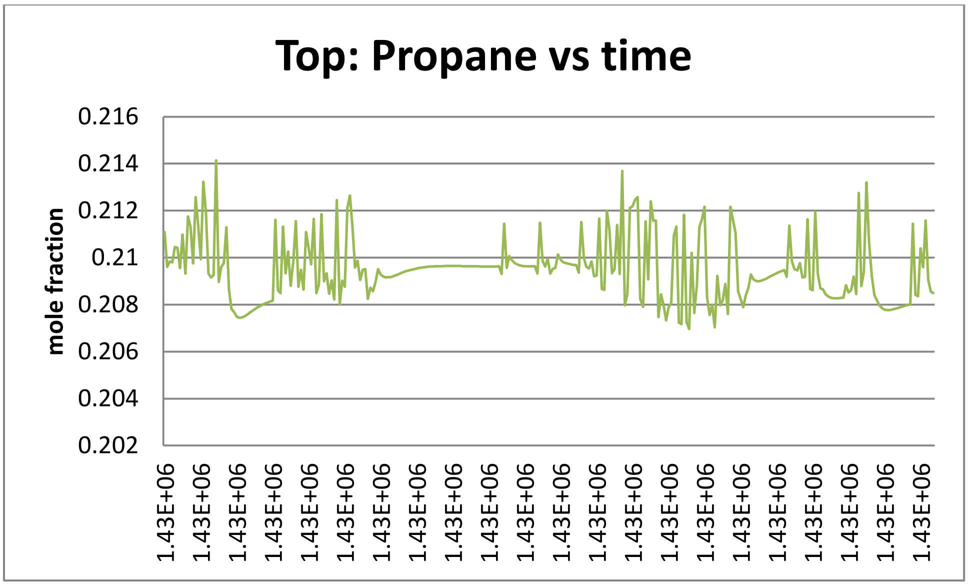

Figure 4.21: Mole fraction of propane at the top exit of Debutanizer.

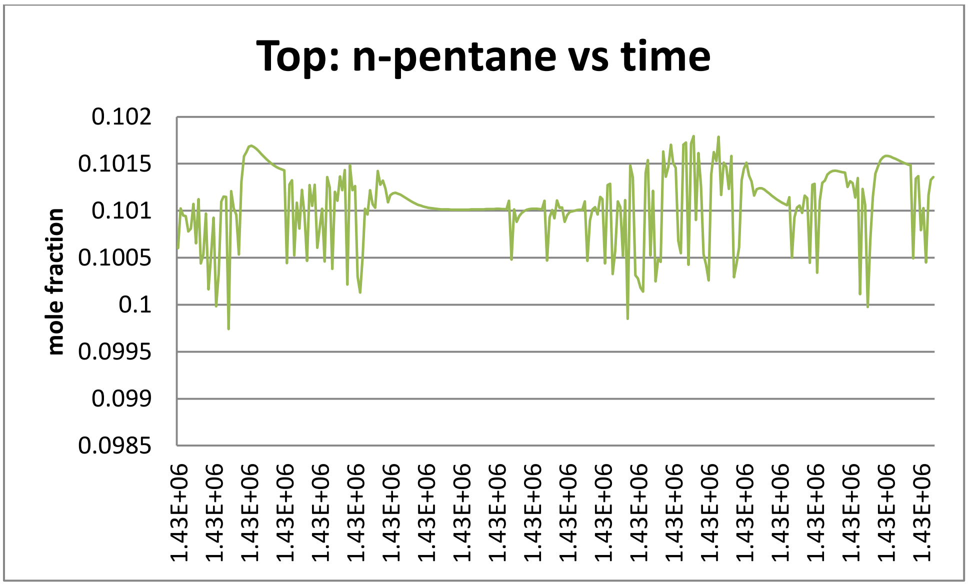

Figure 4.22: Mole fraction of n-pentane at the top exit of Debutanizer.

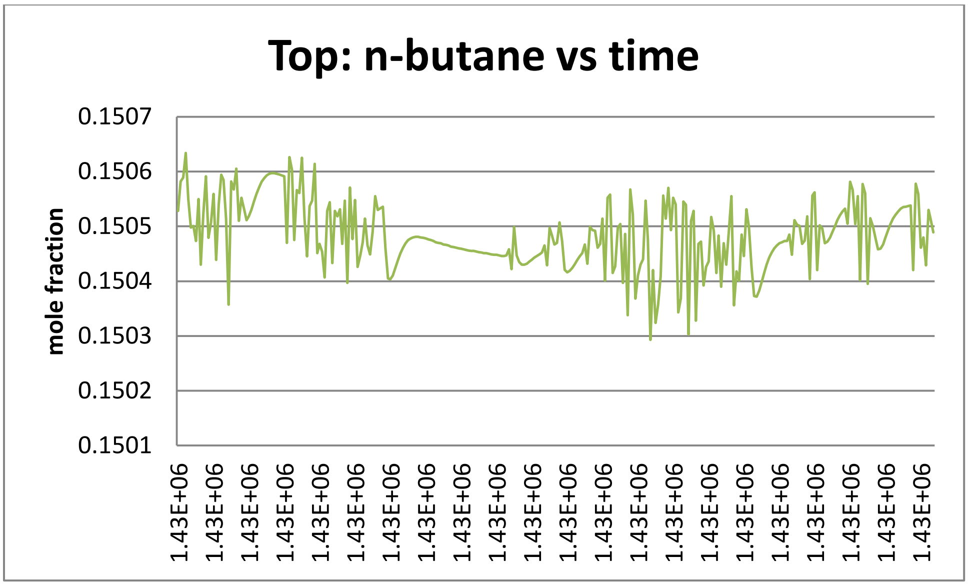

Figure 4.23: Mole fraction of n-butane at the top exit of Debutanizer.

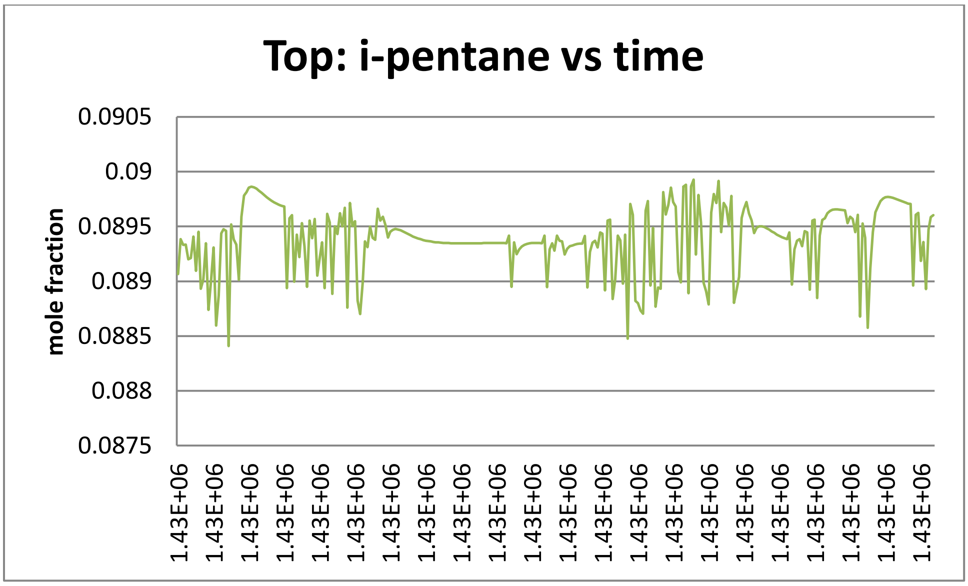

Figure 4.24: Mole fraction of i-pentane at the top exit of Debutanizer.

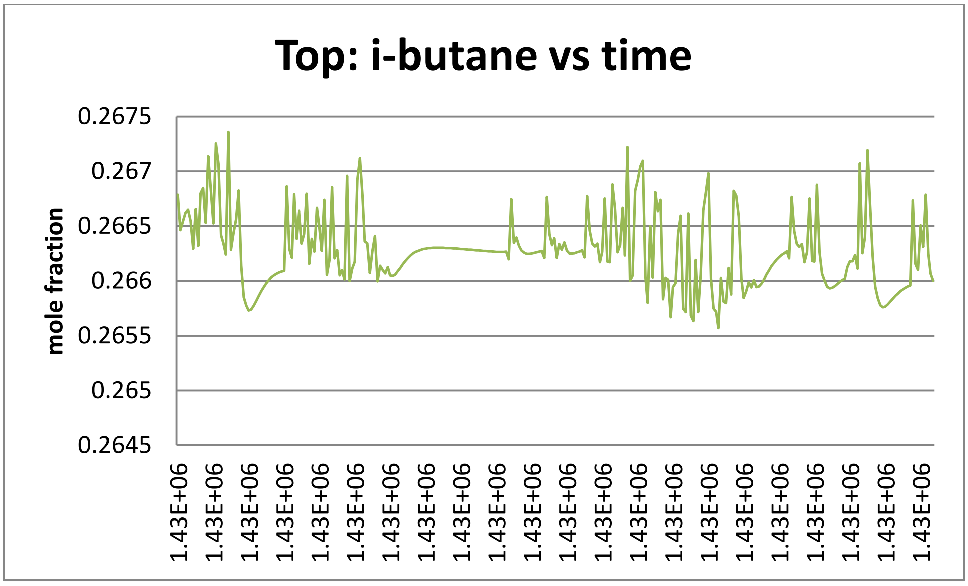

Figure 4.25: Mole fraction of i-butane at the top exit of Debutanizer.

As explained above, open loop step tests were conducted by changing the OP of the other controllers and the responses were observed. Similar results were obtained; therefore it is not shown in this report. However the findings were used for MATLAB programming which will be discussed in the next section.

4.3 MATLAB Programming Results

Once the open loop step test was carried out in HYSYS^{TM}**,** the results were analyzed using the MATLAB software. '2 by 2' matrix method was used. The IDENT system identification tool from MATLAB was used to process the results in order to obtain the transfer functions of each controller. 2 methods were used in MATLAB programming to obtain the overall transfer function of all the controllers which are the 'no constrain method' and 'quadratic programming method'. The response of the controllers with respect to its process variables and manipulated variables were studied. The findings from the MATLAB programming are discussed below according to the method used. The 'no constrain method' does not include the manipulated variable data meanwhile the 'quadratic programming method' includes the manipulated variable data.

4.3.1 No Constrain Method

A total of 7 controllers were studied using this method. How the controllers behave with respect to its process variables and manipulated variables will be discussed. Figure 4.26 shows the Feed Flow controller response. The outputs behave like an ideal controller. Figure 4.27 shows the Flow 1 controller response. The outputs have several peaks which show instability.

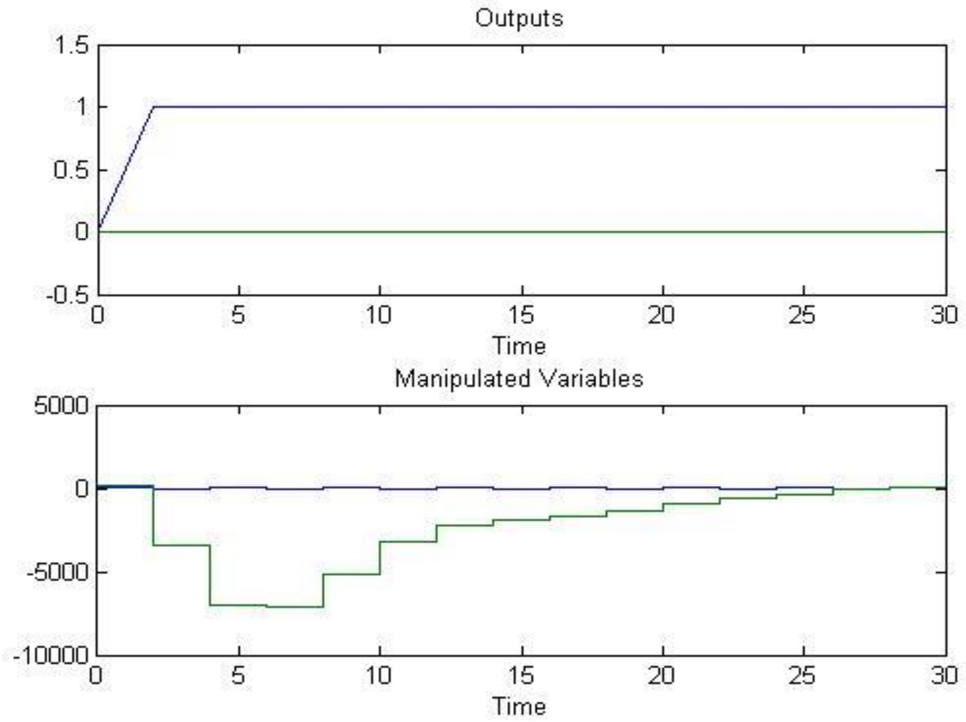

Figure 4.26: Feed Flow Controller response

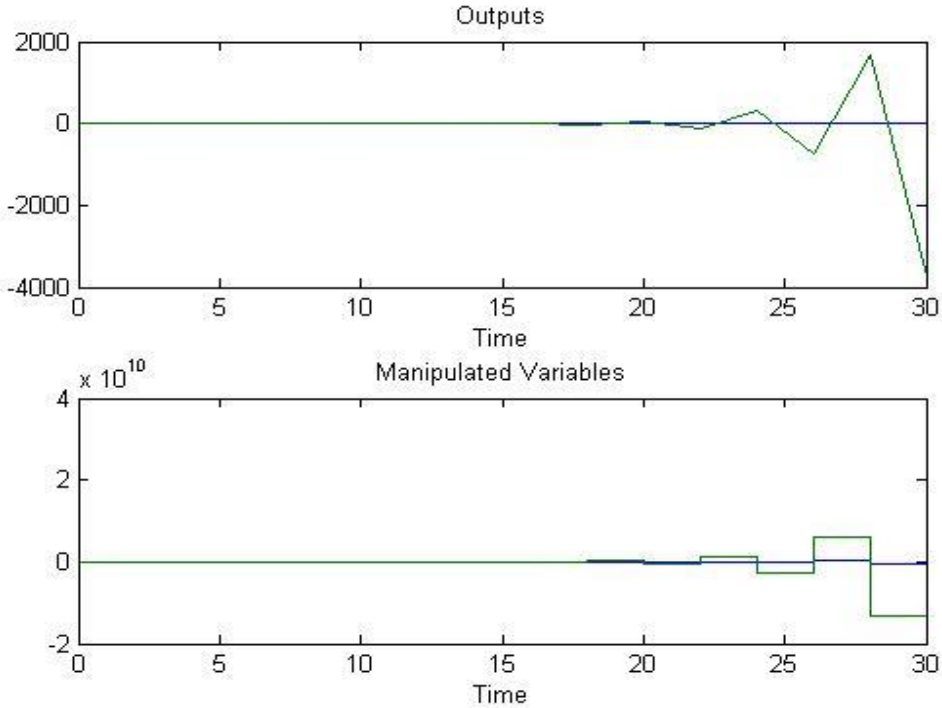

Figure 4.27: Flow 1 Controller response

The response of Flow 2 controller is shown in Figure 4.28. The outputs contain several positive and negative peaks which show controller fluctuation. Settling time is yet to be observed. The manipulated variables show ringing effect. Figure 4.29 shows Flow 3 Controller response. The output behaves accordingly to the manipulated variables. However the process gain, K_p value will be negative.

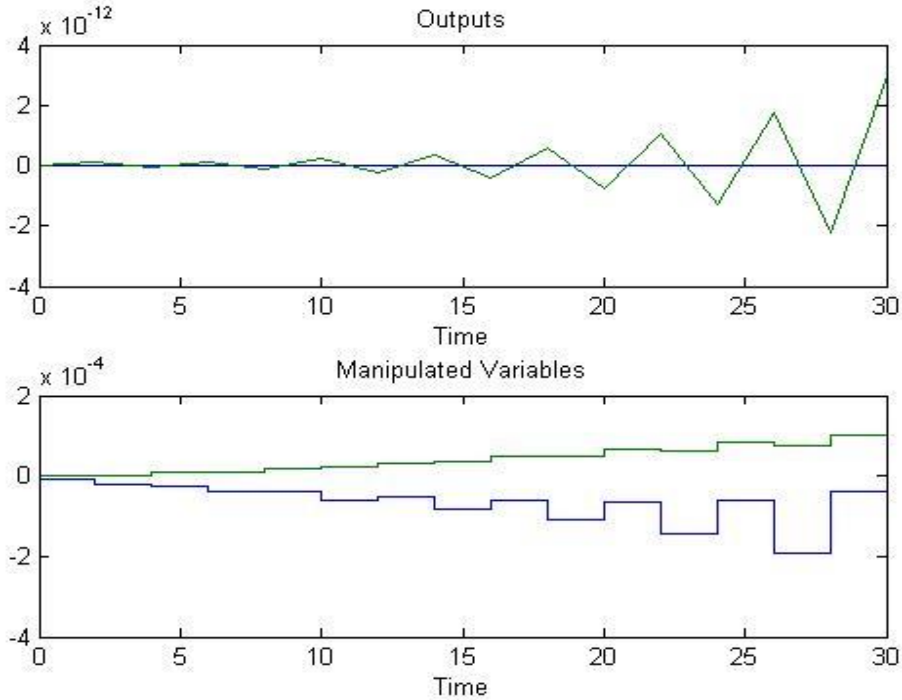

Figure 4.28: Flow 2 Controller response

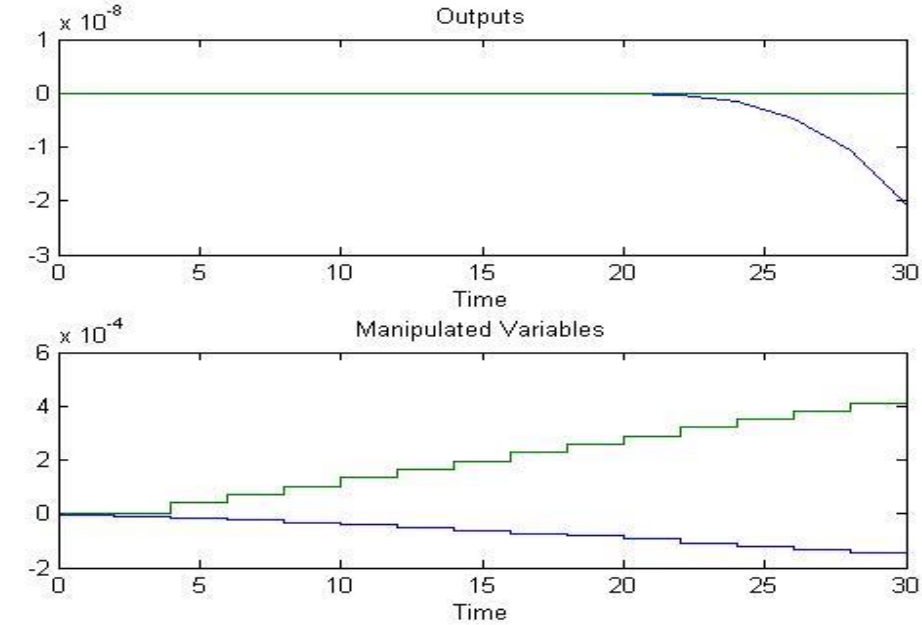

Figure 4.29: Flow 3 Controller response

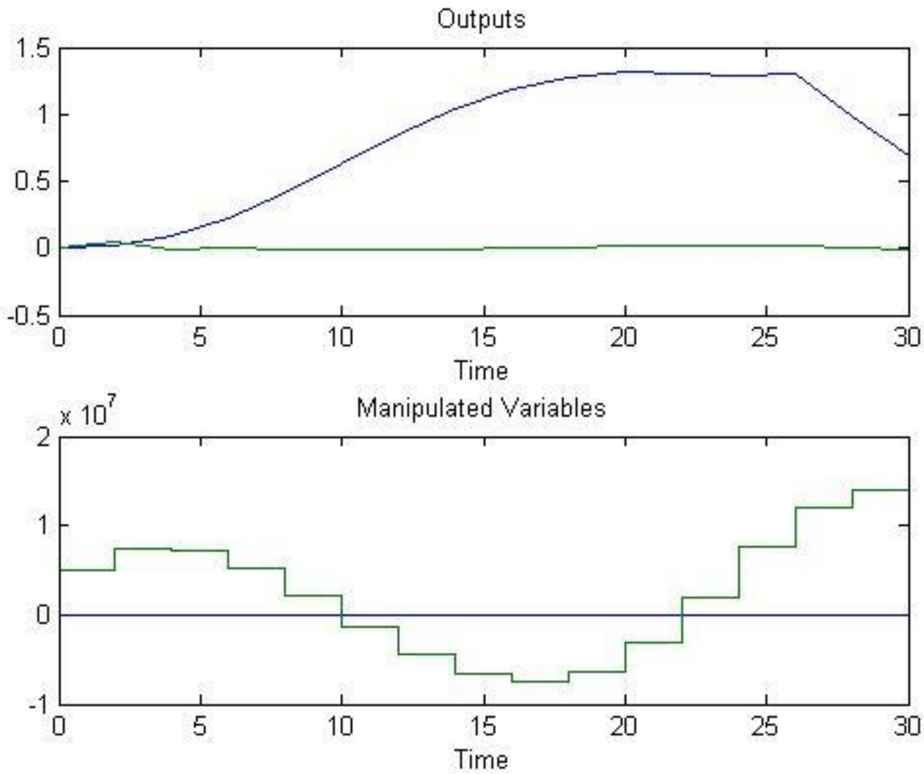

Figure 4.30: Level 1 Controller response

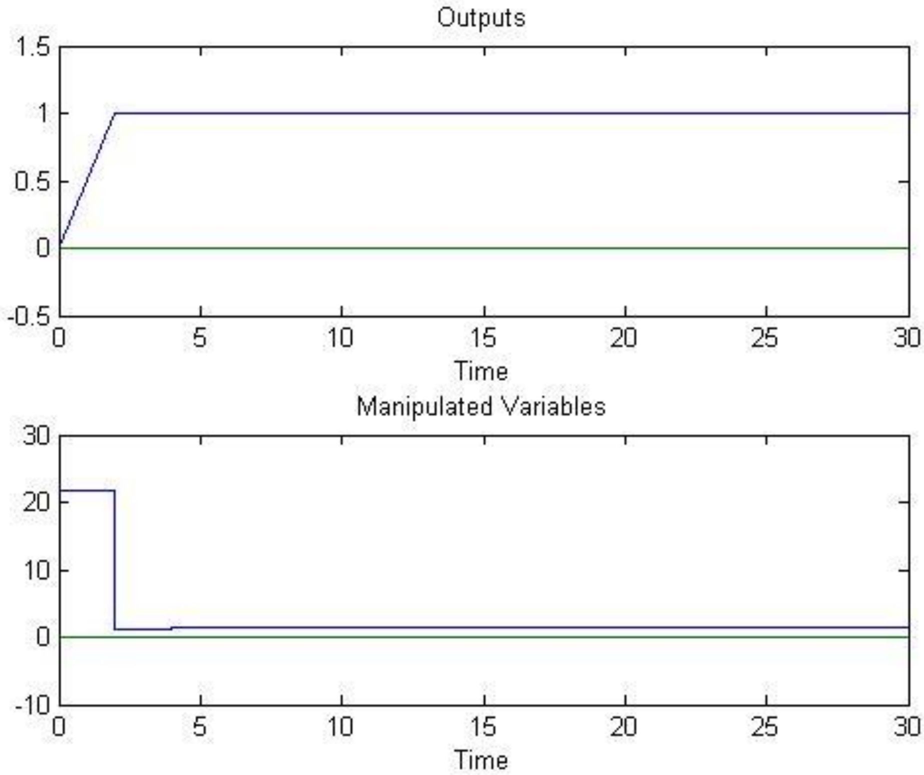

Figure 4.31: Level 2 Controller response

Figure 4.30 shows Level 1 Controller response. A small time delay is observed in the outputs. The manipulated variables show ringing effect. Figure 4.31 shows Level 2 Controller response. The outputs behave similar to an ideal controller. Figure 4.32 shows Temperature 5 Controller response. The outputs behave accordingly to the manipulated variables. However the process gain, K_p value will be negative.

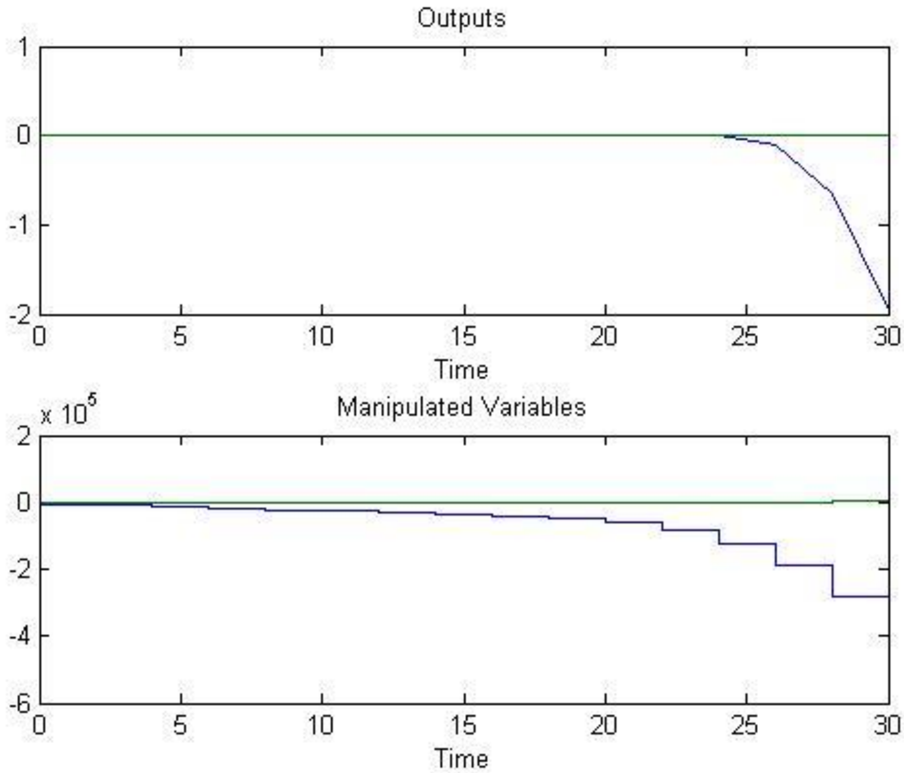

Figure 4.32: Temperature 5 Controller response

4.3.2 Quadratic Programming Method

A total of 6 controllers were studied using this method. How the controllers behave with respect to its process variables and manipulated variables will be discussed. Figure 4.33 shows Feed Flow Controller response. The outputs behave oppositely to the manipulated variables. The process gain, K_p value will be positive. It's a stable process. Figure 4.34 shows Flow 1 Controller response. The outputs behave accordingly to the manipulated variables. A time delay is observed.

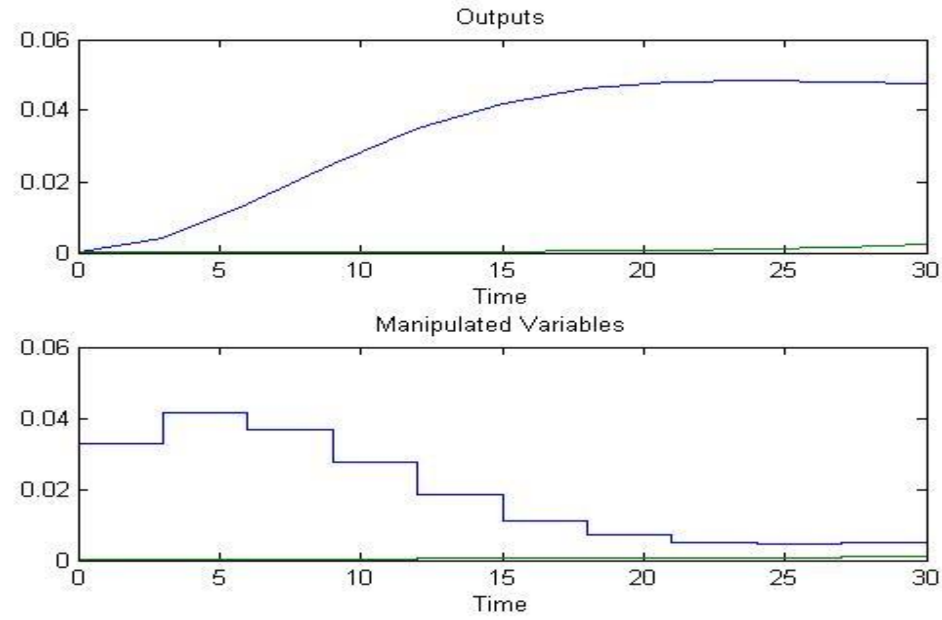

Figure 4.33: Feed Flow Controller response

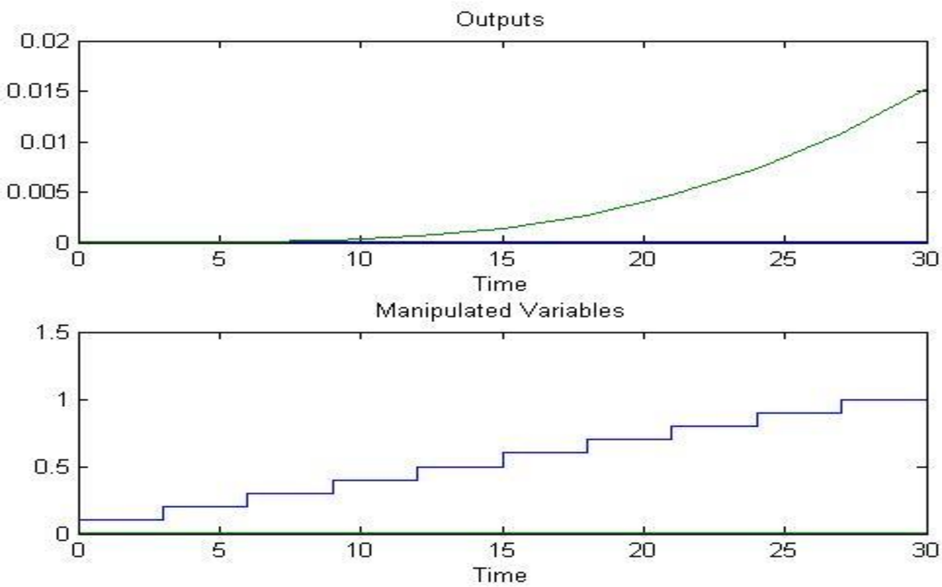

Figure 4.34: Flow 1 Controller response

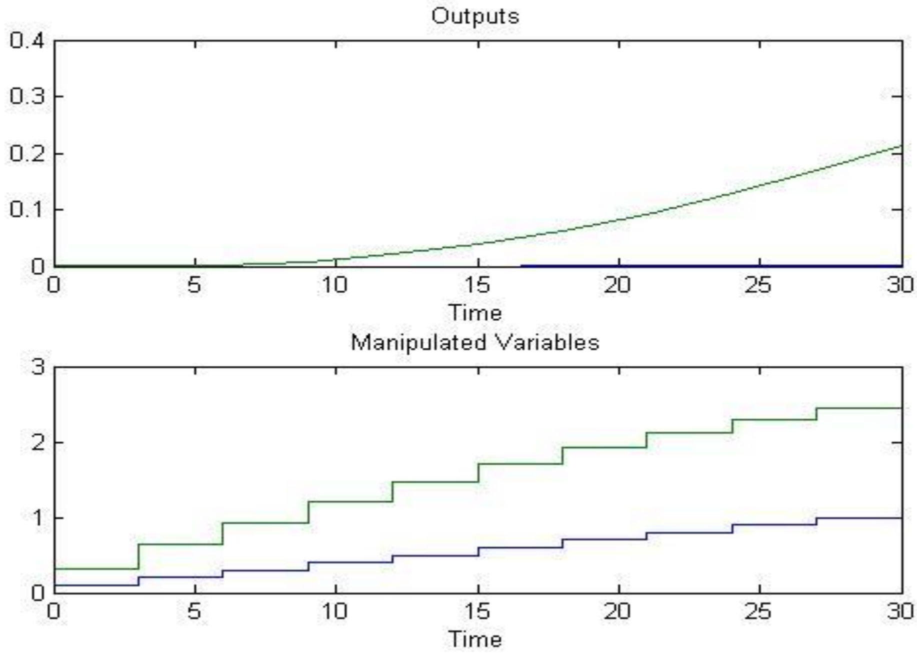

Figure 4.35: Flow 3 Controller response

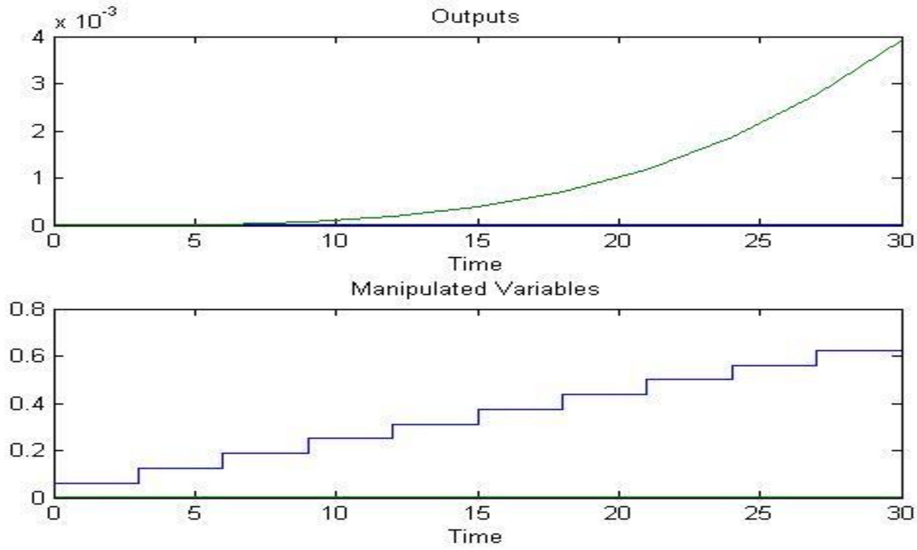

Figure 4.36: Level 1 Controller response

Figure 4.35 shows Flow 3 Controller response. The stable process behaves accordingly to the step change introduced. A time delay is observed. Level 1 Controller response is shown in Figure 4.36. A time delay is observed. The outputs behave accordingly to the manipulated variables. The process gain, K_p is a positive value.

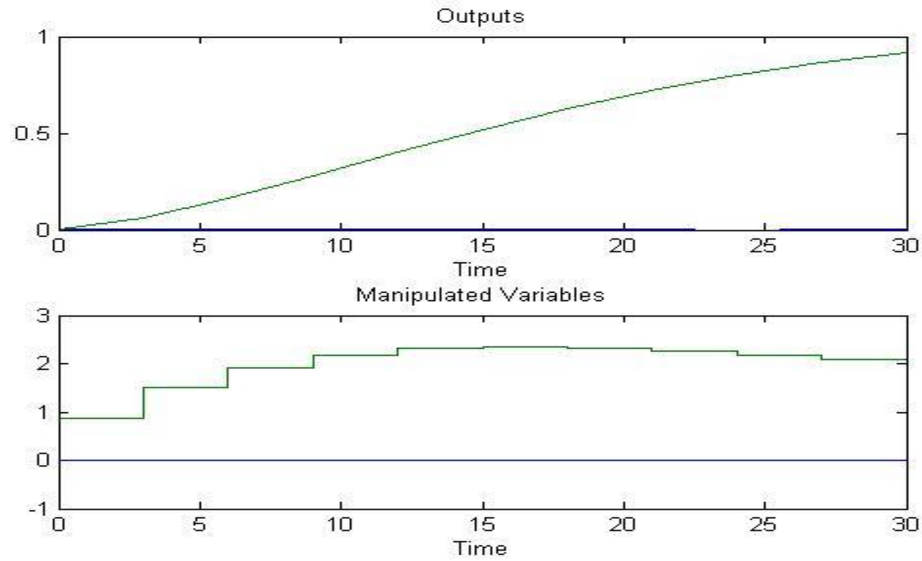

Figure 4.37: Level 2 Controller response

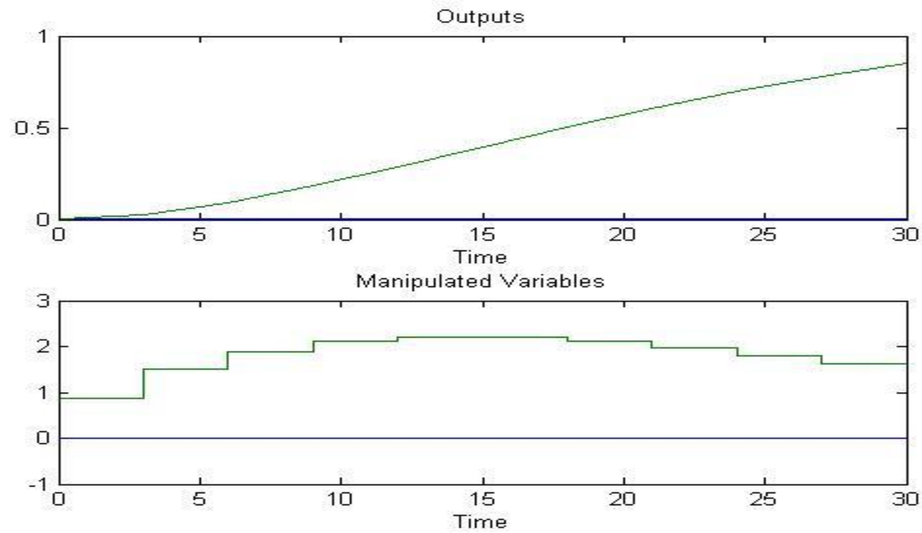

Figure 4.38: Temperature 5 Controller response

Figure 4.37 shows Level 2 Controller response. The manipulated variables have slight ringing effect. The output is stable and has a positive process gain, $K_{p.}$ Figure 4.38 shows Temperature 5 Controller response. The outputs are stable and have a positive process gain, $K_{p.}$

From all the graphs obtained, the process gain, time delay and time constant values were analyzed and used in HYSYS^{TM} for the MPC controllers input values. . Table 4.6 shows the MPC controller properties for 'no constrain method' and Table 4.7 shows the MPC controller properties for the 'quadratic programming method'.

Table 4.6:MPC controller properties for 'no constrain method'.

Controller Name	Process Gain	Time constant (min)	Time delay (min)
Feed Flow	-0.058920987	1	1
Flow 1	-1.62132E-09	29	1
Flow 2	5.94448E-14	29	1
Flow 3	0.000133236	28.7	1
Level 1	5.51966E-05	8.5	1
Level 2	-0.049916754	0.1	1
Temp 5	4.70305E-06	29	1

Table 4.7:MPC controller properties for the 'quadratic programming method'.

Controller Name	Process Gain	Time constant (min)	Time delay (min)
Feed Flow	-1.732487443	12	1
Flow 1	2.92654E-05	27	1
Flow 3	0.000357936	26	1
Level 1	1.46625E-05	27	1
Level 2	-0.009314714	12, 17, 27	1
Temp 5	-0.000856223	27	1

4.4 MPC results from HYSYS™

PID controllers in the dynamic mode of HYSYS™ simulation done previously were replaced by MPC controllers. 2 types of MPC controller settings were compared which are the 'no constrain method' and the 'quadratic programming method'. A general MPC step response method readily available in HYSYS™ was also studied. Therefore the responses of the MPC controllers using these 3 methods with respect to the PID controllers were studied. Figure 4.39 to Figure 4.47 shows the results of the comparison done.

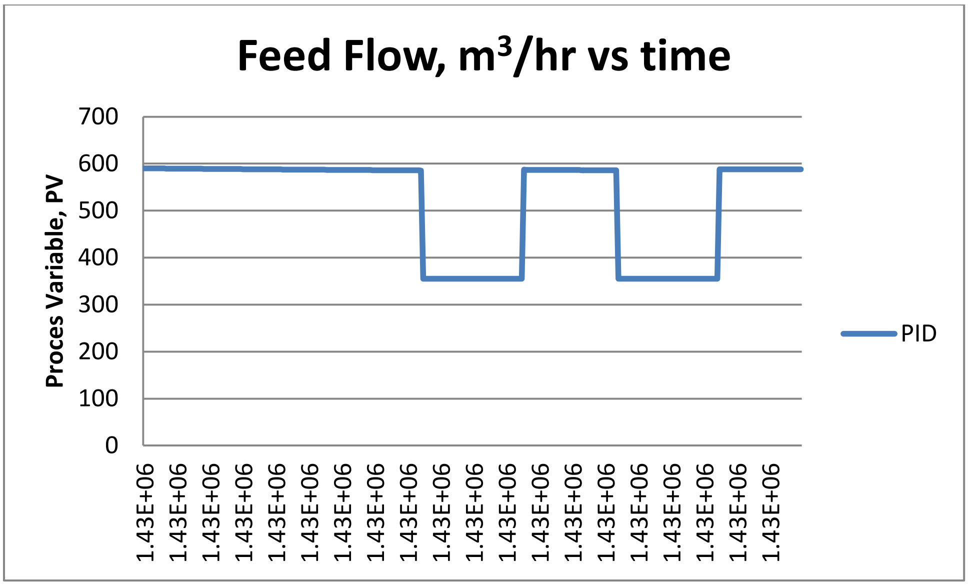

Figure 4.39: Feed Flow - PID controller response.

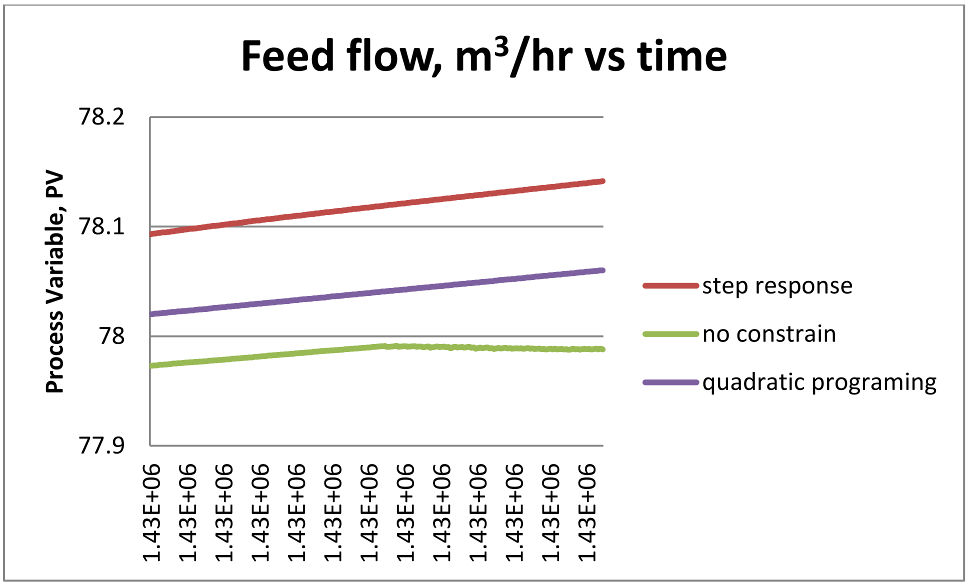

Figure 4.40: Comparison of Feed Flow - MPC controllers' response.

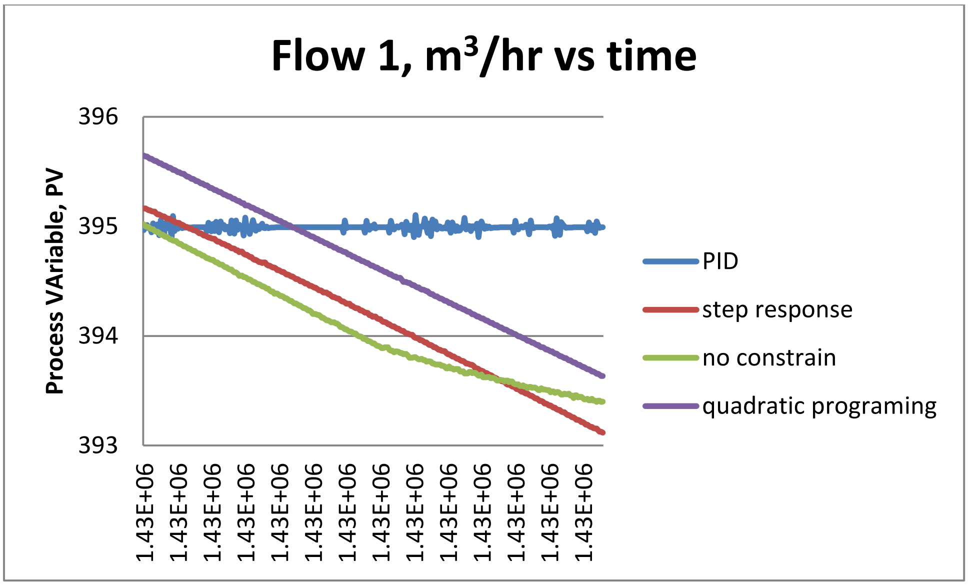

Figure 4.41: Comparison of Flow 1 controllers' response.

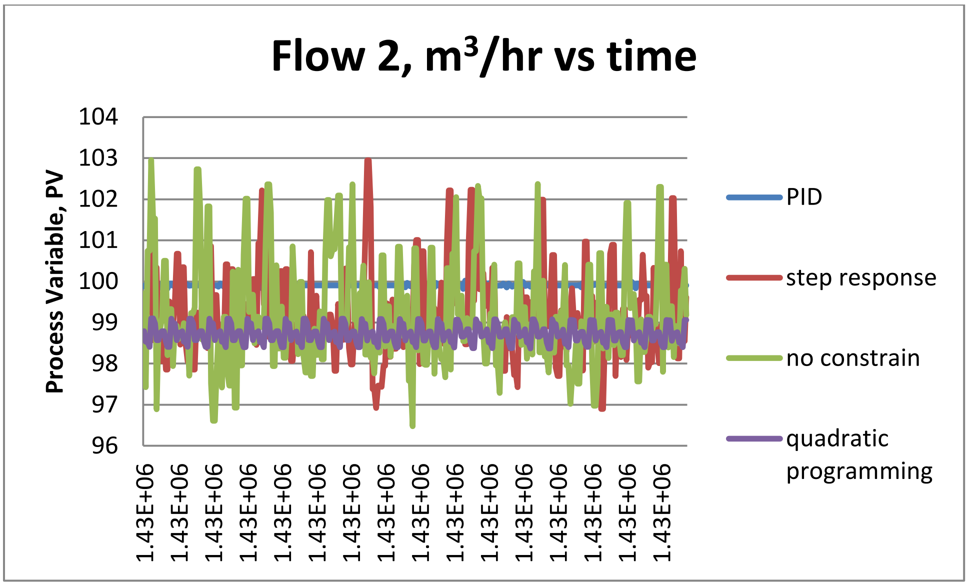

Figure 4.42: Comparison of Flow 2 controllers' response.

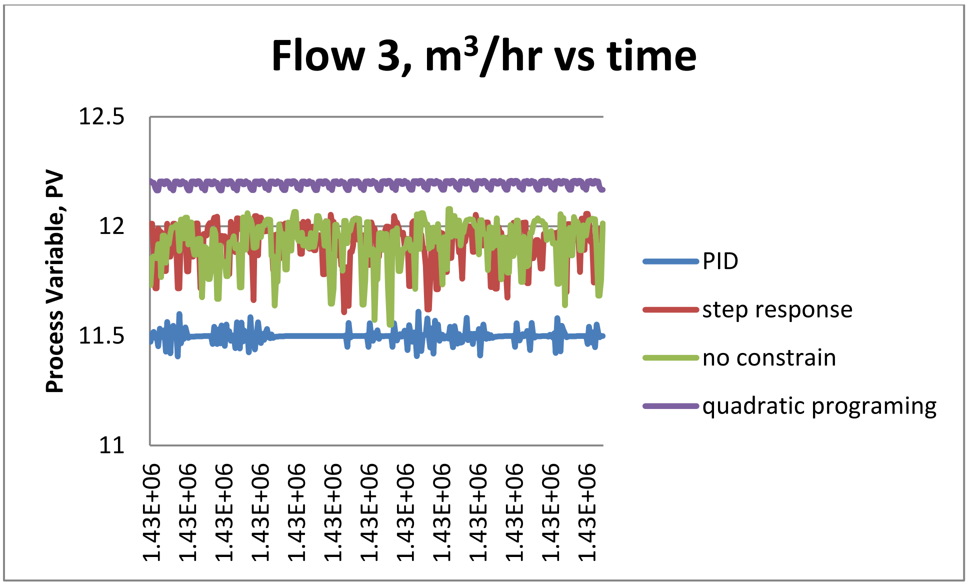

Figure 4.43: Comparison of Flow 3 controllers’ response.

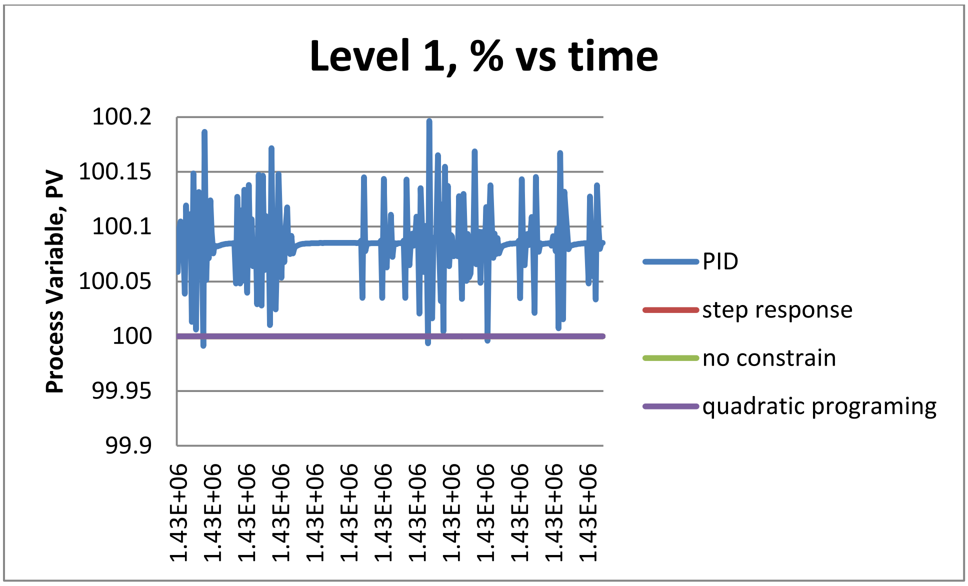

Figure 4.44: Comparison of Level 1 controllers’ response.

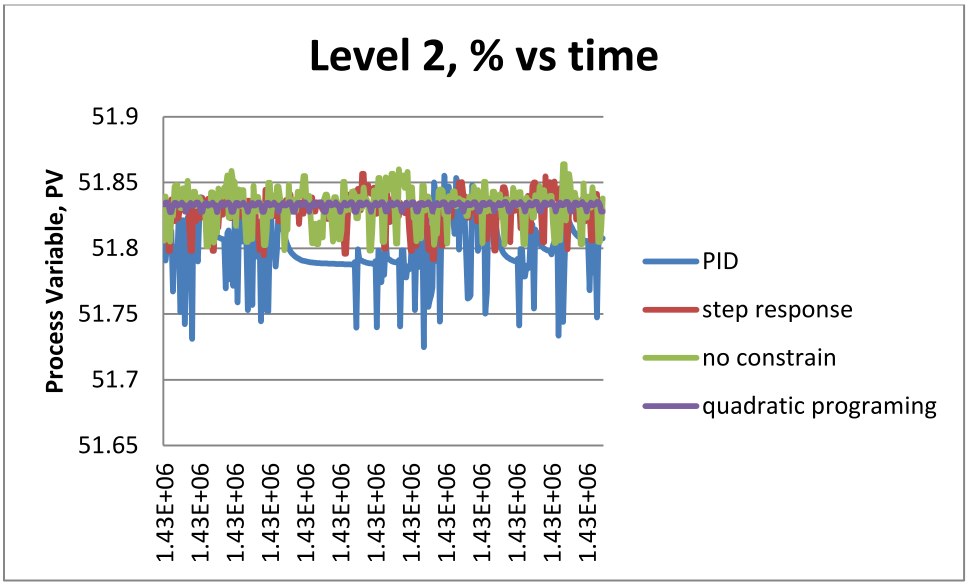

Figure 4.45: Comparison of Level 2 controllers' response.

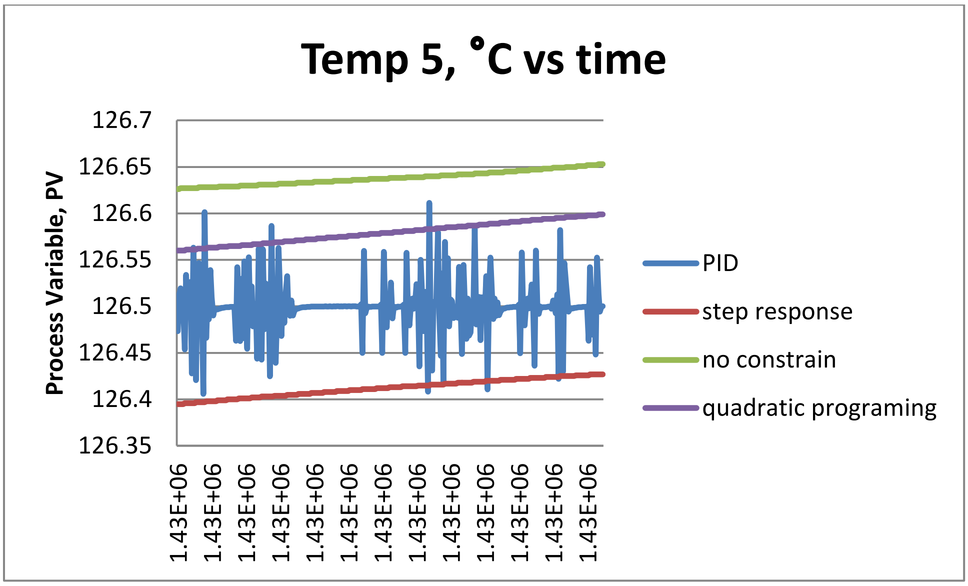

Figure 4.46: Comparison of Temp 5 controllers' response.

From the detailed analysis and comparison done between the PID and MPC controllers, it is proven that the MPC controller settings with the 'quadratic programming method' provides a better control system as a whole to the simulated C-110 Debutanizer column. This method shows lesser fluctuations and the control system contains lower amount of noise. Therefore, from this project, this MPC controller setting is suggested to be used at the plant operation of PP(T)SB to achieve the desired outlet composition for the C-110 Debutanizer column.

4.5 Relevancy to objective

The main objective of this project is to come up with a control strategy that can maximize the outlet composition of the Debutanizer column. As the initial steps, in FYP 1, the steady state model and the dynamic model of the Debutanizer column have been developed using the HYSYS^{TM} software. In FYP 2, MATLAB programming is done to obtain the optimum tuning parameters for the MPC controllers.

4.6 Future Work

Different MPC tuning relations can be further studied and compared in order to achieve the most efficient control strategy that can maximize the Debutanizer product output. Different simulation software like iCON can be used to simulate the real plant environment.

CHAPTER 5

CONCLUSION AND RECOMMENDATION

From this project, we have simulated a Debutanizer model using HYSYS^{TM} software. MATLAB programming was carried out to obtain optimum tuning parameters for the MPC controllers that can maximize the output of the Debutanizer column. From the detailed analysis and comparison done between the PID and MPC controllers, it is proven that the MPC controller settings with the 'quadratic programming method' provides a better control system as a whole to the simulated C-110 Debutanizer column. This method shows lesser fluctuations and the control system contains lower amount of noise. Therefore, from this project, this MPC controller setting is suggested to be used at the plant operation of PP(T)SB to achieve the desired outlet composition for the C-110 Debutanizer column. The findings of this project will help the operations at PETRONAS Penapisan Terengganu Sdn. Bhd., PP(T)SB in order to achieve the desired output of the C-110 Debutanizer column. There are two recommendations for this project. Different MPC tuning relations can be further studied and compared in order to achieve the most efficient control strategy that can maximize the Debutanizer product output. Different simulation software like iCON can be used to simulate the real plant environment.

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APPENDICES

Appendix I: MATLAB coding for quadratic programming method

```
% Declaration of transfer functions
g11=poly2tfd(6.593,[0.0001725 172.5 58.89 1],0,0);
g21=poly2tfd(1.526,[3.528e04 3272 99.88 1],0,3.87);
g12=poly2tfd(517.6,[7.934e04 5.214e08 1.598e09 1],0,0);
g22=poly2tfd(18.7,[7481 3.541e08 1.299e05 1],0,0);

delta=3; % Sampling time
ny=2; % Number of output
tfinal=90; % Internal Execution Time

% Define the model
model=tfd2step(tfinal,delta,ny,g11,g21,g12,g22);
plant=model; % Plant and model are the same

P=10; % Prediction Horizon
M=5; % Control Horizon
ywt=[]; % Weight of outputs
uwt=[1 1]; % Weight of inputs

tend=30; % Sampling time limit
r=[0 1]; % Set point for outputs

% Constraints of inputs and outputs
ulim=[-inf -0.15 inf inf 0.1 100];
ylim=[0 0 inf inf];

% Execution of process
[y,u]=cmpc(plant,model,ywt,uwt,M,P,tend,r,ulim,ylim);

% Output display
plotall(y,u,delta),pause
```

Appendix 2: MATLAB coding for no constrain method

```
delt=2;
ny=2;

g11=poly2tfd(6.593,[0.0001725 172.5 58.89 1],0,0);
g21=poly2tfd(1.526,[3.528e04 3272 99.88 1],0,3.87);
g12=poly2tfd(517.6,[7.934e04 5.214e08 1.598e09 1],0,0);
g22=poly2tfd(18.7,[7481 3.541e08 1.299e05 1],0,0);

umod=tfd2mod(delt,ny,g11,g21,g12,g22);
%Defines the effect of u inputs
g13=poly2tfd(3.8,[14.9 1],0,8);
g23=poly2tfd(4.9,[13.2 1],0,3);
dmod=tfd2mod(delt,ny,g13,g23);
%Defines the effect of w input
pmod=addumd(umod,dmod); % Combines the two model
imod=pmod; % assume perfect modelling
ywt=[]; % default (unity) weights on both outputs
uwt=[]; % default (zero) weights on both inputs
P=5; % prediction horizon
M=P; % control horizon
Ks=smpccon(imod,ywt,uwt,M,P);
tend=30; % time period for simulation
r=[1 0]; % setpoints for two outputs
[y,u]=smpcsim(pmod,imod,Ks,tend,r);
plotall(y,u,delt)
```