Model Predictive Control of Gas Processing Plant Focused on Depropanizer Column

by

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CERTIFICATION OF APPROVAL

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Approved by,

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TRONOH, PERAK

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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.

HOODA BIN-USENG

ABSTRACT

The objective of this project is to improve the energy efficiency and reduce the operation cost for gas processing plant focused on de-propanizer column by implemented the advance process control namely Model Predictive Control. In gas processing plant, 60% of energy used for chemical industries is from distillation processes. To improve the energy efficiency of distillation column for gas processing plant, model predictive control is one of technology introduced to the distillation process control system that will overcome this problem compare to conventional controller. In this project, a study 2x2 model predictive control which consist of two manipulate variable and two control variable for de-propanizer column of gas processing plant. By doing the model predictive controller implementation, plant model development which consists of steady state and dynamic model is required by using HYSYS simulation. Step test is necessary which will then calculate the transfer function by using MATLAB system identification for model predictive control design and implementation. And lastly, Comparison between model predictive control and a conventional controller is desired which shown that model predictive controller has better performance and small energy consumption compare to conventional controller.

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Bissmillahirrahmanirrahim, Alhamdulillah.

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NOMENCLATURE

APC	Advance Process Control
CO2	Carbon Dioxide
CV	Control Variable
DCS	Distrbuted Control System
DV	Disturbance Variable
IEA	International Energy Agency
MPC	Model Predictive Colntrol
MV	Manipulated Variable
PV	Process Model

CHAPTER 1
INTRODUCTION

1.1 Background of Study

Energy is a very important for social and economic development to increase agriculture and industrial activities in the country of any nation with together will also increase the demand of energy (Iniyan and Jebaraj, 2006).

Energy is an important as a part of life that we often use especially in form of electricity. In facts, most of electricity comes from the burning of fossil fuels like coal, gas or oil that can produces carbon dioxide. With the buildup of carbon dioxide in the atmosphere, the risk of global warming which known as the Greenhouse Effect is occurred. However, if the energy efficiency is used it means that the making better use for non-renewable natural resources. These will cause to saving energy and reduce the greenhouse gas emission.

In the chemical process industries, distillation is the most important part which quite energy intensive and account for approximated 3% of the world energy consumption. The energy consumption in distillation and carbon dioxide (CO_2) gases emission to the atmosphere are strongly related due to higher energy demand will cause larger CO_2 produces in atmosphere (Jana, 2010). To improve the energy efficiency in distillation column, many previous paper researches introduces advance process control technology to reduce energy consumption.

Advanced process control (APC) is a general term composed by using computer control algorithm that often used for solving multivariable control problems or discrete control problems. APC can be found in most petrochemical industries and refinery where multivariable control problems are possible to control. Since these controllers contain the dynamic relationships between variables, it can predict the behavior of the plant in the future. Actions of this prediction can be maintaining variables within their limits to prevent the excessive movement of the input. Normally an APC system is connected to a distributed control system (DCS). APC strategy called model predictive control will calculate moves that are sent to DCS for implementation in an optimal manner.

1.2 Problem Statement

In both chemical and petrochemical industries, distillation is the most important separation processes in product recovery and purification. 60% of energy used for chemical industries is from distillation (Diez, et al., 2009). Due to previous statement, energy is important for the distillation. The higher energy demand will cause the higher of CO_2 produces to atmosphere which will cause the global warming which known as Greenhouse Effect. To overcome these issues, conventional controllers and advance process control are introduced in many previous paper researches.

By using conventional controller, it is also applied to maintain the set point of the process but high energy is required for this reason. Besides, the process is difficult to adjust in order to get the product quality due to individually adjust in multivariable control problem. The excessive movement of manipulate variable migth be occurred to effect the product quality and cause to increase an operating cost and an energy consumption.

1.3 Objective and Scope of Study

To enhancing the optimization of gas processing plant, the study of model predictive control is important to develop and improve the process control of gas processing plant in order to achieve the target and propose of the researcher project. The main goal of this paper is considered as below:

1. To implement the MPC controller in gas processing plant focus on depropanizer column.
2. To improve the product quality and the energy efficiency of gas processing plant in order to reduce the CO_2 emission compare to PI controller.

Since many of research has been done in model predictive control in industrial case. Huang & Riggs, 2000 was applied both decentralized PI and MPC controls for a gas recovery unit via a computer simulation (ChemCAD). Aske, et Al., 2005 was implement of MPC on De-ethanizer Column at Karsto Gas Plant by using SEPTIC*MPC tool. The study of the effects of including levels in MPC controller in order to improve distillation control was implemented via DMCPlus software and simulator by Huang and Rigg, 2002.

This research is about the implementation of model predictive control of gas processing plant focus on De-propanizer column by using HYSYS software and simulation. And use MATLAB to calculate the action of MPC.

CHAPTER 2
LITERATURE REVIEW

This chapter will present the background of energy, type of energy resource, the effect of high energy consumption, energy in distillation column of gas processing plant and model predictive control (MPC) technology which is the new technology that has been introduced for the control system in distillation in order to improve energy consumption. And this chapter will also explain the concept of MPC with the example of MPC application and compare MPC with conventional controllers.

2.1 Energy

Energy is the capacity of a physical system to perform work. Energy exists in several forms such as heat, kinetic or mechanical energy, light, potential energy, electrical, or other forms.

By 2030, as the International Energy Agency (IEA) reports, in developing Asian countries, the energy use in an average growth rate of 3% compared with 1.7% for the entire global economy (IEA, 2007). Thus, energy demand is double expected in Asia in the next 20 years (Sovacool, 2009). With the increase in energy demand, CO2 produce in atmosphere will be increase which will cause the global warming as known as Greenhouse Effect.

Energy resources are classified into two categories which are fossil or non-renewables which are included coal, petrol, gas, gas hydrate and fissile material while renewables energy source are hydro, biomass, geothermal, solar and wind energy (Demirbas, 2010).

During the last several year, new concepts of energy planning and management have occurred such as decentralized planning, energy conservation through improved technologies, waste recycling, integrated energy planning, introduction of renewable energy source and energy forecasting (Iniyan and Jebaraj, 2006).

2.2 Natural Gas Energy

Karasalihovic, et al., 2003 stated that natural gas is daily replacing other fuels in residential and commercial sectors. Normally, natural gas is use in industry and power plants as well as in emerging markets such as transportation, cogeneration and cooling by favor with the resource availability, cost and environmental issues. The global primary-energy consumption of natural gas amounts to 23 % and increase steadily.

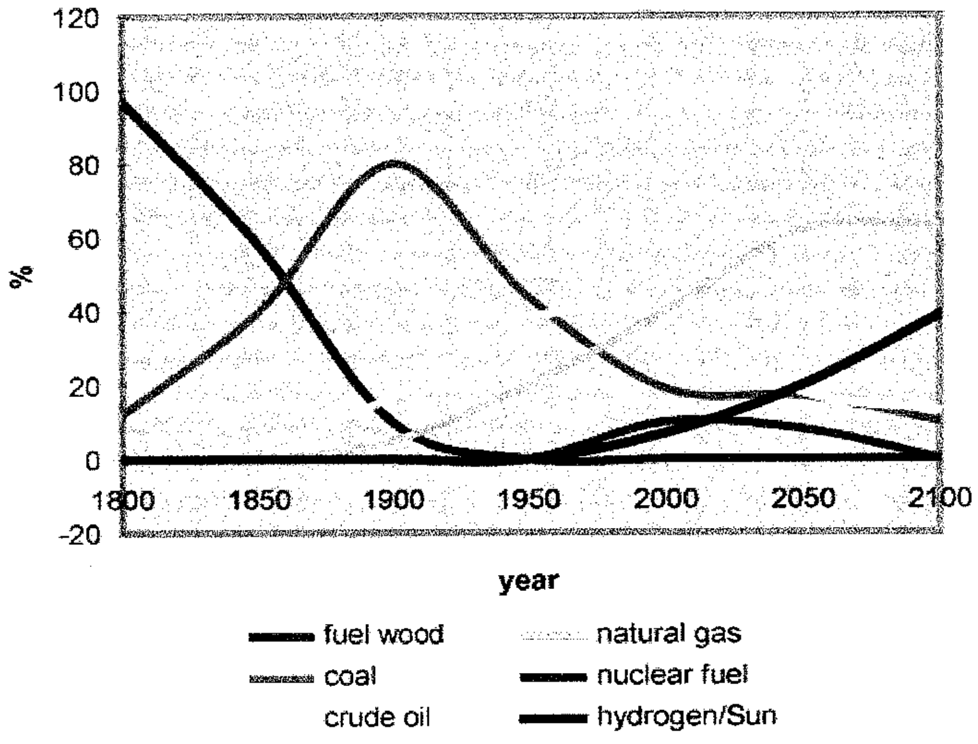

Figure 2.1: Particular fuel in world energy demand (Karasalihovic, et al., 2003).

The highest growth rates in natural gas demand are epected for the developing countries of the world, where the overall demand of natural gas is likely rise by 5% annually between 1995 and 2015 (as shows in Figure 2.1). Much of this growth will be used for electrical generation, industrial energy and also infrastructure construction likes cooking flue in major cities.

In industrialized countries, where natural gas market are most growth up which will also increase their confidence on natural gas. Over the next two decades, the industrialized countries demand is expected to grow by 2.6% annually, more than twice the rate of increase in oil use.

2.3 Energy Efficiency in Distillation Column

Distillation is important for chemical process industries. It is quite energy intensive and accounts for an estimated 3 % of the world energy consumption. It is the fact the energy consumption in distillation is strongly related with CO_2 gases produced in the atmosphere. With increase energy consumption will cause the larger CO_2 emissions to the atmosphere because mostly the energy is generated through the combustion of fossil fuel or non-renewable energy (Jana, 2010).

Distillation columns are used over 95% of the separation process in the chemical processing industries and it is also usually produced the final products in the chemical processing industries. As a result, product quality is usually determined by distillation control for the chemical processing industry (Enagandula and Riggs, 2006). Many studies recognized several of sector-specific and cross-cutting energy efficiency improvement opportunities. Innovative industrial technologies not only to reduce energy consumption, but also improve productivity, reduce capital cost, reduce operation costs, improve reliability as well as reduce CO_2 emission and improve working condition (Worrell and Price, 2001).

Thus, many of technologies discussed will improve the productivity and increase in a globalizing economy. Advance process control namely model predictive control is one of technologies control that is introduced for distillation column in order to improve energy efficiency, productivity, capital cost, operation cost and etc.

2.4 Model Predictive Control

Model predictive control (MPC) is an advance process control that usually uses to solve the multivariable control problem in order to predict the future response of a plant to achieve their optimal target. MPC algorithm use to optimize how the future plant behaves by computing the sequence of input. The first input is sent into control calculation while the other set of input will be repeated for entire calculation (Qin and Badgwell, 2002).

The classification of model types use in industrial MPC algorithms consist of 3 types which are (as shown in figure 2.1):

- Non-linear first principles models which are use by NOVA-NCL and PFC algorithm.
- Nonlinear empirical models which are use by Aspen target, MVC algorithm and process perfecter.
- And linear empirical models which are use by DMCplus, HIECON, RMPCT, PFC and SMOC algorithm.

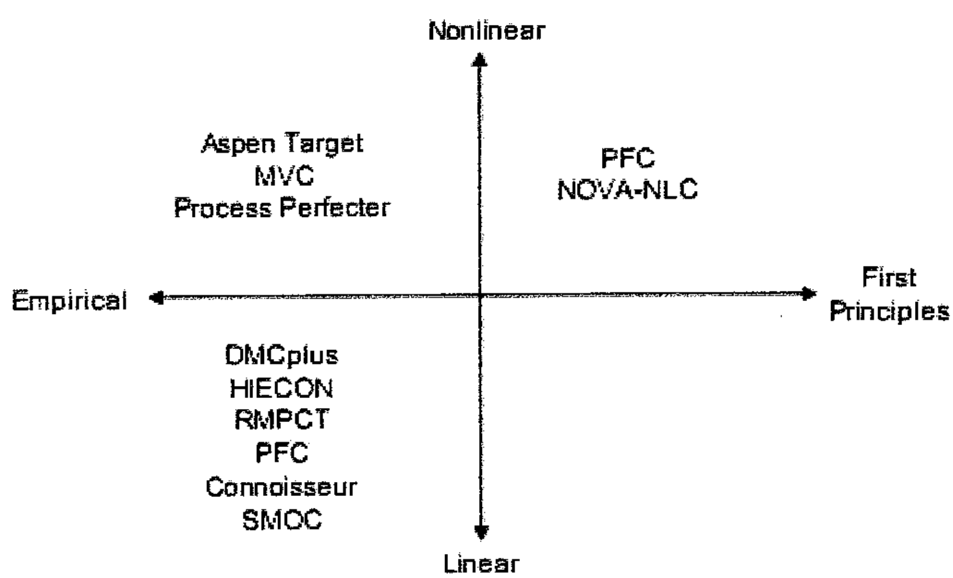

Figure 2.2: Classification of model types use in industrial MPC algorithms (Qin and Badgwell, 2002).

2.5 Principle of Model Predictive Control

Qin and Badgwell (2003) have been summarized the overall objectives of MPC controller that: MPC provide the input and output within constrains limit which can also move some control variable to their optimal target, while the other control variable still within their range. The movement of manipulated variable can be in control of their limit and it is also can be control process plant as much as possible when the sensor or actuator cannot detect in order to control the plant.

The purpose of the MPC control calculations is to consider a sequence of an input changes to predict the future output in order to achieve the optimal target or set point. Figure 2.3 shows that the basic concept of model predictive control which MPC calculates a set of M values of the input at the current sampling instant denoted by k. At each control move, the input will be constant. These inputs are calculated to give a set of predicted outputs achieves the optimal set point (target). The number of prediction P is call prediction horizon and the number of control move M is call control horizon. This concept is likely same with playing chess. Every time the chess move, the player should be predicted the strategy in order to get the best solution. The player will be move the chess again and again until get into the target.

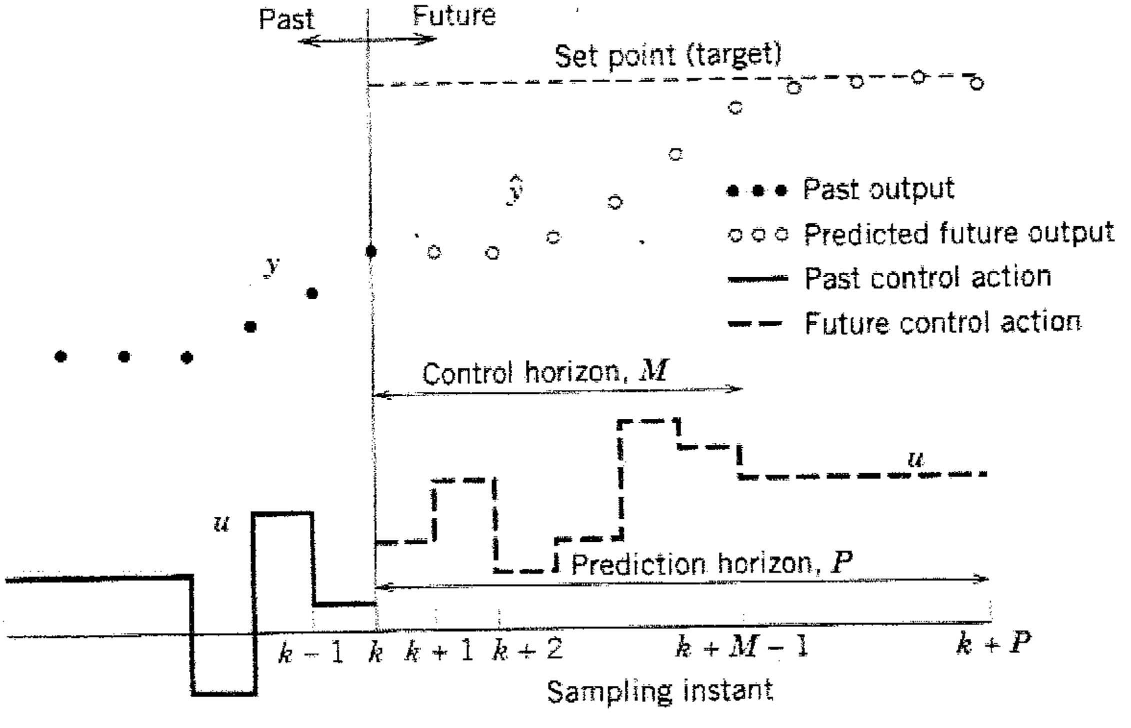

Figure 2.3: Basic concept for model predictive control (Edgar and co-workers, 2004).

2.6 Example of MPC Applications

There are several example of model predictive control implementation in gas processing plant such as comparison PI and MPC for gas recovery unit, implementation of MPC on a de-ethanizer at Karsto gas plan and include levels in MPC to improve distillation control.

2.6.1 Comparison PI and MPC for Gas Recovery Unit

For this application, Huang & Riggs, 2000 was applied both decentralized PI and MPC controls for a gas recovery unit which consists of three distillation columns operated in series: a de-ethanizer, a depropanizer and a debutanizer (as shown in figure 2.4) via a computer simulation (ChemCAD) in order to compare PI and MPC controllers.

The implementation of the decentralized controls was presented by considered the configuration consideration for the quality controls, constraint handling and tuning PID controllers. Then the comparison of three different MPC control implementation which is use PI controls the level control loop closed without MPC control, the level control loop closed with MPC move set point to level controller and direct MPC control the level by moving the bottom flow rate without the level control loop.

By compared between decentralized and MPC controls due to adjusting multi-manipulated variables to maintain the operation within the constrain limit, the comparison was found that the MPC controllers have an economic benefit compared to conventional controllers. For three different MPC implementations, the result was found that when input variable for level control has effect on product composition, all can improve control performance. But the MPC move set point to level controller has advantage of easy for step test and tuning.

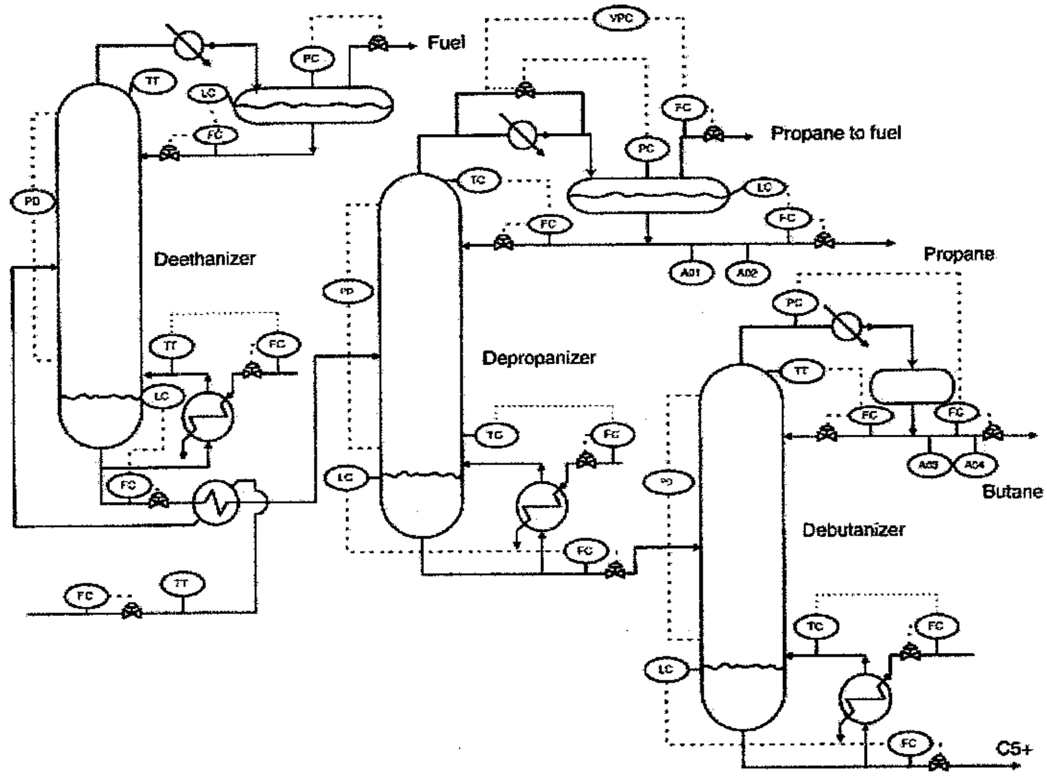

Figure 2.4: Process diagram of the gas recovery unit (Huang and Riggs, 2000).

2.6.2 Implementation of MPC on a De-ethanizer at Karsto Gas Plan

This application shows the implementation of MPC on De-ethanizer Column at Karsto Gas Plant by using SEPTIC*MPC tool which include with design, development of estimator, development of model and MPC tuning. By introducing the MPC algorithm to De-ethanizer Column, variation of product quality can be reduce for both top and bottom product compared to the operation before MPC implementation as shows in Figure 2.5.

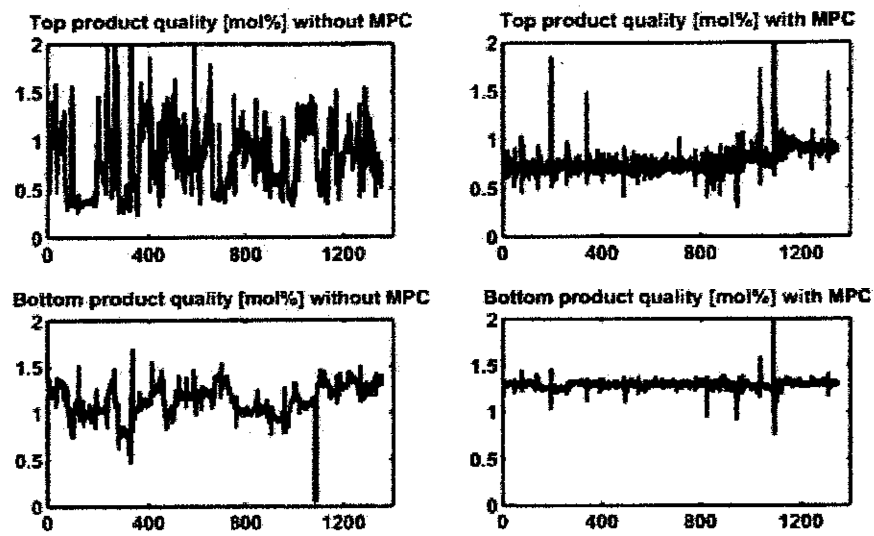

Figure 2.5: Trains of product quality of top and bottom product from the column without (left) and with (right) MPC (Aske, et al., 2005).

Figure 2.6 shows the De-ethanizer column with PID controller which control reflux drum level, reflux flow, bottom column level, tray 1 temperature, column pressure and LP steam pressure control before apply MPC modeling. These PID controllers give the large variation of both top and bottom product quality due to the disturbance of feed. The temperature set point and reflux flow rate are not easy to get the right value. This is because the temperature of column and reflux flow rate are changed with feed flow and feed composition which is difficult to control. The operator must be aware and proper adjust the temperature and reflux flow many times in order to get the right values (Aske, et al., 2005).

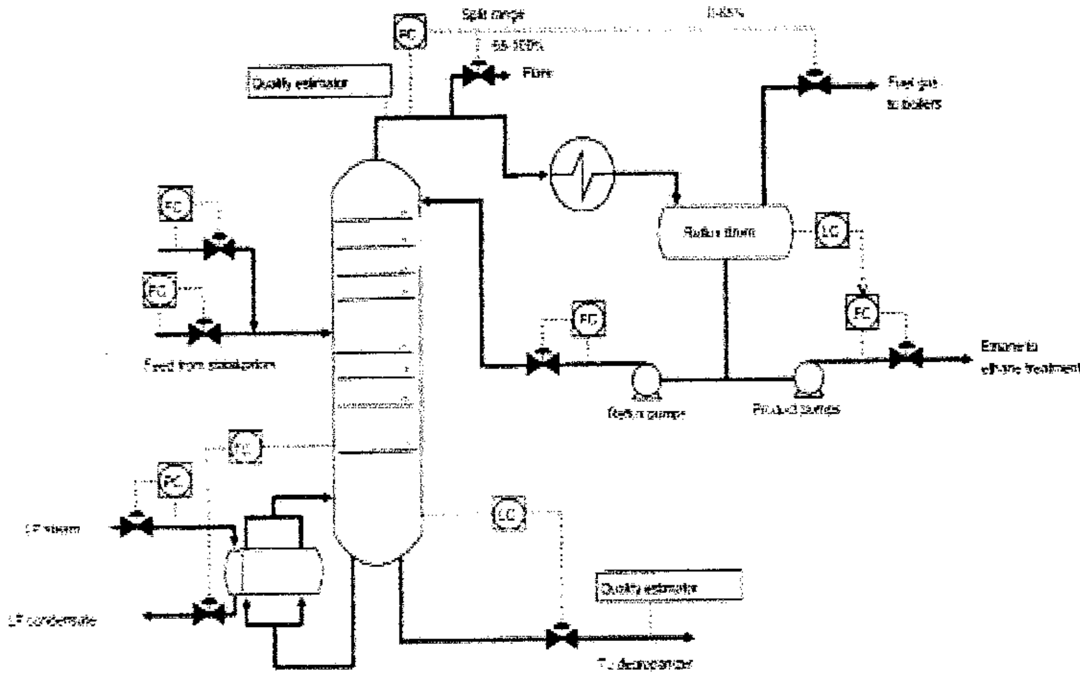

Figure 2.6: The de-ethanizer with basic controller (Aske, et al., 2005).

2.6.3 Include Levels in MPC to Improve Distillation Control

The study of the effects of including levels in MPC controller in order to improve distillation control was implemented via DMCPlus software and simulator of two columns which are a depropanizer and a propane/propylene splitter (C3 splitter) column. These two columns are a four by four system (four inputs and four outputs) which are reflux flow (L), distillate flow (D), hot steam flow (V) and bottom flow (B) as shown in Figure 2.7.

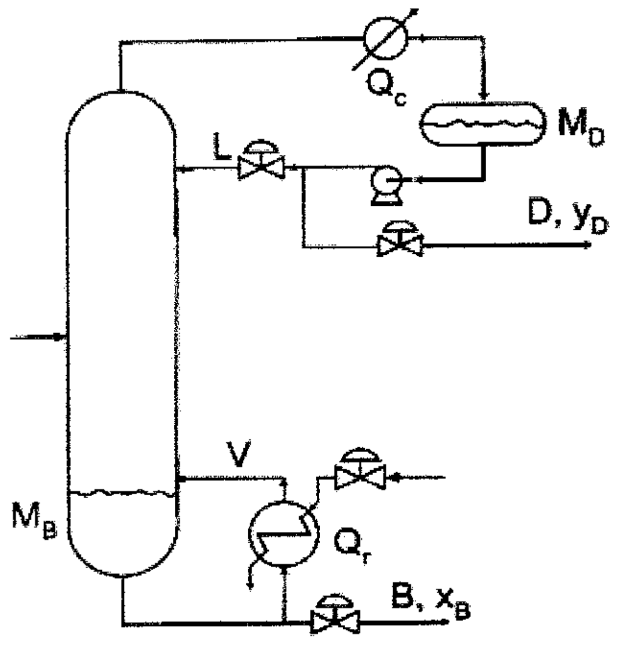

Figure 2.7: Two-product column (Huang and Rigg, 2002).

Three different MPC implementation (as shows in figure 2.8) were compared which are two regular MPC implementation for bottom and reflux flow and using PI controller for level control, direct level control by MPC directly in manipulating flow rate and cascade implementation by MPC moves the set point to the level controller.

The result was shown that both direct and cascade ensures the MPC controller to move all four manipulates in order to improve level control and composition control. The cascade implementation shows the improvement of both depropanizer and splitter where the direct MPC controller is performed well only in depropanizer case. Actually the direct MPC controller has a high reliability because it is independence from regulatory level controller and it should be perform better than cascade. But the direct MPC controller is difficult to apply step test and ill-condition is introduced for a certain case while this situation is not occurred in the cascade (Huang and Rigg, 2002).

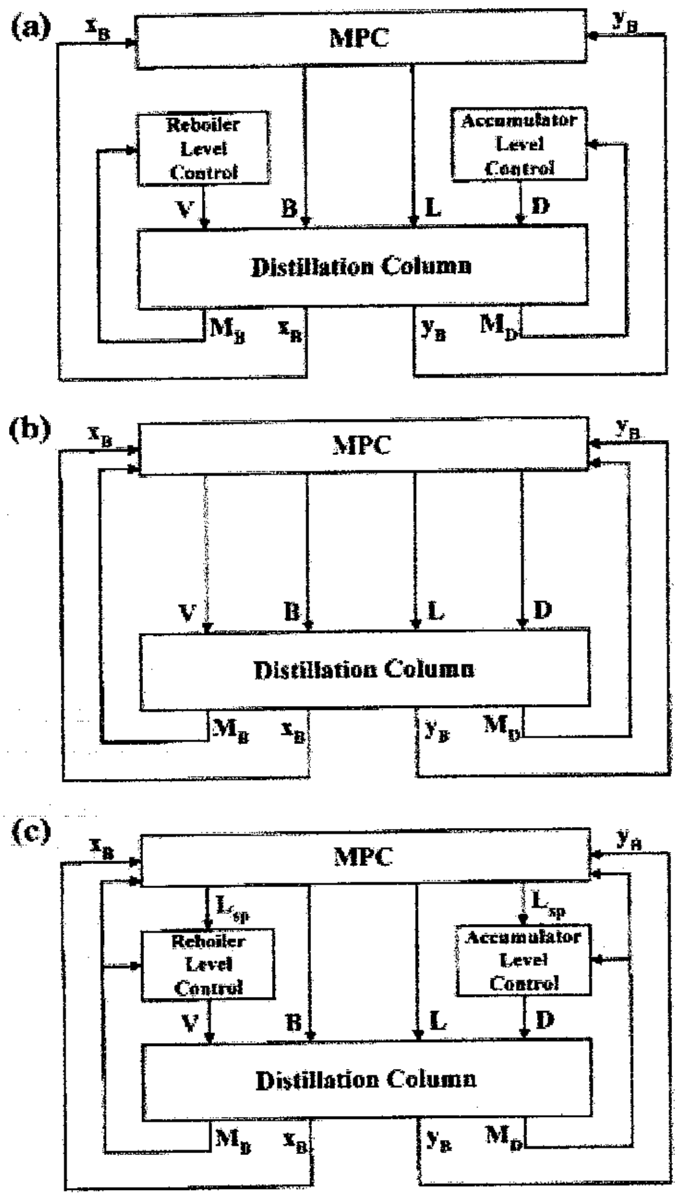

Figure 2.8: MPC implementations a) regular MPC strategy for [L, B] configuration, b) direct MPC for level control, c) MPC through cascade level control for [L, B configuration] (Huang and Rigg, 2002).

2.7 Conventional versus MPC controller

MPC algorithm are introduced in multivariable control instead of using conventional controller because the conventional controller controls the variable separately which is difficult to control and conflict between two output might be occurs. By using

MPC controller, the process variable can be now control all variable together which can move the variable within their constrain limit to prevent the excessive movement of input variable. A small change output constrain in the MPC have the effect by a small change in input constrain, this action can be shown that the operating cost of plant can be reduce and cause to reduce in energy usage. The figure shows the comparison of conventional control structure and MPC control structure.

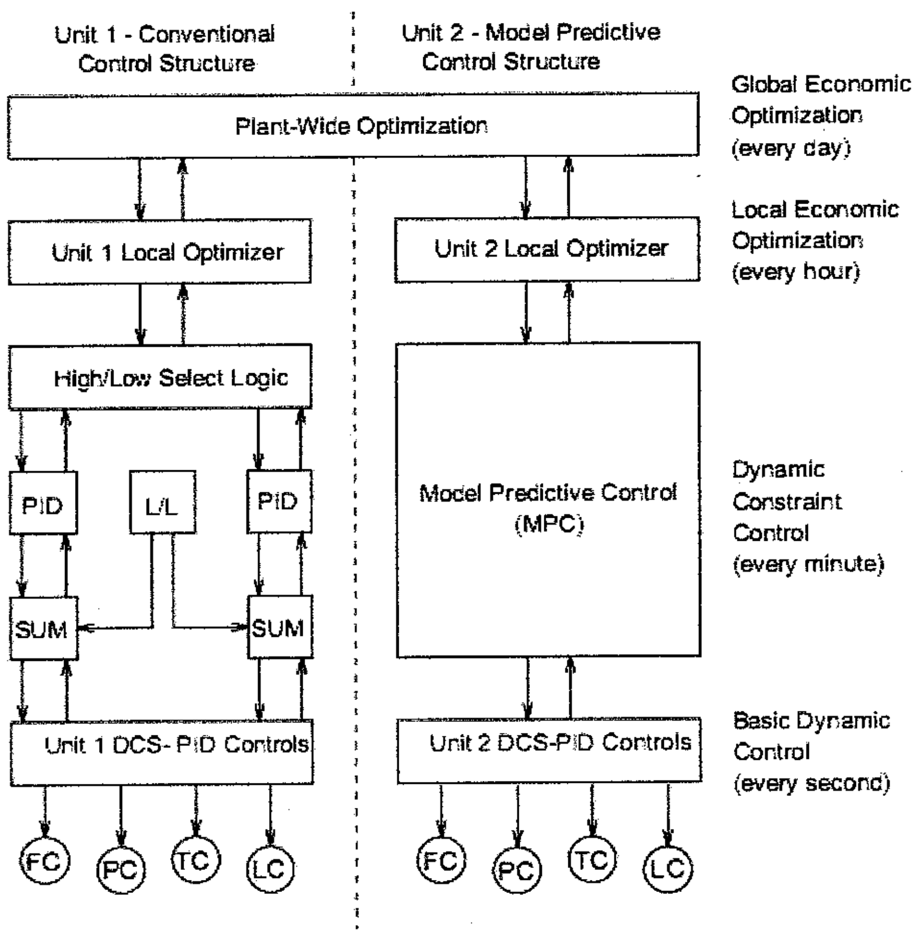

Figure 2.9: Control structure for conventional controls (at the left) and MPC controls (at the right) (Qin and Badgwell 2002).

Energy is very important as a part of daily life as well as in gas processing plant. With increase in energy consumption, CO2 release in the atmosphere will be increase. The excessive of CO2 in atmosphere is the main cause of global warming. Many of research paper are doing on model predictive control applies to distillation column in order to improve their product quality and energy efficiency instead of using conventional controllers. For the project research methodology including the step of MPC implementation will be discuss in the next chapter.

CHAPTER 3
METHODOLOGY

This chapter will provide the information about the methodology used in project and briefly explain the project activities based on project research methodology. The tools required in order to develop the project as well as the schedule of the project in form of Gantt chart will also present at the end of this chapter.

3.1 Project Research Methodology

For the project methodology, the project will start with literature review follow by plant model development which consists of steady state and dynamic model by using HYSYS process flow diagram. In the next step, APC design implementation which are involve with plant testing, APC design and APC implementation are introduced and lastly, the comparison with base layer control are implemented.

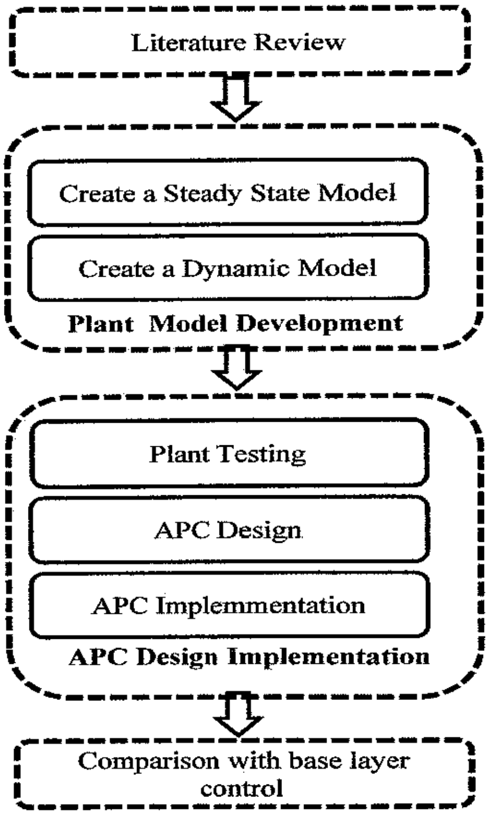

Figure 3.1: Project Methodology.

3.2 Project Activities

3.2.1 Literature Review

For the literature review, first of all, the researcher needs to research through available works that have been developed by many researchers around the world which showed the recently knowledge and technology that relates to the project. In this paper, the research of technologies provide for reduce the energy consumption for gas processing plant are needed. The technology that is focused in this paper is advance process control namely model predictive control.

3.2.2 Plant Model Development

For the simulation work will start with steady state model of unit operations which is de-propanizer column as shows in Figure 3.2 with consist of 23 number of stage, stage 16 is feed location with identify the composition, temperature, pressure and molar flow of feed stream to simulate, and condenser and reboiler are simulated a propane product purity in Aspen HYSYS software. Once the steady state model is set up, the sizing of unit operation, specification of flow/pressure condition at boundary streams and installation of controller are needed in order to prepare the simulation of dynamic model.

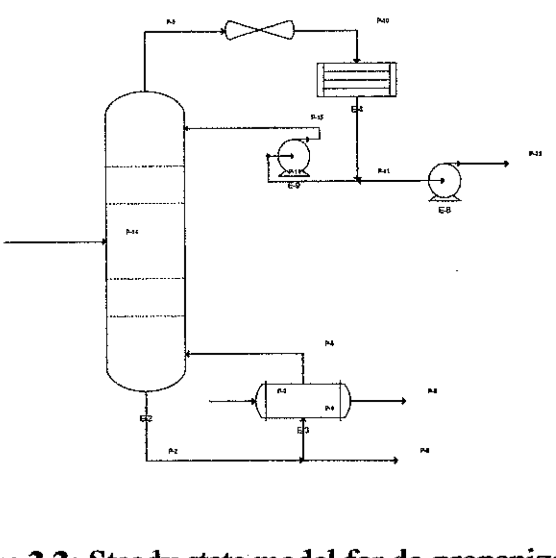

Figure 3.2: Steady state model for de-propanizer column.

3.2.3 APC Design Implementation

APC design implementation is the next step for the research methodology after steady state and dynamic model are available. For the APC design implementation, Figure 3.3 shows the flow chart of MPC calculation modified from Qin and Badgwell, 2003. There are seven steps including in MPC calculation.

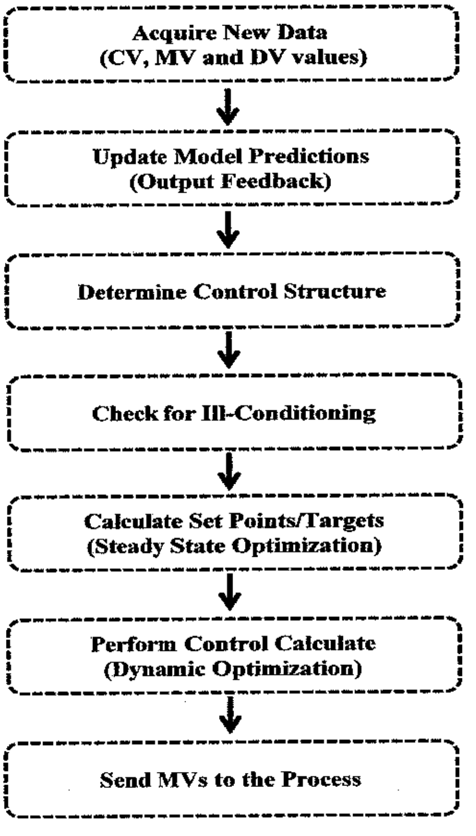

Figure 3.3: Flow chart for MPC calculation (Qin and Badgwell, 2003).

In step 1 is obtain the current value of manipulated variable, disturbance variable and process variable from Distribute Control System (DCS). Then new prediction output will be calculate by using process model and the current value of control process (MV, DV and CV) in order to update model predictions. In step 3, control structure should be determine to make sure the manipulate variable can be proper manipulated and control variable should be control. If manipulate variable is disable to control valve then this manipulate variable cannot be used to control but it can be served to be a disturbance variable. Ill- conditioning should be considered in the next step before move to step 5 and 6. Ill- conditioning can be occurred when the effect of input on two or more output are too small.

Next step is calculated the set point or target of the process. Control calculation should be performed in order to move the process to their set point. For the last step of MPC calculation is sent the manipulate variable to the process for control calculation to move the process into the target at DCS level. Any error that might be occurs for MPC calculation, the review back of literature review is needed in order to make sure understanding of MPC and the accuracy of the process model should be revising for the success of MPC implementation.

3.2.4 Comparison between MPC and Base Layer Control

For the last step of this project, the result will shown the energy efficiency of gas processing plant on distillation part and the comparison between MPC and Base layer controller is needed in order to compare which controller is covered more on energy efficiency.

3.3 Requirement Tools:

3.3.1 AspenTech HYSYS Dynamics

AspenTech HYSYS dynamic software is use to implement the MPC controller and create plant model development which included with steady state model and dynamic model.

3.3.2 MATLAB

By solve the MPC calculation, MATLAB is needed in order to calculate the transfer function of MPC to implement the MPC controller in HYSYS simulation.

However, these two software will link together in order to implement the MPC controller to improve the process control of gas processing plant in the way of energy efficiency.

3.4 Project Gantt Chart:

Table 3.2: Project gantt chart for FYP1.

Activities	Months						
	Jan	Feb	Mar	April	May	June	July
1. Literature Review.							
2. Model Development of De-Propanizer Column.							
- Steady State.							
- Dynamic.							
3. Report Writing.							

To make the project run smoothly and will be finish on time, gantt chart is needed. For the FYP 1 progress, literature reviews are needed to study for the researcher to make sure the understanding on project throughout the semester. For the steady state model simulation, the model will use a maximum one month in order to finish by June. After steady state model is simulated, the dynamic model is the next step by using maximum two months which expect to finish by the end of July.

Table 3.3: Project gantt chart for FYP2.

Activities	Months				
	Jul	Aug	Sep	Oct	Nov
1. Plant Testing.					
2. APC Design.					
3. Simulation and APC Implementation.					
4. Comparison with Base Layer Control.					
5. Report Writing.					

For FYP 2 planning progress, APC design and implementation are planning to finish within two months. For APC design will be focused from August until September. After APC design are settle, APC implementation is the next step which expected to finish within October.

CHAPTER 4
RESULT AND DISCUSSION

In this chapter, process description of the plant, steady state and dynamic model have been described. The result of step test and the comparison of PI and MPC controller by using two methods which are disturbance rejection and set point tracking have been discussed in this chapter.

4.1 Process Description

In real gas processing plant, there is consist of many process unit which are important such as mercury removal unit, dehydration unit, acid gas removal, NGL recovery and fractionation unit in order to get the specific product required such as Sales Gas and Liquid Gas Petroleum as shown in the figure below.

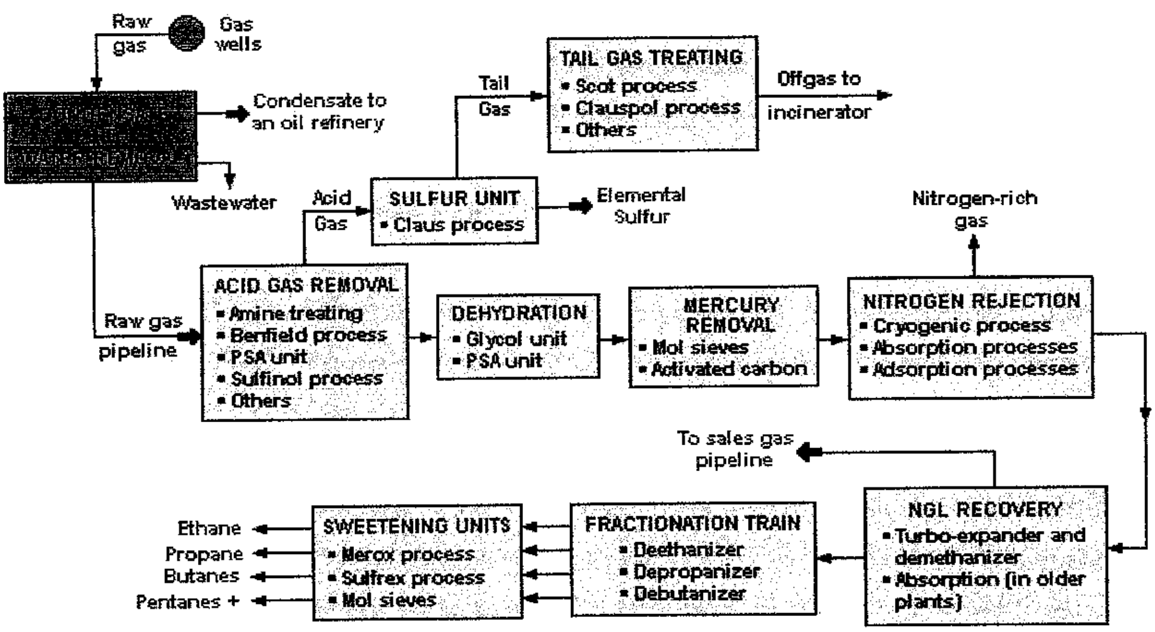

Figure 4.1: Process flow of natural gas processing plant.

The De-propanizer column is one of the important unit uses to separate propane and lighter composition from the feed gas by using a different of each boiling point to produce propane product purity. This column consist of ten numbers of stage with stage five is feed location. Condenser is another unit operation to remove heat to condense gas into liquid to recycle back to the top of the column. To heat the feed gas, reboiler is needed in order to heat a feed gas at a proper temperature to separate the required product.

4.2 Steady State Model

To start the simulation part, steady state model is required. Stream and unit operation are installed in process flow diagram of HYSYS software as figure 4.2. For this research, the basic distillation column is used with already consist of condenser and reboiler with follow the condition of table in the figure below. And feed composition as table 4.1:

Composition	Mole fraction
Nitrogen	0.001947
CO2	0.004502
Methane	0.234483
Ethane	0.252815
Propane	0.259128
i- Butane	0.125903
n-Butane	0.121222

Table 4.1: Feed gas composition.

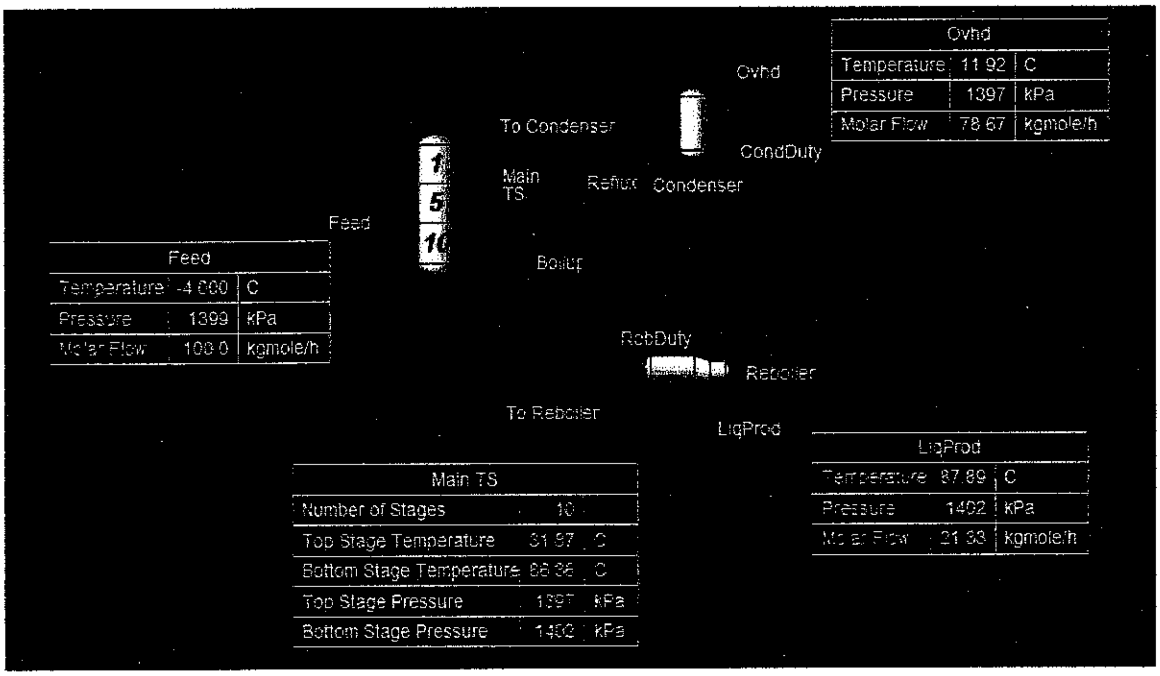

Figure 4.2: Steady state model.

4.3 Dynamic Model

After steady state model is converged, next step is dynamic model development. Three steps are required which are sizing the equipment in order to get a realistic model, flow or pressure specification of boundary stream and add the controller as figure 4.3. Make sure the dynamic process is stable in order to move on to the next step which is step testing.

For De-propanizer column, only internal part that need to be sized. The tray or packing type should be specified. For this De-propanizer column, tray type has been choosing. The dimensions such as tray spacing, tray diameter, weir length and weir height should be indicated.

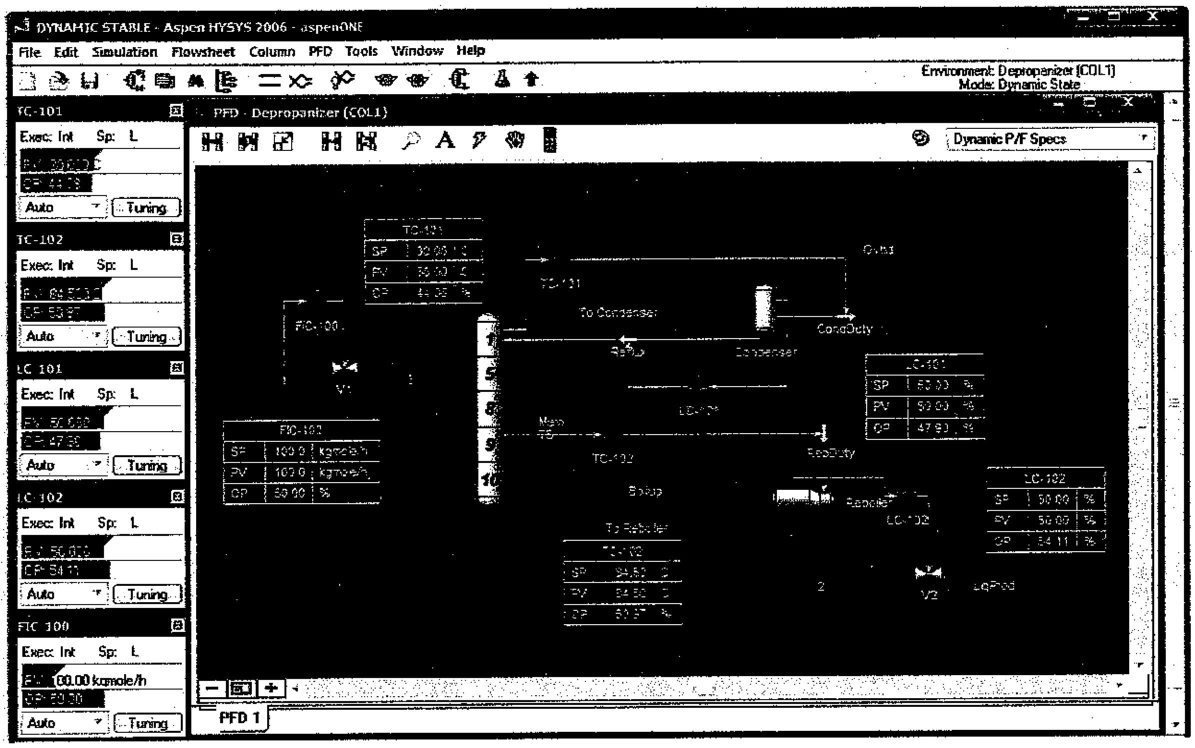

Figure 4.3: Dynamic model.

4.4 Step Test

To install MPC controller, step test is required to see the response of output variable when input change is increase and decrease. And to make sure the process is in stability. Step test is to measure the dynamic responses which will again required for MATLAB system identification toolbox to calculate the transfer function by using first order plus time delay (FOPTD) dynamic model.

For 2x2 model predictive control consist of two input variables which are condenser duty and reboiler duty and two output variables which are impurity i-butane overhead product composition and stage ninth temperature Table 4.2 and 4.3 shows the input move of each input variables.

Input Move TC101	
%OP Before	%OP After
45.89	50.89
45.89	55.89
45.89	40.89
45.89	35.89
45.89	50.89
50.89	55.89
45.89	40.89
40.89	35.89
45.89	45.89

Table 4.2: Input move of condenser duty, u1.

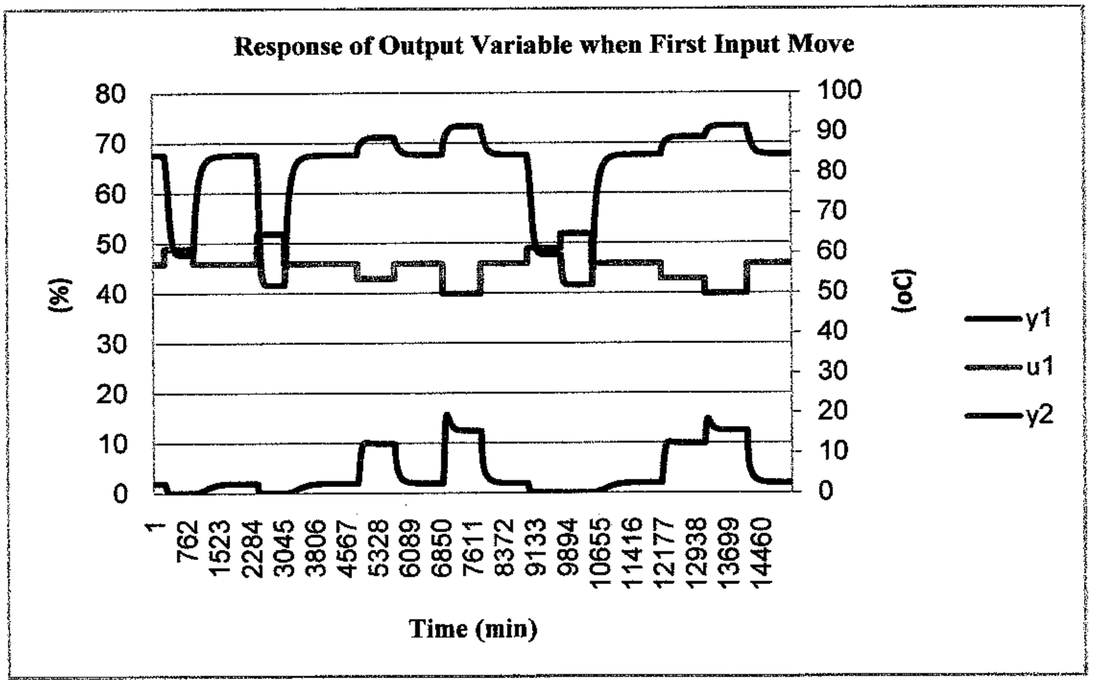

Figure 4.4: Response of output when first input move, u1.

From figure 4.4, an initial percent valve open at 45.89% which have i-Butane Composition overhead product is 1.962% and stage ninth temperature is 84.50 °C as the set point of this process. By open valve TC 101 which is increase in condenser duty, heat will be removed to condense more liquid and will cause column temperature decrease as well as stage 9th temperature. As our product is in gas phase, propane is more condense into liquid in order to reflux back to the column and will cause the overhead propane product is decrease in composition. i-butane is heavier component than propane. As decrease in propane product, it causes i-Butane product overhead decrease as well. By the way, when decrease in valve opening, less heat will be removed from condenser which will cause to increase the column temperature and stage ninth as well. As decrease condenser duty, the propane is also decrease to condense into liquid. It cause propane and i-Butane are increase at the overhead product. By increase in temperatures, the impurity of i-butane will boil up to the overhead product which will cause i-Butane at the overhead product is increase.

Input Move TC102	
%OP Before	**%OP After**
51.46	56.46
51.46	61.46
51.46	46.46
51.46	41.46
51.46	56.46
56.46	61.46
51.46	46.46
46.46	41.46
51.46	51.46

Table 4.3: Input move of reboiler duty, u2.

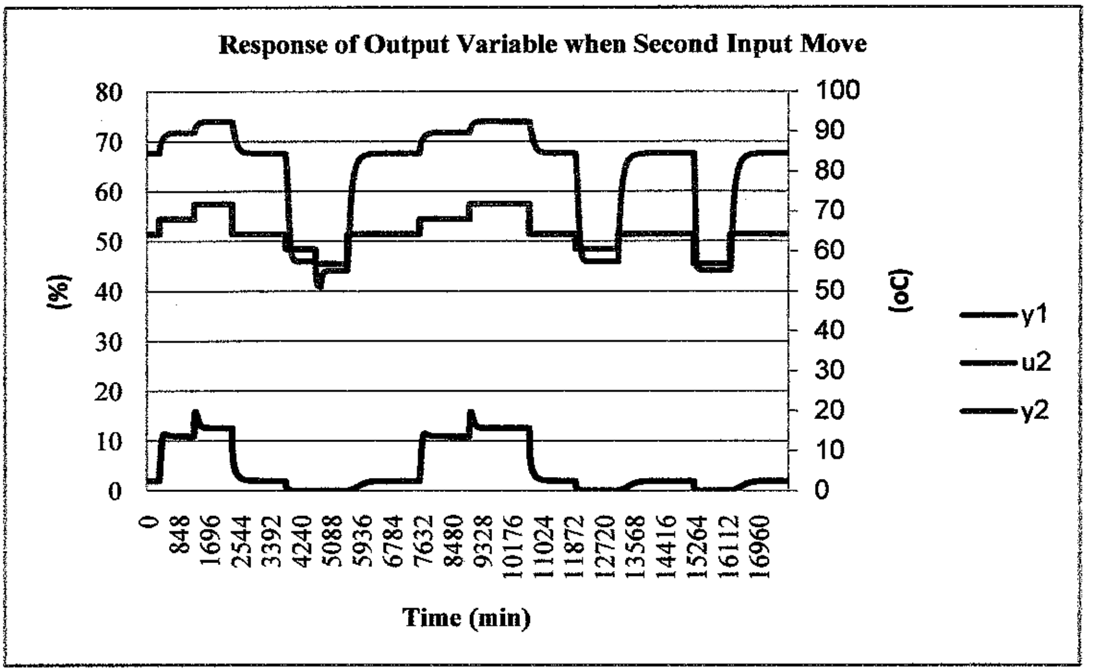

Figure 4.5: Response of output when second input move, u2.

As our i-butane product composition overhead product set point is 1.962% and stage ninth temperature is 84.50 °C with percent opening of valve TC 102 (Reboiler duty) is 51.46%. Figure 4.4 shows the response of output when reboiler duty percent opening valve moves. As increase in valve opening, heat in reboiler is increase to boil up the gas into the column and will cause stage ninth temperature and top stage temperature is increase. As increase in both temperatures, it will cause i-Butane overhead product composition increase due to most of lighter carbon will boil up to the overhead product like propane composition. Thus, the impurity i-Butane product boils up to increase the composition of i-butane at the overhead product. To decrease reboiler duty, temperature of boil up gas will be decrease and will cause the top stage temperature and stage ninth temperature is also decrease. With decrease the temperature, propane composition will be decrease due to less propane component to be vaporized which cause the propane composition will still remain in the bottom product as well as the impurity i-butane still in the bottom due to i-butane component is lighter than propane component.

4.5 MATLAB System Identification Toolbox

After step test data is recorded, MATLAB system identification is required in order to calculate the transfer function of 2x2 model predictive controls by using first order plus time delay (FOPTD) model. The theoretical method of the model parameters of 2x2 transfer functions which consist of process gain, time constant and time delay are obtained as follow:

1. Process gain, $Kp = \dfrac{\text{Steady state changed in measured process variable,}\Delta\text{PV}}{\textit{Steady state changed in controller output,}\Delta\text{CO}}$
2. The dead time, Θ_p

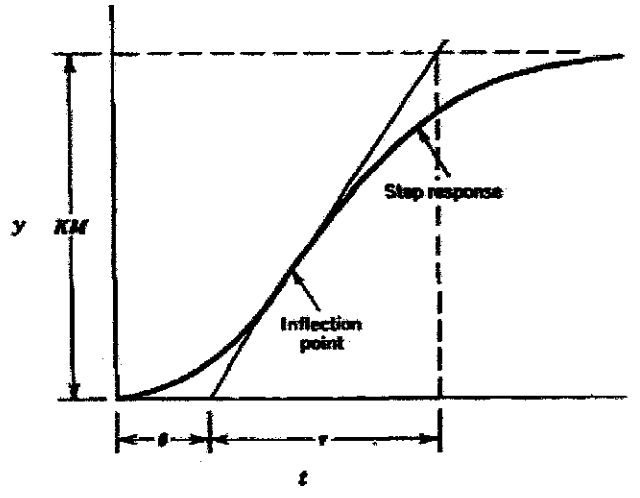

Figure 4.6: Graphical analysis of the process reaction curve to obtain parameters of a FOPTD.

3. Time response, τ_p

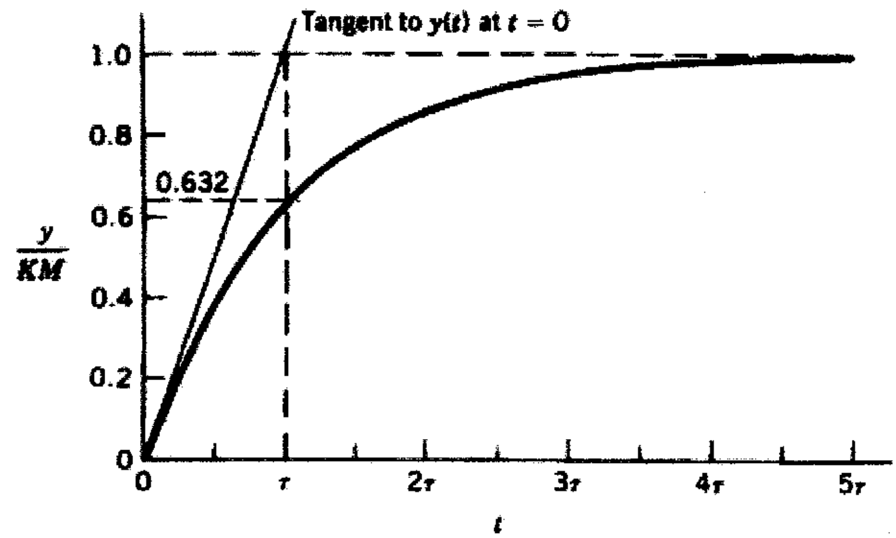

Figure 4.7: Graphical Constructions Used To estimate The Time Constant.

By using system identification (MATLAB), model parameter can be obtained as follow:

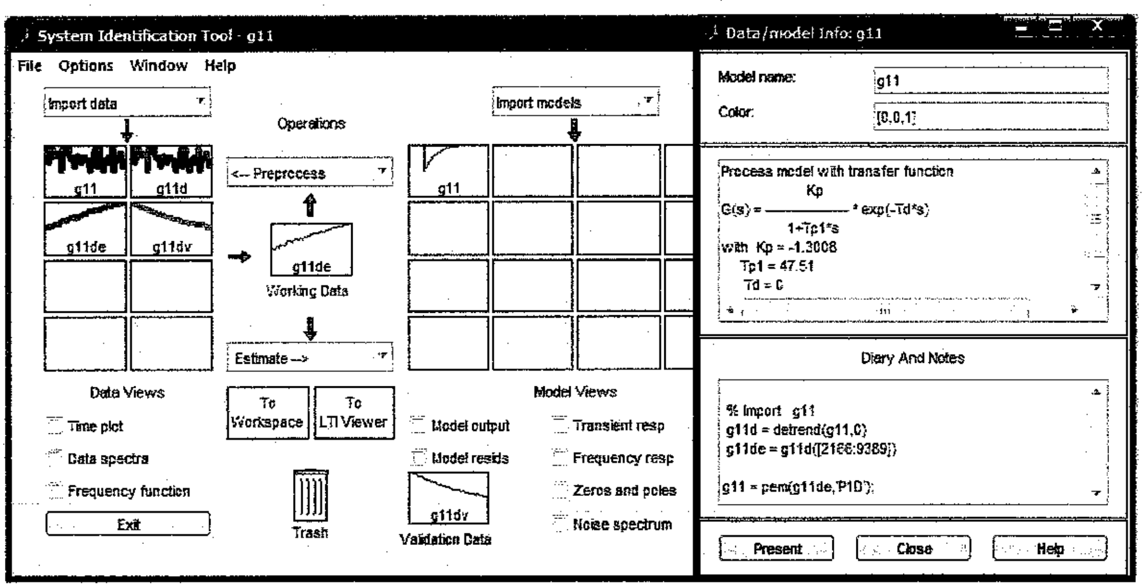

Figure 4.8: MATLAB system identification toolbox

Once model parameter is obtained, the transfer function of MPC will create based on FOPTD model:

$$g_{ij} = \frac{K_{p,ij} e^{-\tau_{d,ij} s}}{\tau_{p,ij} s + 1}$$

g_{ij} is the transfer function relate to output y_i and input u_j. $K_{p,ij}$, $\tau_{p,ij}$ and $\tau_{d,ij}$ are process gain, time constant and time delay respectively. Process gain, time constant and time delay of the system of 2x2 MPC are shown in the table 4.4 below:

Transfer Function	**Model Parameters**		
	K_p(°C/%)	**τ_p(min)**	**τ_d(min)**
G11	-1.3008	47.5100	0.0000
G12	1.0982	65.8750	0.0000
G21	-3.4813	90.0890	5.6126
G22	3.5539	79.1740	0.0000

Table 4.4: FOPTD model parameter.

4.6 MPC Controller

After the transfer function of MPC is obtained, the next step is to install MPC controller by using HYSYS simulation. To install MPC controller, the connection to process variable and input variable should be connected. Once connect the process and input variable, the operation parameter should be specified to get the set point of the process. Transfer function that calculated from system identification is now put in the process models tab to create the process model of MPC. To run the MPC controller, auto mode of MPC controller is set and the other two TC101 and TC102 is set in off mode.

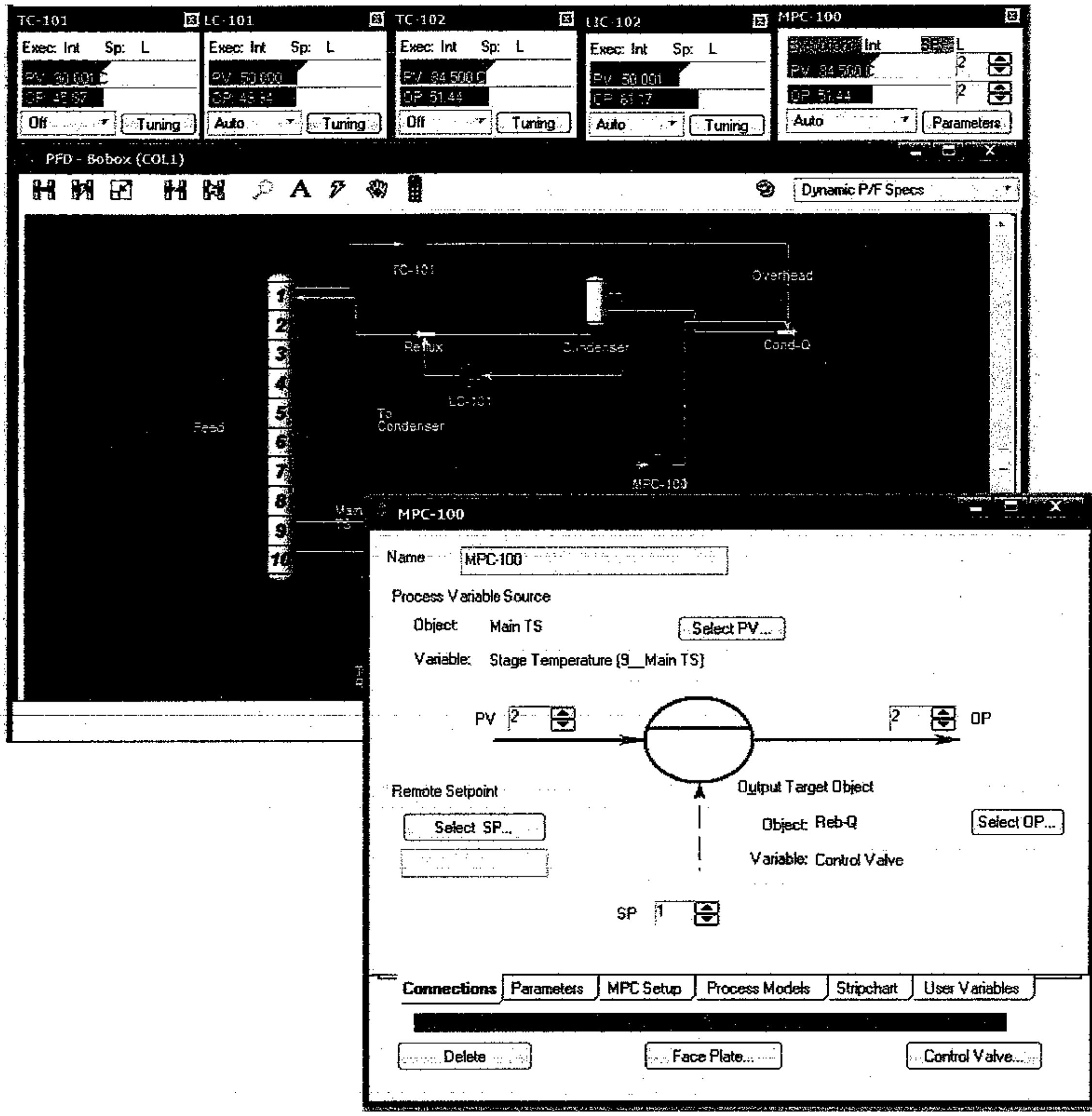

Figure 4.9: MPC controller installation.

4.7 Disturbance Rejection

There are many methods to compare the performance of PI and MPC controllers. Disturbance rejection is one of the methods to compare the performance of PI and MPC controllers. For disturbance rejection assessment is to compare the performance in term of the ability to maintain the i-butane overhead product composition and stage ninth temperature of both controllers by introduce noise disturbance at the feed from 5-10 %. The introduce noise disturbance start after 500 min.

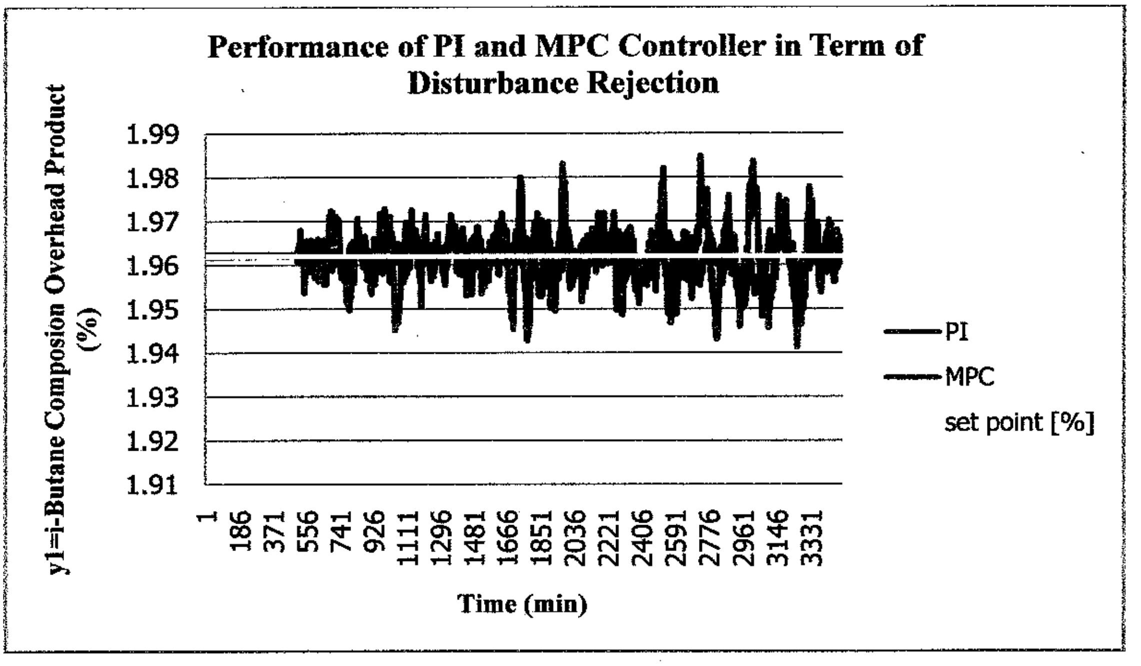

Figure 4.10: Performance of MPC and PI controllers for i-butane overhead product composition, y1.

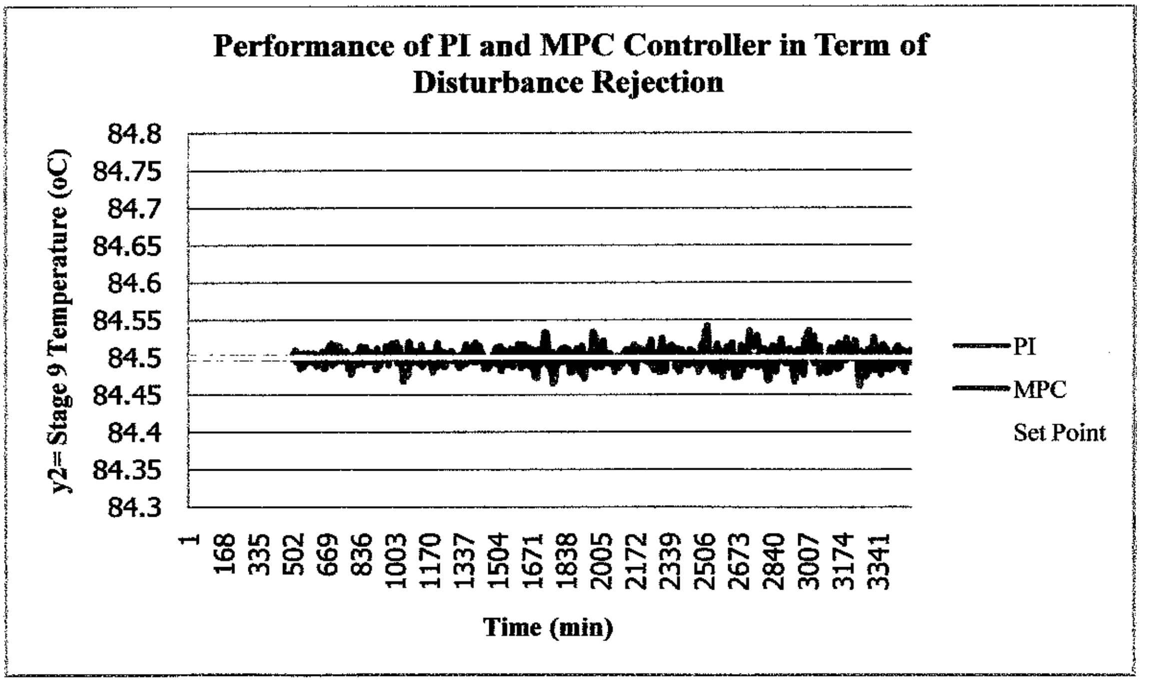

Figure 4.11: Performance of MPC and PI controllers for stage ninth temperature, y2.

From figure 4.10 shows the performance of PI and MPC controllers based on the ability to maintain i-butane overhead product composition. The initial time 500 minute is steady at design set point of 1.962. After 500 minutes, noise disturbance is introduced at the feed from 5-10 %. From the graph shows that both PI and MPC can maintain the i-butane overhead product nearly the set point. From figure 4.11 shows the performance of PI and MPC controllers based on the ability to maintain stage ninth temperature. Noise disturbance is introduced in the feed from 5-10% after 500 minutes as well. The graph shows that both PI and MPC can maintain the stage ninth temperature near the set point which is 84.5. Thus, it means that both controllers able to handle the disturbance rejection. But from the performance of these two graphs, the high oscillation or deviation from set point in PI controller is higher than MPC. Thus, MPC is smaller error compare to PI which means that MPC can be maintain the product quality better than PI controller.

4.8 Set Point Tracking

Set point tracking is another method to compare the performance of PI and MPC controllers by changed the set point of i-butane overhead product composition and stage ninth temperature that these controllers can move to the new set point or not. For set point tracking, some case studies have been tested as shown in the table below:

Case	y1	y2
1	1.962	83.5
2	1.862	84.5

Table 4.5: Case study of set point tracking.

4.8.1 Performance of the Process Variable

For case 1, the set point of i-butane overhead product composition is maintain in 1.962 % while change the set point of stage ninth temperature from 84.5 °C to 83.5 °C for MPC controller. Since PI controller cannot control the composition, PI controller is control in top stage temperature with initial set point 30 °C instead of i-butane overhead product composition with actually related each other.

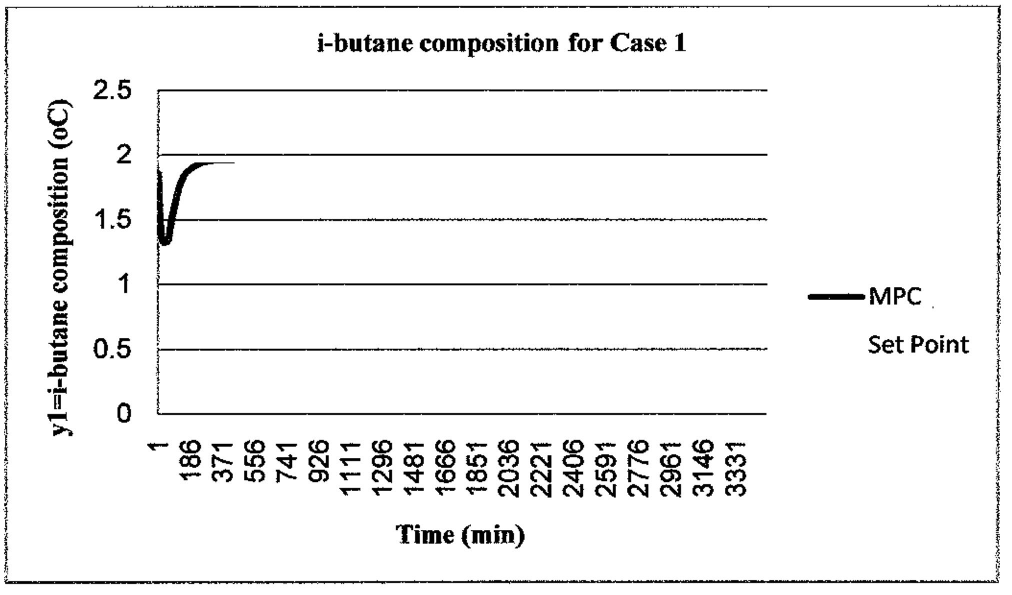

Figure 4.12: The performance of i-butane overhead product composition of MPC controller for case 1.

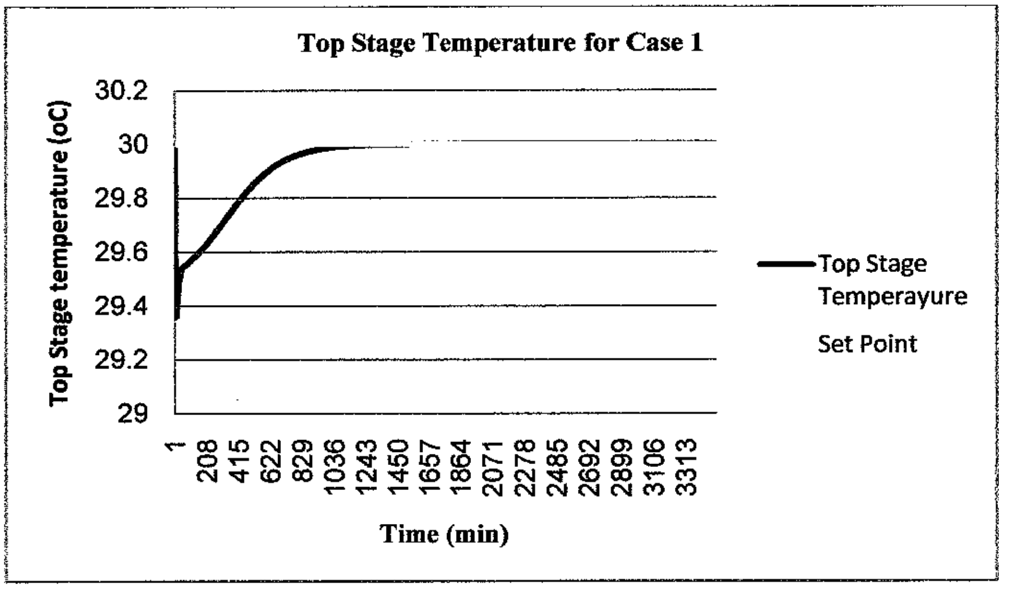

Figure 4.13: The performance top stage temperature of PI controller for case1.

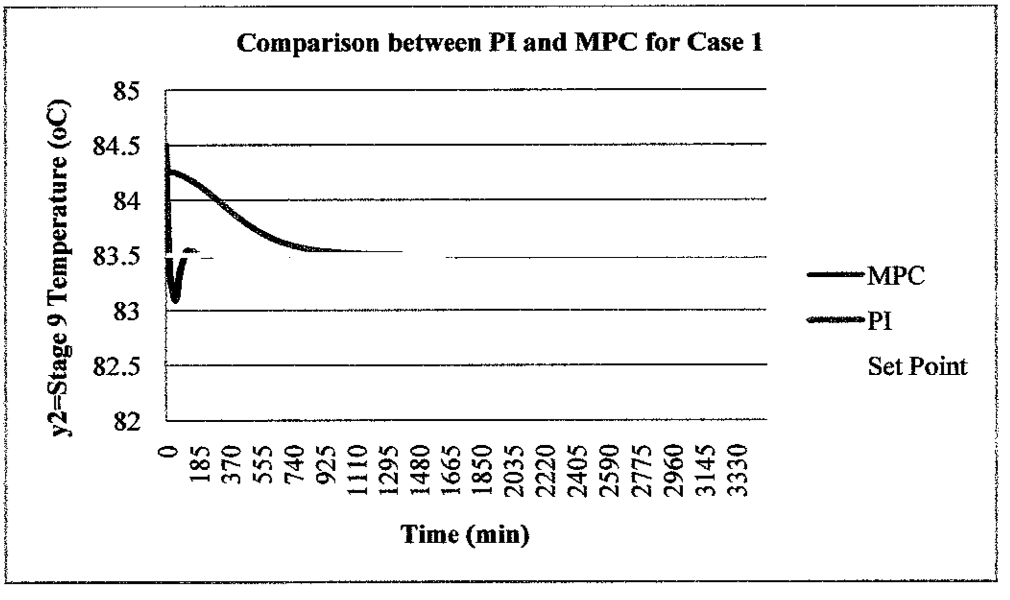

Figure 4.14: The comparison of performance between PI and MPC for case 1

From figure 4.12, 4.13 and 4.14 shows that both PI and MPC can be move to the new set point. But PI controller is move to the new set point faster than MPC controller. This is not means that PI is better than MPC. The controller which reach the set point faster means that it is higher in energy use due to the process have to be force in other to maintain in the set point.

For case 2, the set point of i-butane overhead product composition is changed from 1.962 % to 1.862% while maintain the set point of stage ninth temperature which is 84.5 °C for MPC controller. Since PI controller cannot control the composition, PI controller is control in top stage temperature change from 30 to 29 °C instead of i-butane overhead product composition with actually related each other as mention before. And the other variable is maintained.

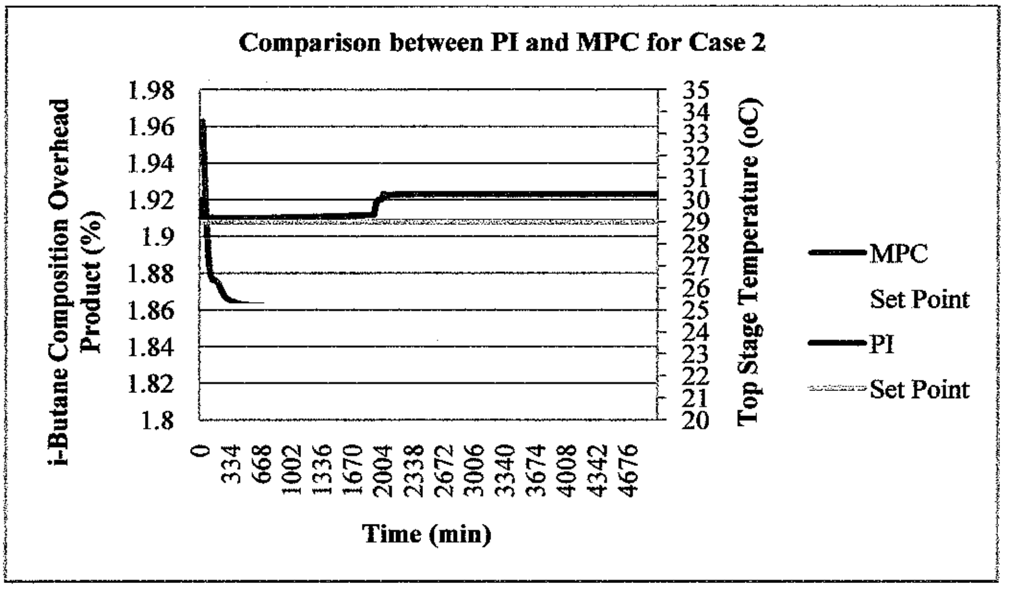

Figure 4.15: Comparison the performance of PI and MPC controller for y1 in case 2.

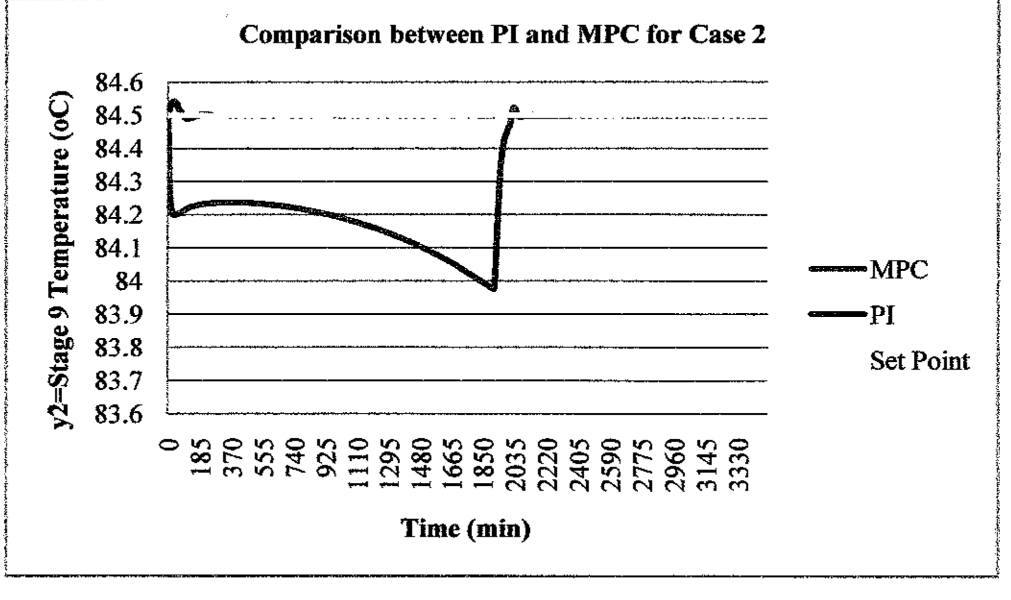

Figure 4.16: Comparison the performance of PI and MPC controller for y2 in case 2.

From figure 4.15 shows that PI controller cannot move the top stage temperature to the new set point 29 °C. But the MPC controller can move to the new set point of i-butane overhead product composition which is 1.862. From figure 4.16 shows that both PI and MPC controllers can be maintain the set point of 84.5 °C. But PI controller is reached the set point faster than MPC controller.

4.8.2 Energy Consumption

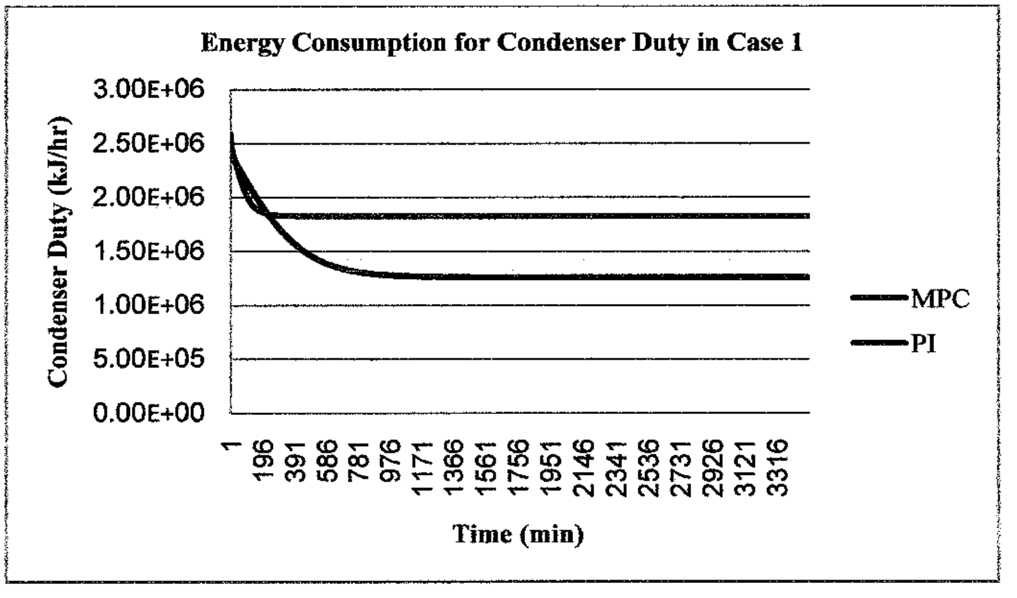

Figure 4.17: Energy consumption of condenser duty for case 1

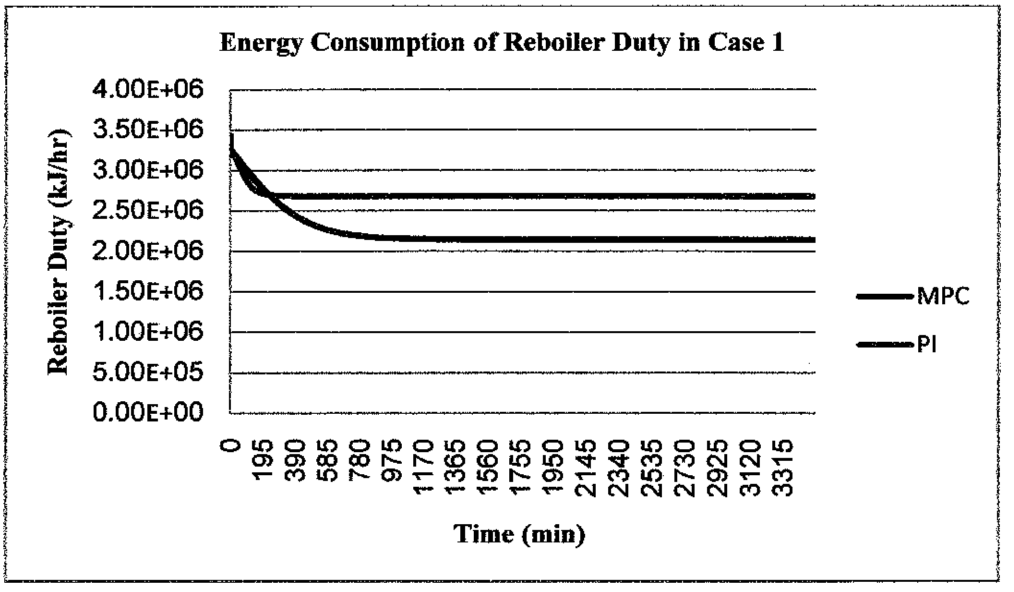

Figure 4.18: Energy consumption of reboiler duty for case 1

Figure 4.17 shows the energy consumption of condenser duty of both PI and MPC controllers for case 1. Figure 4.18 shows the energy consumption of reboiler duty of both PI and MPC controllers for case 1. It can be seen that both figure shows MPC controllers have smaller energy consumption compare to PI controller. So, it can conclude that MPC is better than PI controllers.

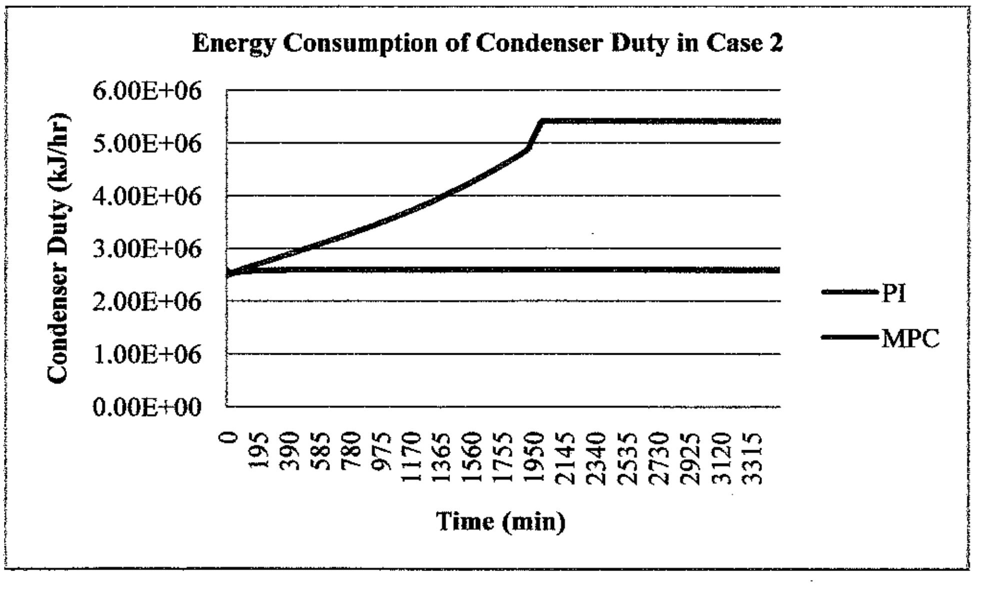

Figure 4.19: Energy consumption of condenser duty for case 2

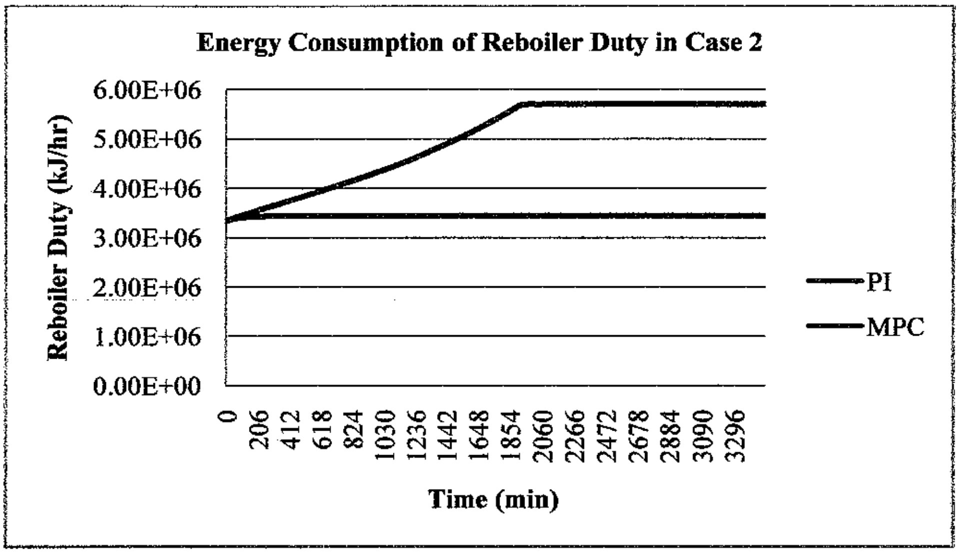

Figure 4.20: Energy consumption of reboiler duty for case 2

As well as case 2, figure 4.19 and 4.20 show the energy consumption of condenser and reboiler duty of both PI and MPC controller respectively. The graph is also shows that MPC controller has smaller energy consumption compared to PI controller.

CHAPTER 5
CONCLUSION

Advanced process control which is Model Predictive Control is used to control the plant. MPC controller has achieved better performance of product quality and reduces energy consumption compare to PI controller. By reducing the energy consumption will also result in the reducing amount of CO_2 released to the atmosphere which is the main cause of global warming.

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