Modeling and Predictive Control of Carbon Dioxide Removal Unit by Aqueous Alkanolamine

by

Noree Stirak

Dissertation submitted in partial fulfilment of

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CERTIFICATION OF APPROVAL

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Approved by,

____

(Dr. Nooryusmiza Yusoff)

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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.

NOREE STIRAK

ABSTRACT

The objectives of this project are to develop modeling and predictive control for Carbon dioxide removal unit by aqueous alkanolamine, compare its performance with an existing PI controller and to reduce carbon dioxide emission and energy consumption from a gas processing plant. Carbon dioxide removal unit is a plant to remove and eliminate carbon dioxide by aqueous alkanolamine. It absorbs impurities of natural gas; carbon dioxide, mercaptant and hydrogen sulfide. Modeling and Predictive Control is an advance technology which can be used to control and implement in process and overcome the problem. By reducing carbon dioxide and energy consumption, it will also result in the reducing amount of carbon dioxide released to the atmosphere which is the main cause of global warming, corrosion of equipment, pipeline and reduce the heating value of the process.

There are 4 methods to complete the project which are step testing, system identification, MPC installation and lastly compare Modeling and predictive control with existing PI control. The performance of MPC and PI are compared by using disturbance rejection method which is shows the performance to achieve and maintain the set point of percentage mole fraction of CO_2 and main stage temperature at tray no.17.

Modeling and Predictive control is a better performance than PI control according to its performance to achieve and maintain at a set point. Consequently, develop Modeling and Predictive Control in amine adsorption technology helps the process to reduce carbon dioxide emissions of natural gas manufacturer and minimize energy use of reboiler duty.

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NOMENCLATURE

Acronyms and Abbreviations

AI1 AI2	Analyzer
APC	Advance Predictive Control
C_3	Propane and propene
C_4	Butane, butene
CV	Controlled Variable
DV	Disturbance Variable
E	Heat exchanger
F1	Feed flow rate
F2	Reflux flow rate to the top of column
F3	Flow rate of hot oil
LPG	Liquefied petroleum gas
MPC	Modeling and Predictive Control
MV	Manipulated Variable
PI	Proportional Integral
PID	Proportional–integral–derivative controller
T-01	Distillation column
T1	Temperature and first stage of column
T2	Temperature of hot oil
V	Designates a process vessel

Chemical Symbols and Formulae

CO_2	Carbon dioxide
DEA	Diethanol amine
DGA	Diglycol amine
DIPA	Di-isopropanol amine
H_2O	Water
H_2S	Hydrogen sulfide
MEA	Monoethanol amine
MDEA	Methyl-diethanol amine

Variable

$A\ B\ C\ D$	Dynamic state space system matrices of the process
$A\,z^{-1}\ Bz^{-1}$	Polynomial in the backshift operator, z^{-1}
$A_{i,j}$ (s) $B_{i,j}$ (s)	Polynomial
$b_{i;j;k}$ $a_{i;j;k}$	The model coefficients,
$Cz^{-1}Dz^{-1}$	Polynomial in the backshift operator, z^{-1}
$e\ (t)$	White noise (innovation) sequences
f_i	The sampled output
f_j	The sampled output
g_i	The sampled output
I_{ny}	The identity matrix with dimension *ny*
j	Number of block columns in the block-Hankel matrices
N	The open loop stabilizing time of the system
nb na	The orders of polynomials B_{ij} and A*ij*
N_u	Control horizontal
N_s	The settle time of the process
Q_j	Weight matrix
R_j	Weight matrix
r_t	System set point at sampling instant t
S1 S2	The step response coefficients.
(u_i, y_j)	Input and output
u_t	System input(s) at sampling instant *t*
v_t	Input measurement noise
x_t	System states(s)at sampling inst*ant t*

$\hat{y}$	Predicted out put
y_t	System output(s) at sampling instant *t*
$\theta_{i,j}$	the process time delay and

CHAPTER 1
INTRODUCTION

1.1 Background

As natural gas becomes one of the high demand energy sources, the companies or manufacturers realize that natural gas needs to be commercializing in a high quality. The manufacturers look forward the technology to remove and separate of contaminants in natural gas. Natural gas consisted with a mixture of different gases and the main ingredient is methane, ethane, propane, butane and other hydrocarbon. Natural gas usually contains some impurity of carbon dioxide (CO_2), hydrogen sulfide (H_2S) and heavy hydrocarbon such as mercaptans and water vapor (H_2O).

Nowadays, natural gas that contains some impurity of carbon dioxide (CO_2) needs to be removed. According to the combination of carbon dioxide with water can make highly corrosive and rapidly destroy in pipeline and equipment inside plant. This problem of carbon dioxide within natural gas also can reduce heating value of natural gas stream and waste pipeline capacity.

Many technologies have been developed to enhance the process of removing carbon dioxide from the gas stream. There are many processes for gas sweetening such as batch solid bed absorption process, reactive solvent (Monoehanolamine (MEA), Diehanolamine (DEA) and Methyldiethanolamine (MDEA) processes, physical solvent to remove CO_2 and Membrane process to remove CO_2 out from natural gas. The most attracts many manufacturers to sweetening the natural gas is reactive solvent process of aqueous Monoethanolamine (MEA) and Methyldiethanolamine (MDEA) according to its high reactivity, low solvent cost, high equilibrium loading capacity and low heat or requires lower energy (Mandala et al, 2001).

1.2 Problem Statement

A global warming or climate change condition is major problems and this is the effect of rising temperature of the world and environment. The main reason of this problem comes from the greenhouse gas. Greenhouse gas is a sort of carbon dioxide or methane gas that retains heat and not reflects back to atmosphere. The increase of carbon dioxide to atmosphere from many sources such as the car industrial, fossil fuels (coal, oil, natural gas and hydrocarbon compound). Even though CO_2 is only 10–40 percent from the total post-combustion outlet into atmosphere, it still contributes to undesired global warming (Ahmad et al, 2010). According to above problem many researchers become aware of carbon dioxide (CO_2) emission problem.

Carbon dioxide (CO_2) that produce in natural gas manufacturer not only cause the global warming or climate change condition but it still cause the corrosion problem inside pipeline and equipment of the process plant and reduce the heating value of the gas. Therefore Carbon dioxide (CO_2) needs to be removed to eliminate or reduce the disadvantages in daily operations.

Currently there is several methods use for removing carbon dioxide from natural gas, flue gas or synthesis gas. Practically in industry use amine adsorption as an absorbent for removing carbon dioxide which is categorized under acidic gas group and MPC is a tool to implement in the process and overcome the problem that has been stated.

1.3 Objectives and scope of study

According to above problem statement, carbon dioxide is a major of environmental problem and it makes corrosion problem of equipment in the plant. Hence the objectives of this research project are:

- To develop Modeling and Predictive Control in amine adsorption technology to reduce carbon dioxide emissions of natural gas manufacturer.
- To reduce the energy consumption of plant.

The scopes of study for this research process

As consider for some constraints and time limitation, this project has planned to start off with narrow scope of Modeling Predictive Control for gas separation process plant of CO_2 removal unit in order to reduce carbon dioxide emission, reduce energy consumption and increase the heating value of the gas. The process should be taken in this project are as follow:

- Plant model development

  In this part of plant model development the process will develop using Aspen HYSYS and MATLAB's simulink; MATLAB simulink can use in constrained MPC.
- APC design and implementation
- Comparison with base layer control

  Lastly after finished APC design of plant or equipment, compare the APC design with the existing Base layer control; PI control.

The remaining discussions of this report are as follows:

This chapter explained about the background of the project which concern on the current situation of the global warming. Following with the next chapter on literature review part the main research is about the gas sweetening process to reduce and eliminate carbon dioxide. In this literature review explains why natural gas need to treat before enter into the process and what is the effect if the sweet gas; carbon dioxide is enter into the process. And another main in this literature review explain about the basic concept of Modeling and Predictive control and its application. Next is methodology part, this chapter includes with the method calculation of Modeling and Predictive control, and the Gantt chart of the project. Lastly is work progress; update on the current work and discussion on the step to accomplish the project.

CHAPTER 2

LITERATURE REVIEW / THEORY

In this chapter discusses on sweet gas from offshore and its disadvantage to environment and process plant. In order to sweeten gas the process of gas sweetening is described. The two processes of gas sweetening are presented; which are chemical absorption process and physical absorption process. Next is the example of carbon dioxide removal unit by aqueous alkanolamine. Lastly, introduce of Modeling Predictive and Control and its application.

2.1 Sweet gas

Natural gas from offshore is usually contains some impurities such as carbon dioxide (CO_2) hydrogen sulfide (H_2S) water vapor (H_2O) and the heavy hydrocarbons such as mercaptant. The main component contains in natural gas are methane, ethane, propane and a few of heavy hydrocarbon such as butane and pentane. The demand of consumption of natural gas is projected to increase from 95 trillion cubic feet in 2003 to 182 trillion cubic feet in 2030 (Xiao et al., 2009). The increase number of consumption of natural gas is the major problem of global warming.

Normally carbon dioxide is an impurity of natural gas from the offshore. This known as “sweet gas” it is usually desired to remove and eliminate carbon dioxide to prevent it from the corrosion problem inside the pipe and equipment of the process and to increase the heating value of the gas. Carbon dioxide emission from the natural gas manufacturer is the main causes that make the rising temperature of the earth and confront the climate change. The effect of carbon dioxide to the environmental and equipment plant is described by topic below.

The effect of sweet gas emission

The following below are the effect of carbon dioxide emission through the environment and in the natural gas processing.

- Global warming

Global warming phenomenon or climate change is one of the carbon dioxide (CO_2) emissions problems from the burning of fuel transportation and industrial production through environment. It causes and effect through the temperature rise of the earth. According to figure below is the correlation of increase in carbon dioxide through year and the temperature rise.

The figure below showed the increase of temperature from year 1950 to 2005 because of increase of carbon dioxide in the earth. The increase of carbon dioxide produce an effect of globally temperature (Global warming) and has an effect on climate change, rising of sea level, lead to change of rainfall and impact on plant animal and human.

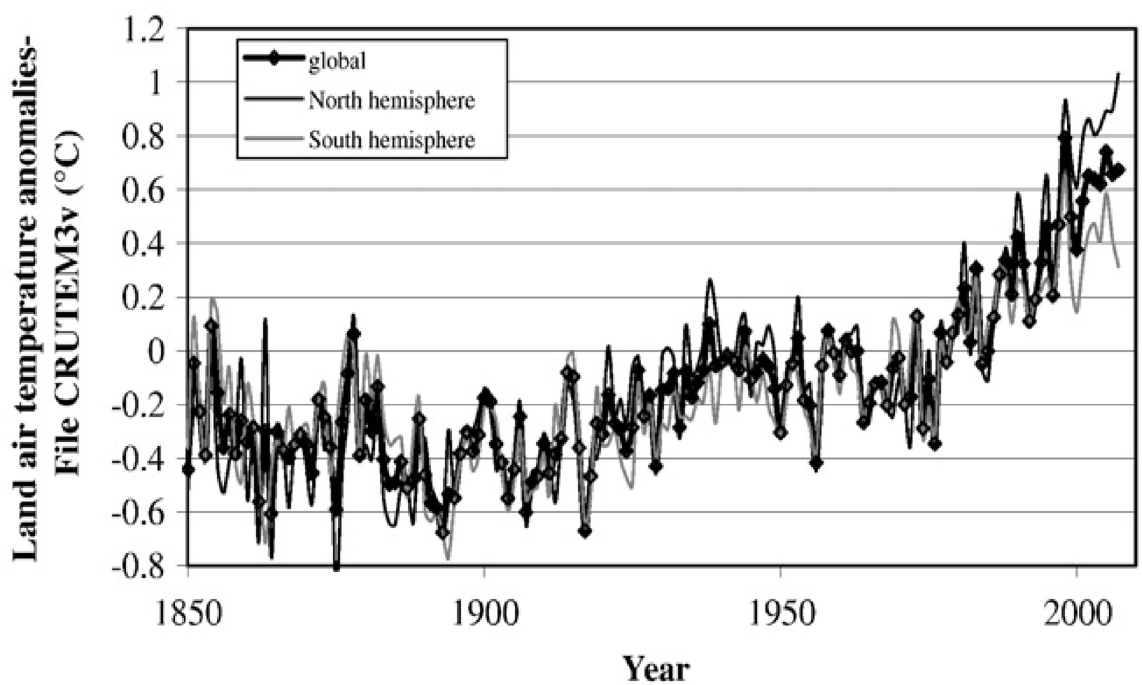

Figure 2.1: The correlation of increase of Carbon dioxide through year and temperature rise (source: Florides et al., 2009)

From the figure above, the land air temperature of global, north hemisphere and south hemisphere are increase through year 1850 to 2005 around 1.1 ^{0}C, it means that if carbon dioxide is increasingly , one day ice from the northern hemisphere and southern hemisphere may dissolve.

In addition, the effect of global warming is effecting of El Nino and La Nino phenomenon.

- Corrosion problem

The presence of carbon dioxide, hydrogen sulfide and free water can cause severe corrosion problems in oil and gas pipelines. Internal corrosion in wells and pipelines is influenced by temperature, CO_2 and H_2S content, water chemistry, flow velocity, oil or water wetting and composition and surface condition of the steel. A small change in one of these parameters can change the corrosion rate considerably, due to changes in the properties of the thin layer of corrosion products that accumulates on the steel surface. **(**Mora and Turgoose, 2002) mention that **"**The corrosion of carbon dioxide has many variable associated such as PH, temperature, pressure, flow steel composition, inhibitor, brine chemical composition on, surface films, etc."

- Reduce heating value of the process

Carbon dioxide fraction reduce the heating value of the gas, this is measured by the calorific value of the gas. As the CO_2 is a non combustible component in the natural gas, carbon dioxide will reduce the heating value of the gas. By this way if the gas contains high carbon dioxide content it is not economic to transport this gas through the pipe line. Therefore, carbon dioxide has to be removed.

The removal process of carbon dioxide is known as gas sweetening process to further understand the process, it is elaborated next.

2.2 Gas sweetening Process

According to above effect of Carbon dioxide; corrosion of equipment plant and environmental problem (Global warming). Gas sweetening is the one of the most important step of process to reduce and eliminate carbon dioxide in natural gas. There are many processes to purify the gas and remove carbon dioxide such as cryogenic process, adsorption process (pressure swing adsorption, PSA and thermal swing adsorption) hybrid solution and also membranes technology. The most desirable of sweetening process for natural gas manufacturer is absorption into aqueous blend by methyldiethanolamine (MDEA) because of, it can use to remove even in large amount of carbon dioxide. The gas sweetening process of aqueous solution of alkanolamines (Chemical absorption process) and physical absorption further described below.

2.2.1 Chemical absorption

Chemical absorption process is a carbon dioxide removal process or gas sweetening process by absorption of carbon dioxide in a solvent. The chemical absorption can be classified in to three main categories which are the hot potassium carbonate process, alkanolamines process and other chemical compound absorption process. (Refer to figure 2.4)

The most widely used for sweetening of natural gas are aqueous solutions of alkanolamine or alkanolamines process. It is usually used to remove a large amount of carbon dioxide and Hydrogen sulfide. MDEA or methyl-diethanolamine is a chemical compound used for gas sweetening. It is a tertiary amine, less basic and can be used in significantly higher concentration. According to (Abedini et al., 2010) MDEA is high solution concentration up to (50 to 55 wt%), high acid gas loading, low corrosion, slow degradation rates, lower heat of reaction and low vapor pressure and solution losses.

2.2.2 Physical absorption

Physical absorption is a process to absorb carbon dioxide and hydrogen sulfide at low temperature and high pressure. There are 4 organics liquid (solvents) such as Selexon, Purizon, Sulfurol and flour solvent. Physical solvent is more favor over chemical solvent when high concentration of acid gas (H_2S and CO_2). At normal pressure the compression of the gas for physical absorption is expensive. The physical absorption will be the better choice process to remove acid gas if the gas is available at high pressure (Barry, 2008) According to above organic liquid of physical absorption, the Purisol solvent is the most selective because of it has the highest capacity for absorption of acid gas (H2S and CO_2).

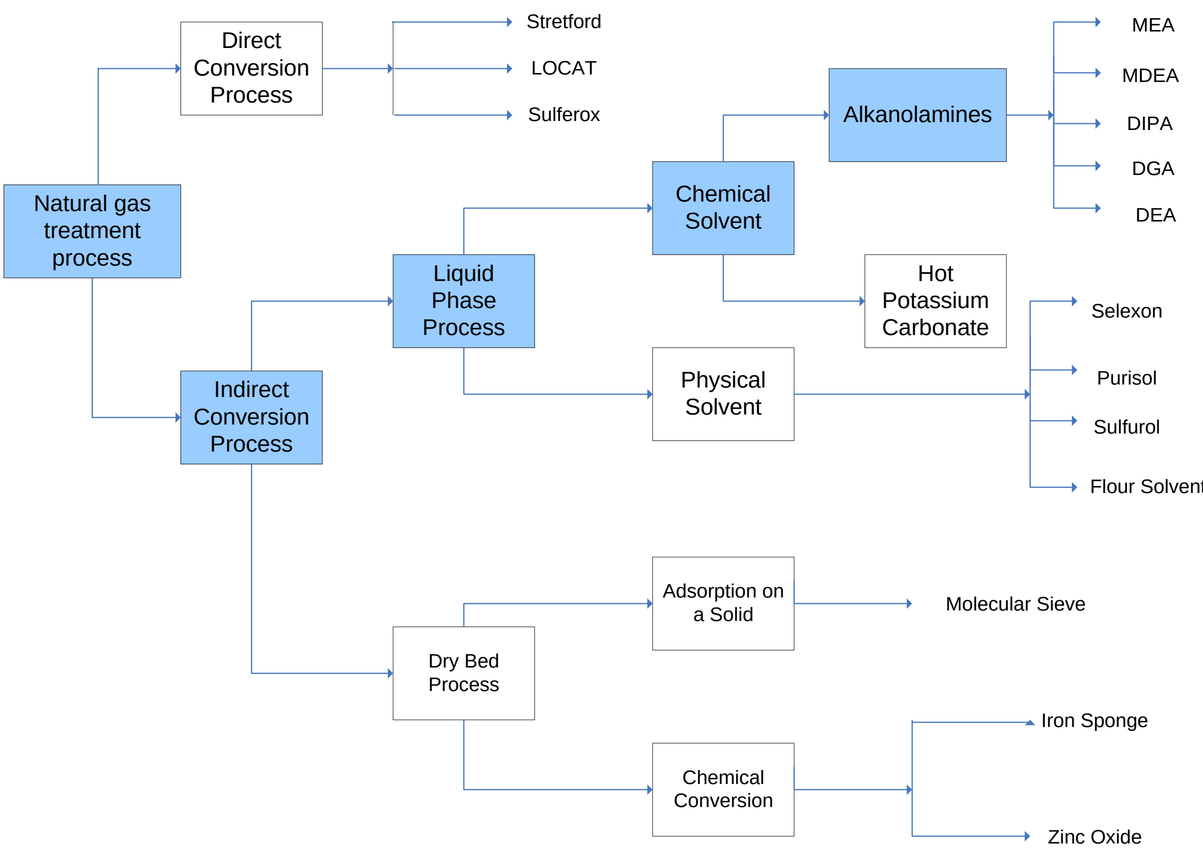

Figure 2.2: Alternative for natural gas sweetening (Source: Tennyson)

2.3 Example of gas sweetening process

Carbon dioxide removal unit or gas sweetening is a process to remove carbon dioxide, base on figure 2.3 there are six main equipments which are Absorber, Rich Pump, Rich/Lean Heat Exchanger, Lean Pump, Stripper and Lean cooler. The aqueous solution of MEA (Monoehanolamine) is used to remove carbon dioxide. The process of sweetening gas or carbon dioxide removal unit will further described as:

The sour gas from the power plant is fed through the bottom of absorber while lean MEA amine is fed into the top of the column and flow counter current of the feed gas then carbon dioxide are absorbed with lean MEA. The tray column absorber is used to provide intimate contact between gas and amine solvent (MEA) so carbon dioxide can transfer from gas phase to the solvent liquid phase. The chemical reaction takes place as:

$$CO_2 + NRH_2 \longrightarrow RH_2^{+}NCOO^{-}$$

$$RH_2{+}NCOO^{-} + NRH_2 \longrightarrow RH_2NCOO\text{-}NRH_2^{+}$$

The treated gas leave the top of absorber while the gas outlet from the bottom of absorber as rich gas. It is pumped by rich pump to transfer rich MEA to lean/ rich heat exchanger. The rich MEA from the bottom of absorber is heated before entering the top below the wash tray of the striper. At the striping section the rich MEA being regenerated via heating (Steam stripping). The acid gas is stripped and exists at the top of the stripper column, while the lean MEA is recycle back to booster pump and then exchange heat with the rich MEA solution at lean/ rich heat exchanger. The lean MEA needs to be cooled by lean cooler before enter through the top of absorber.

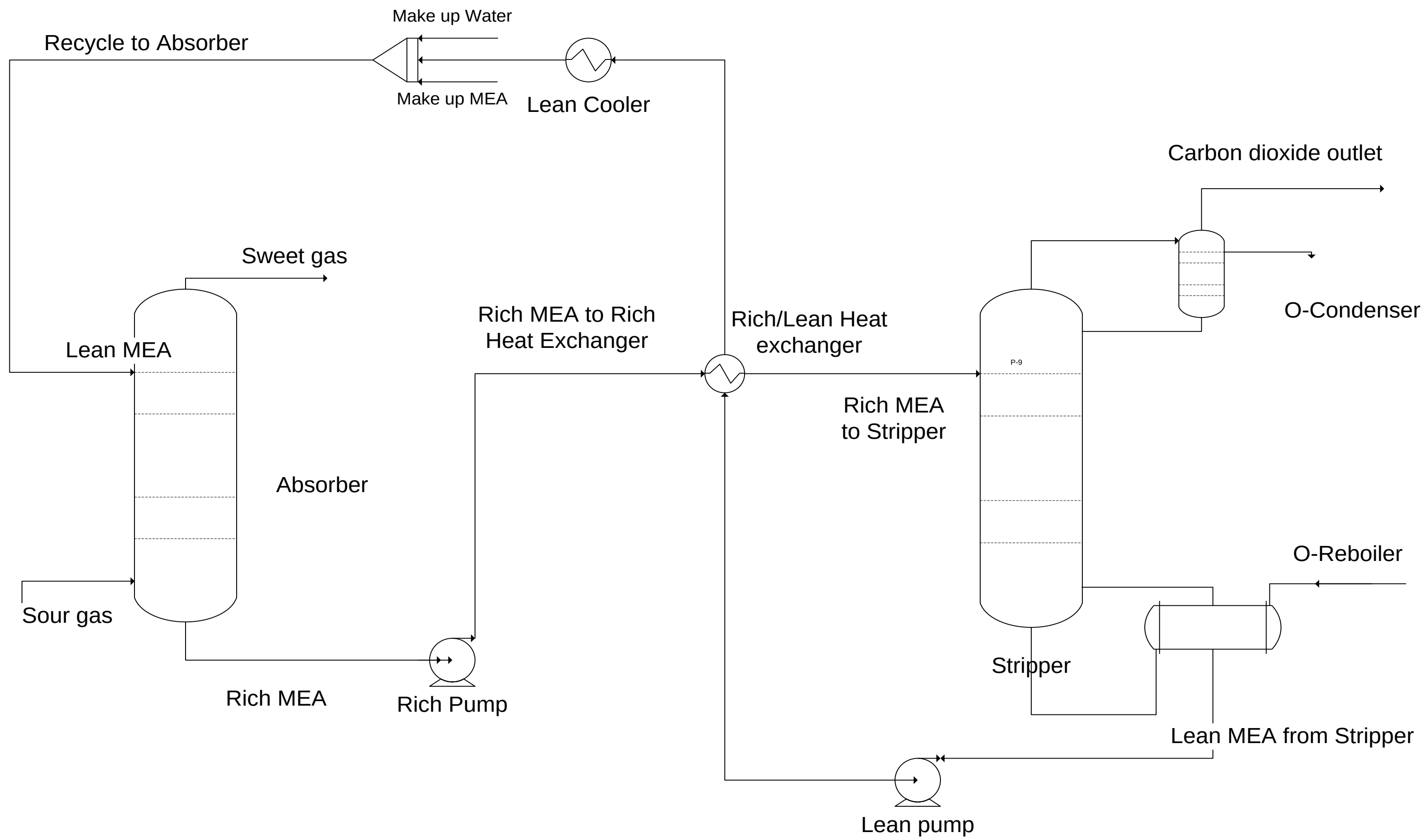

Figure 2.3: Process flow diagram of Carbon dioxide removal (Source: SIMS2007 Conference)

2.4 Modeling and Predictive control

Modeling and predictive control is a computer control that utilizes the process model and it is a tool to implement in this process and overcome the problem that has been stated. It is a popular subject for academic and industrial research it utilizes the process model for two central tasks, first explicit prediction of future process behavior, and second computation of appropriate corrective control action required to drive the predictive output as close to desired target value.

2.4.1 Basic concept of modeling and predictive control

Modeling Predictive Control is an advance technology which can be use to control a great variety of process with simple dynamics to more complex. A process model is used to predict the current value of the output variable.

Camacho and Bordons (2004) point out that, in order to apply this strategy, the basic structure of MPC shown below in Figure 2.4. A model is used to forecast the future output. It calculated by optimizer taking into account the cost function (where future error has to be considered) and also the constraints.

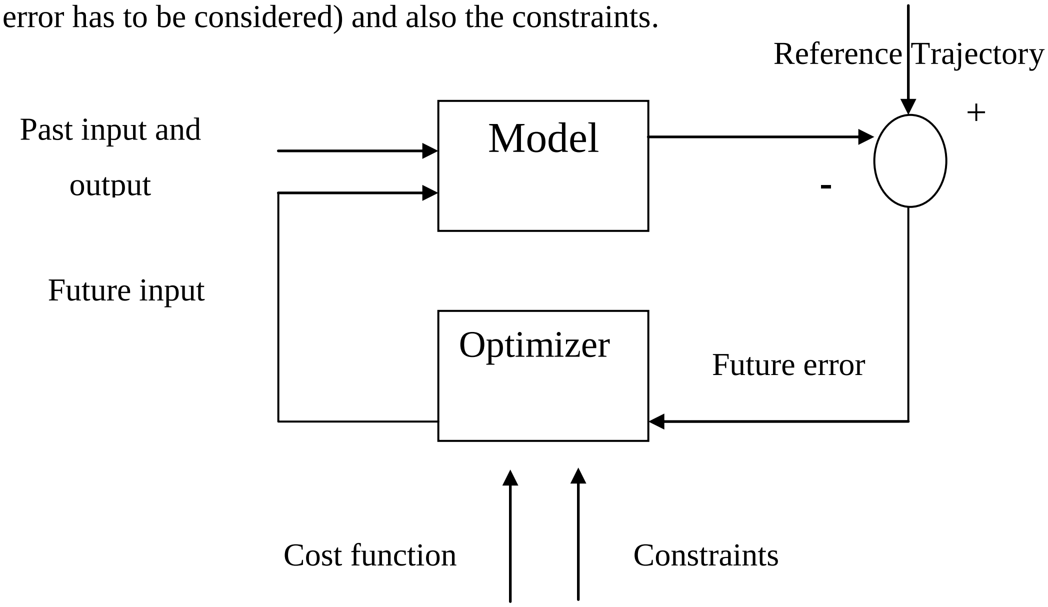

Figure 2.4: Basic structure of MPC

The process model runs, in consequence, a decisive role in the controller. The chosen model must be able to capture the process dynamics to precisely predict the future outputs and be simple to implement and understand. As MPC is not a unique technique but rather a set of different methodologies, there are many types of models used in various formulations.

2.5 Modeling Predictive Control Elements

Camacho (1999) argued that the MPC algorithms possess common elements and different option can be chosen for each of these elements as

- Prediction model
- Objective function
- Obtaining the control law

2.5.1 Prediction model

The used and propose of prediction model is to predict the process output at future time. According to (Huang and Kadali, 2008) MPC generally consisted of two parts which are process model and disturbance model. Both part of MPC technology need for the prediction and forecasting.

Process model

The most commonly used of MPC formulation appears in a given below:

Impulse response model

$$y_t = \sum_{i=1}^{\infty} f_i u_{t-i}$$

Where f_i is the sampled output, when the process is excited by an impulse response or unit response. This sum is truncated and N_s value are considered, N_s is the settle time of the process starting from instant N_s+1, impulse responses are approximately zero thus

$$y_t = f_i u_{t-i} = (f_1 z^{-1} + f_1 z^{-2} + \cdots . + f_1 z^{-Ns}) u_t)$$

Step Response Model

$$y_t = \sum_{i=1}^{\infty} g_i \Delta u_{t-i}$$

Where g_i is the sampled output for the unit step input $\Delta u_t = u_t - u_{t-i}$. g_i Will be constant after settling time N_s for the stable system, as an impulse response coefficient can be considered as the difference between two consecutive step response coefficients, the following relationship hold:

$$f_i = g_i - g_{i-1}$$

$$g_i = \sum_{j=1}^{i} f_j$$

Transfer function model

$$y_t = \frac{B(z^{-1})}{A(z^{-1})} ut$$

Step Space Model

$$x_{t+1} = Ax_t + Bu_t$$

$$y_t = Cx_t + Du_t$$

Where x is the state and *A*, *B*, *C* and *D* are the matrices of the system, input and output respectively.

Time series model for the disturbance

$$v_t = \frac{C(z^{-1})}{\Delta D(z^{-1})} e_t$$

Where $\Delta = 1 - z^{-1}$, v_t is the disturbance, *D* and *C* are often chosen as 1 in practical MPC

Disturbance model

The disturbance model is as important as process model, the differences between the measured output and the output can be calculated by:

$$n(t) = \frac{C(z^{-1})e\,(t)}{D(z^{-1})}$$

The polynomial $D(z^{-1})$ explicitly include integrator$\Delta = 1 - z^{-1}$, $e\,(t)$ is the white noise of zero mean and polynomial C consider equal to one. This appropriate for random change and Brownian motion.

2.5.2 Objective function

The various MPC algorithms propose different cost function for obtaining the control law. The general aims are: (Huang and Kadali, 2008)

- The future output should follow a determined reference signal over the considered horizon.
- The control effort necessary for doing so should be considered in the objective function.

The general expression for such an objective function is

$$j = \sum_{j=N1}^{N2} [r_t + j - \hat{y}\,(t + j|t)]^T Q_j [r_t + j - \hat{y}\,(t + j|t)] + \sum_{j=1}^{N_u} [\Delta u_{t+j-1}]\,^T R_j\ [\Delta u_{t+j-1}]$$

Where Q_j and R_j are the weighting matrices.

2.5.3 Control Law

In order to obtain the control law $u(t + k|t)$ it need to minimize the function *J* from objective function and equate the derivative to zero. This is a least square problem. According to (Huang and Kadali (2008)) “if there are hard constraints on ut, $\Delta ut\ or\ \hat{y}(t + j|t)$ analytical solutions are not possible and numerical optimization is necessary."

When numerical optimization are needed, obtaining the solution is not trivial because there will be N_u decision variables in the optimization". The control horizon is used to impose a structure on the control law. Under this concept it is considered that after a certain time window N_u, ut become constant or equivalently $\Delta u_t = 0$.

$$\Delta u_{t+j-1} = 0 \qquad J > N_u$$

To sum up, the design of Modeling and predictive control involve the requirement of prediction model, objective function and optimization to get the control laws.

2.6 Example of MPC application

2.6.1 Process description of industrial of C_3/C_4 splitter

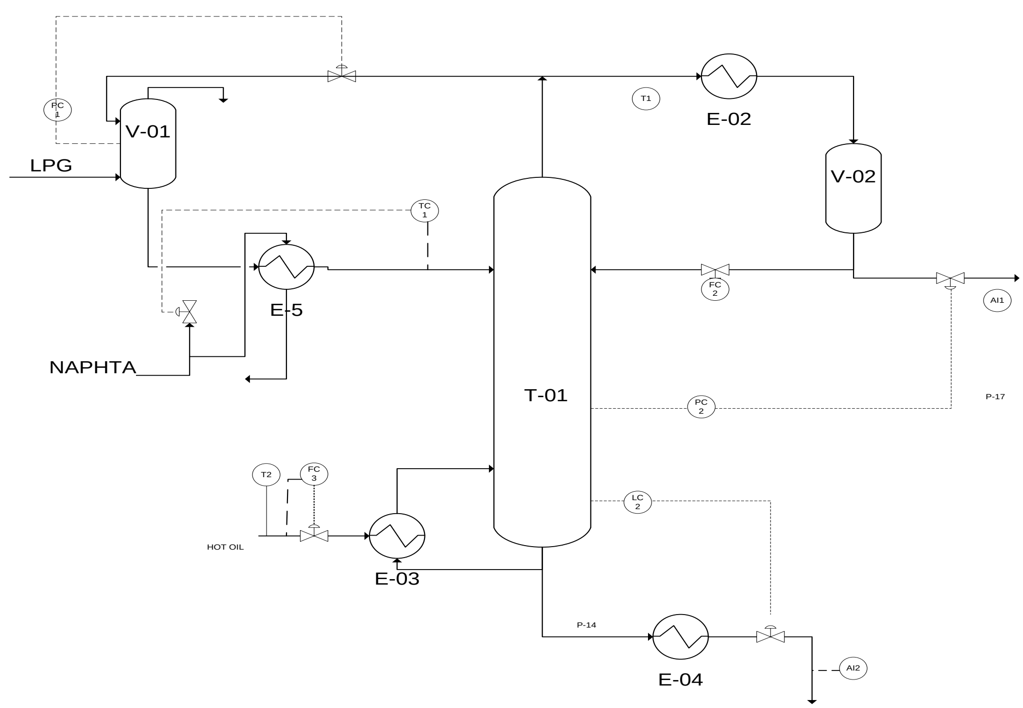

Figure 2.5 Schematic representation of the C3/C4 separation system.

This is an application of MPC that has been applied to industrial of C_3/C_4 splitter. (Porfirio et al, 2003)

From the above Figure 2.5 shown the schematically distillation system, C_3 stream (propane and propene) is separated from a C4 stream, which contains butane, butene and other hydrocarbons. PID controllers are represented for this system of C_3/C_4 separation. Liquefied petroleum gas (LPG) is fed from the top of a debutanizer column that separates the LPG from gasoline. From the figure above T-01 is represented as the distillation column, E stands for heat exchanger and V designates a process vessel. The C_3 stream is produced as the top stream of the distillation column and the C_4 stream is produced as the bottom stream of the column. AI1 and AI2 are analyzer. The variable of this process are shown below

Table 2.1: The variable of C3/C4 separation system

Controlled out put	T1 (temperature and first stage of column) percentage of propane and propene in Analyzer AI2
Manipulated input	F3 (flow rate of hot oil) F2 (reflux flow rate to the top of column)
Disturbance	F1 (feed flow rate) and T2 (Temperature of hot oil)

MPC with state-space model

This process C3/C4 separation systems is based on state-space modeling. It is more economical of number of state compared with impulse response model. Where the step-space model of the form are

$$x_{k+1} = A_{xk} + B\Delta u(k) \tag{1}$$

Where

$$x_{k+1} = \begin{bmatrix} y(k+1) \\ y\ (k+2) \\ \vdots \\ y(k+N) \end{bmatrix}, x_k = \begin{bmatrix} y(k+1) \\ y\ (k+2) \\ \vdots \\ y(k+N-1) \end{bmatrix}$$

$$\begin{bmatrix} 0 & I_{ny} & 0 & \dots & 0 \\ 0 & 0 & I_{ny} & \ddots & \vdots \\ \vdots & \vdots & \ddots & \ddots & 0 \\ & & & 0 & I_{ny} \\ 0 & \dots & & 0 & I_{ny} \end{bmatrix}, B = \begin{bmatrix} s_1 \\ s_2 \\ \vdots \\ s_3 \end{bmatrix} \tag{2}$$

Where N is the open loop stabilizing time of the system, I_{ny} is the identity matrix with dimension *ny;* which is the number of outputs and S1; S2 are the step response coefficients.

Assume the above pair (input and output) = (u_i, y_j) there is a Laplace transfer function model

$$G_{i,j}(s) = (\frac{B_{i,j}}{A_{i,j}(s)} e^{-s\theta i,j}), \tag{3}$$

Where $\theta_{i,j}$ is the process time delay and $B_{i,j}$ (s) and $A_{i,j}$ (s) are polynomial given by

$$G_{i,j}(s) = b_{i,j,0} + b_{i,j,1}s + b_{i,j,2}\ s^2 + \cdots b_{i,j,nb}\ s^{nb}$$
$$A_{i,j}(s) = 1 + a_{i,j,1}s + a_{i,j,2}\ s^2 + \cdots a_{i,j,na}\ s^{na}$$

Where $b_{i;j;k}$ and $a_{i;j;k}$ are the model coefficients, *nb* and *na* are the orders of polynomials B_{ij} and A*ij* ; respectively.

Assumed that none of the roots of A *ij*= 0. Therefore, the step response of the system represented by (3) can be written as follows:

$$S_{i,j}(n) = c_{i,j,0} + \sum_{g=1}^{na} c_{i,j,g} e^{r_{i,j,g}(nT-\theta i,j)}\ , nT\ > \theta_{i,j}\ , \tag{4}$$

This system, a state space model, which is equivalent to (1) and (2) but has a reduce number of state, given by

$$x_{k+1} = Mx_k + S\Delta u(k), \tag{5}$$

2.6.2 Dynamic Modeling to Minimize Energy Use for CO_2 Capture in Power Plant by Aqueous Monoethanolamine

This is an application of dynamic modeling to minimize energy use for CO_2 capture in power plant by aqueous monoethanolamine. The model was developed in aspen Custom Modeler program for stripping in CO_2 removal unit.

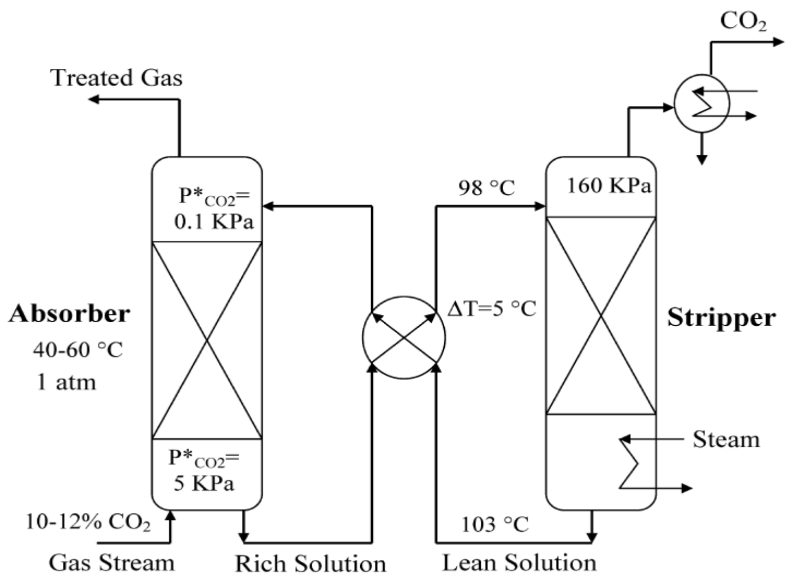

Figure 2.6: Typical absorption/stripping process for CO2 removal with monoethanolamine. (Sepideh and et al, 2009)

This is absorption and stripping process for CO_2 removal with monoethanolamine. In this process the Absorber column is operated at Temperature during 40-60 ^oC and 1 atmospheric pressure. The gas stream enters at the bottom of the column (Absorber) which contains 10-12% of CO_2. The lean amine is loaded at the top of the absorber column and it counter current contact with the CO_2 gas from the bottom of column. The CO_2 is absorbed by amine by physical and chemical absorption then come out of the absorber column as rich solution which contains with high concentration of CO_2 in the amine. The amine was pumped to the heat exchanger before entre in to stripper. The temperature of reboiler operating at 100-120 ^{o}C, it influent the removal of CO_2 from the amine solution at the top of stripper and amine will leave the stripper column as Lean amine. It was pumped and cooled by the cool rich solution before enter to the top of absorber.

Dynamic strategy For CO_2 Capture

The main objective of CO_2 capture is to reduce cost of energy during loading the electricity peak load and reduce the duty of reboiler stream. The simulation was done and dynamic are set, there are possible manipulated variable or input variable of this process is lean loading in absorber column, the overhead pressure of column and reboiler liquid level are controlled variable. According to this process there are 2 dynamic strategies which are

Strategy1: Reduce rich solvent flow rate from the absorber while the lean loading of amine at the top of the absorber column constant, refer process strategy1 at appendix A.

Strategy2: Increase loading strategy by regenerate all the rich solvent in the stripper, refer process strategy2 at appendix B.

During the step change and dynamics model of the process there are negative values of 10% step change for both strategies and the result of the dynamic of this process is shown below

Table 2.2: Detailed simulation Result

	CO2 removal		Lean loading		Preb KPa		Treb^{o}C		τL packing S
	Initial	final	Initial	Final	Initial	Final	Initial	final	average
Strategy 1	90%	81%	0.42	0.4199	162.76	162.36	103.23	103.19	4.98
Strategy 2	90%	80.3%	0.42	0.4315	162.76	162.27	103.23	101.93	5.1

According to step testing of this process the result and the relation of each variable is presented below:

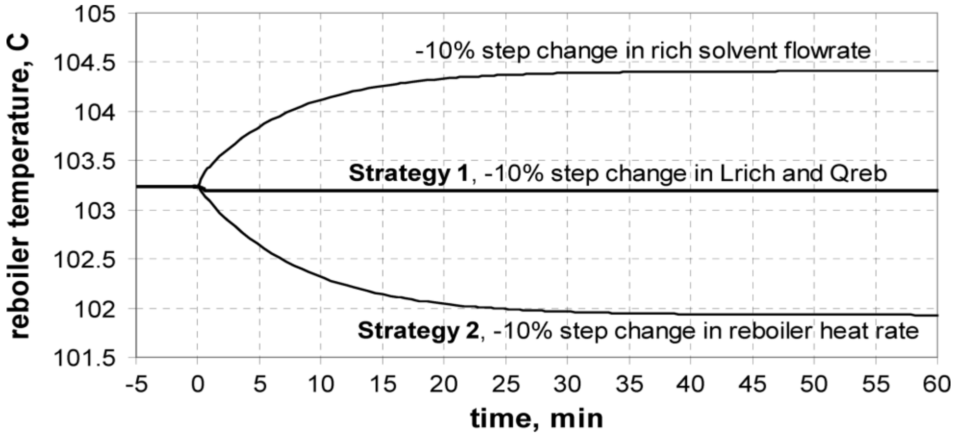

Figure2.7: The reboiler pressure responses to change of rich solvent flow rate and reboiler heat rate.

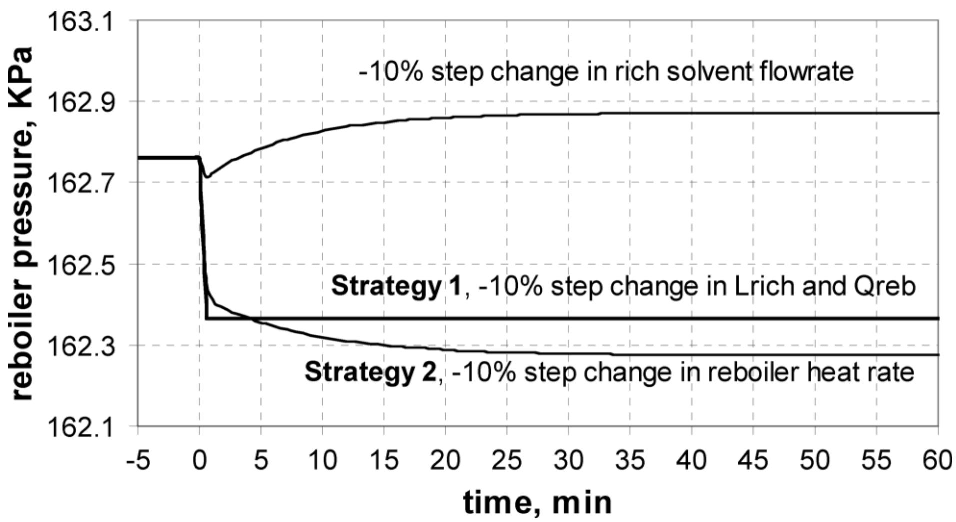

Figure2.8: Reboiler pressure responses to the change of rich solvent flow rate and reboiler heat rate.

According to figure above; the solvent flow rate decrease when the reboiler temperature increases. When reboiler pressure is decrease it not change of strategy 1 and it make strategy1 faster to reach the steady state while the strategy 2 is slow, the consequently both reboiler pressure and lean loading influence the temperature in the reboiler. (Sepideh et al, 2009)

2.6.3 Model Based Control of Absorption Tower for CO2 Capturing

This work of CO_2 capturing by aqueous MEA is concerned and considered. The model predictive control of CO_2 capturing at absorber is developed to reduce percentage release of CO_2 gas to the atmosphere.

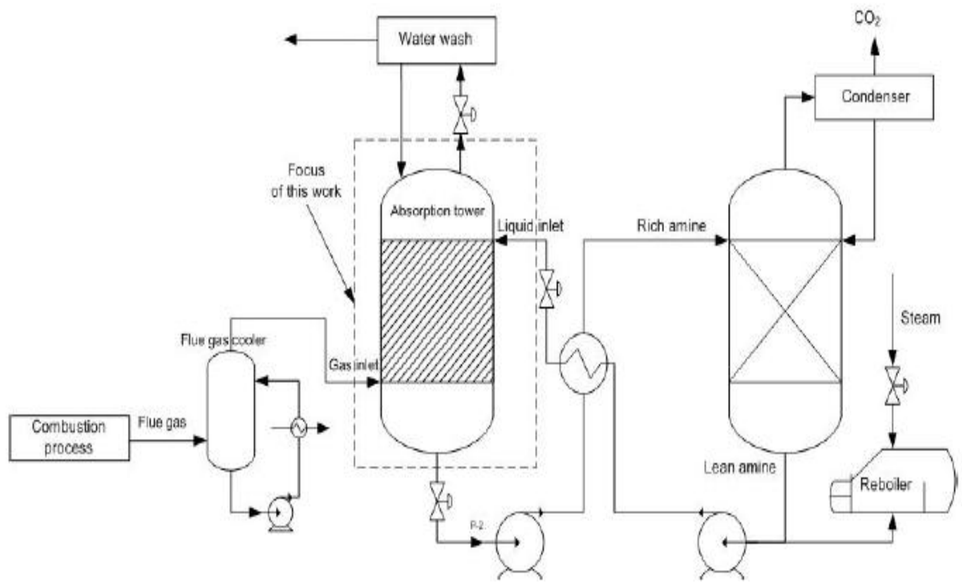

Figure2.9: Absorption/stripping process flow diagram (Bedelbayev et al., 2008)

The absorption and stripping process flow diagram shows above is a process to reduce and remove CO_2 removal by alkanolamine acid gas removal process. The fuel gas from the combustion process enters to the bottom of absorption tower and contact counter current with lean amine MEA (any alkanolamine) which coming from the top of the tower. During contact counter current of fuel gas and lean amine, chemical sorption and physical sorption occurs. The CO_2 diffuse in amine then become rich CO_2 (Rich gas) leave at bottom of absorption tower. Mean while the gas that have been absorbed from the amine will move to the top of the absorber tower. Before leave the absorber tower as sweet gas, water wash pass through the gas to wash and purified gas from alkanolamine acid solution. Rich gas pass through heat exchanger and enter to the top of the striper. The reboiler at stripper column will strip the CO_2 gas from amine and CO_2 leave at the top of the stripper. At the same time, lean amine leaves the stripper through the bottom of column and pump back as lean amine enter to the top of absorber tower.

MPC and Process control

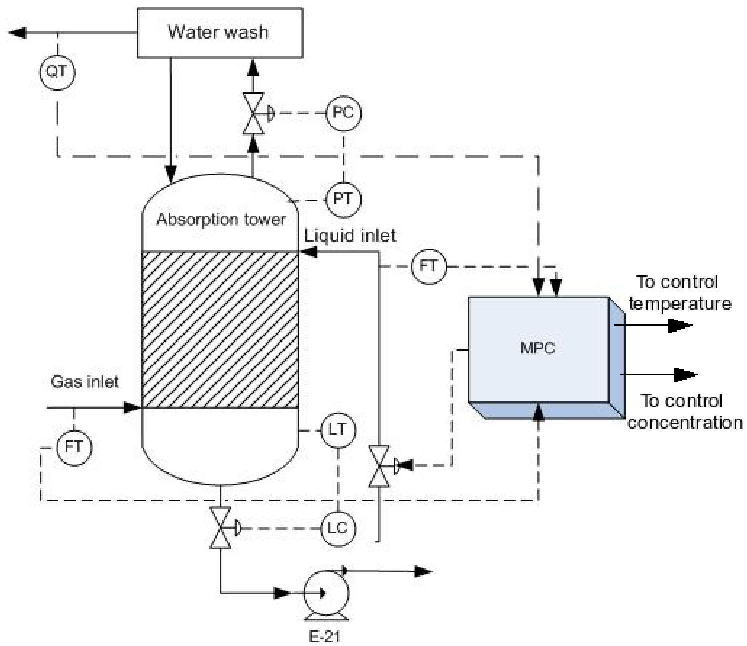

Figure2.10: MPC for the absorption tower control (Bedelbayev et al., 2008)

According to this process the output variable or controlled variable is the concentration of the CO_2 at the top of absorber tower. The input variable or manipulated variable for this process are the liquid velocity, liquid concentration of MEA and liquid temperature. (Bedelbayev et al., 2008) The main manipulated variable for this process is liquid velocity of absorber tower. The inlet of CO_2 gases, temperature of inlet gas and inlet velocity of gas are disturbance of the process. MPC is implemented for this process to improve the system by control the manipulated variable. The result of the simulation shows in good result of 91.01 % satisfactory is achieved of 15 height tower and 0.005 m/s of liquid velocity. This MPC (model predictive control) is used to improve the operation and it is calculated in MATLAB Tool box.

This chapter discussed on basic concept of natural gas, modeling and predictive control and its application. Next chapter will further discuss on methodology and project activities for the first and second semester.

CHAPTER 3

METHODOLOGY

The methodology of modeling and predictive control for carbon dioxide removal unit by aqueous alkanolamine, in this chapter discusses about the methodology/procedure, project activities, Gantt chart and equipment used.

3.1 Project activities

The Figure3.1 shows the step to complete project that consisted of three main steps together; which are Plant model development, APC design and Implementation and lastly Comparison MPC technology with BLC (Base layer control).

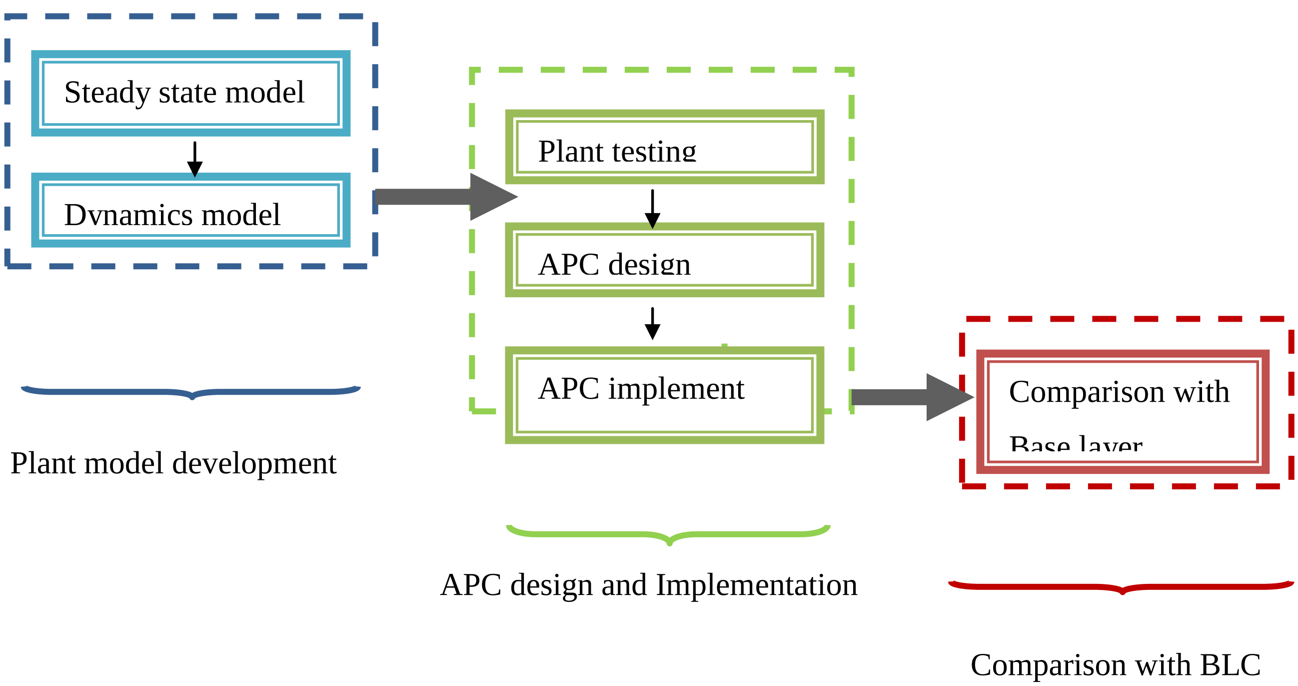

Figure 3.1: The flow project activities

There are three main activities to complete these project activities there are.

- Plant model development

Carbon dioxide removal unit is a process to remove CO_2 from natural gas. In this project plant used aqueous alkanolamine solvent which is a chemical absorption process to absorb CO_2 from the feed gas. The aqueous alkanolamine absorb the contaminant of CO_2. The treated gas leaves the top of the absorber while the rich gas out to the bottom of absorber. The rich gas is heated and regenerates via stripper and circulates back to the absorber as lean anime solution. The Manipulated Variables (MV) are flow rate at the top of absorber and temperature at stripper column. The Controlled Variables (CV) are the temperature in stripper and the composition of rich amine. The Disturbed Variables (DV) are the flow rate and temperature at the top of stripper and the flow rate of inlet gas.

The project will start plant model development which has to form the steady state model and dynamic model by using engineering software (Aspen HYSYS 2006 and Aspen HYSYS 3.2). The design of the process based on Figure 2.4 carbon dioxide removal units source by (SIMS, 2007 conference). In this design used amine properties package to simulate the design according to solution of aqueous alkanolamine. And use MATLAB simulink to simulate the process for constraint MPC.

- APC design and implementation

This is 2x2 MPC project. In this project's activity is divided into 3 steps together which are first the step testing, in step testing normally deal with changing of manipulated variable and observe the relationship of manipulated variable and control variable. APC design and implementation step can refer to topic 3.2 "Modeling and predictive control calculation" which consisted of 7 steps together and further elaborated in each step below as topic 3.2.

- Comparison with Base layer control

The last step of project activities is to compare MPC project with the existing of base layer control (PI control). In this step the MPC project was completed then compare with PI control.

3.2 Modeling and Predictive Control Calculation

Figure 3.2 (Qin and Badgwell, 2003) provides an overview of the flow for MPC calculation. Each step performed at each control execution time. It consisted of 7 steps as follow the figure below:

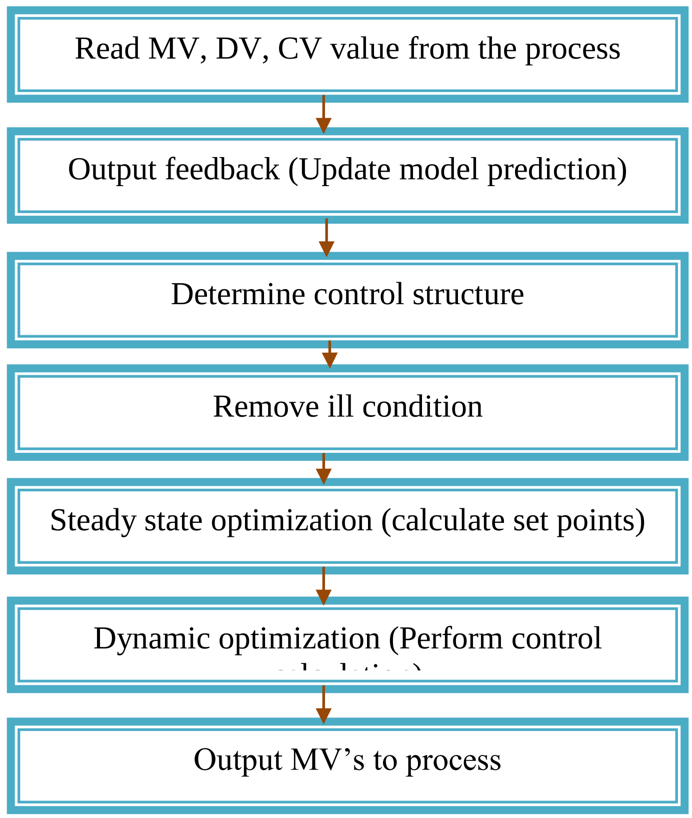

Figure 3.2: Flow of MPC calculation at each control executions

(Source: Qin and Badgwell 2003)

The MPC calculation is elaborated below:

3.2.1 Read MV, DV, CV value from process

The first step of MPC calculation, it is important to know, DV (Disturbance Variable), MV (Manipulated Variable) and CV (Controlled Variable). Furthermore; each measurement has its own sensor status to indicate whether is properly functioning or not. If the MV controller is disable or unavailable for control, it can be consider as disturbance variable DV.

3.2.2 Output feedback (state estimation)

This step will estimate the dynamic state of the system. (Qin and Badgwell, 2003) argued that the most of state estimation is not incorporate in industrial MPC products at all.

3.2.3 Determine the controlled structure

The controller need to determine which MV should be manipulated and which CVs should be controlled. If the operator has enabled control of the CV and the measurement status of CV is good, therefore it should be controlled. MV has to meet the same criteria also and the lower level control function must also be manipulated if the lower level control function is disabled, the MV cannot be use for control.

3.2.4 Removal of ill condition

Ill-condition occurs when the available inputs have very similar effects on two or more outputs. (Maciejowski, 2002; Qin and Badgweel, 2003) is very definite “If ill-condition is detected 3 effective strategies are available for remove”. First, if ill conditioning is detected, low-priority outputs are sequentially removed until ill condition is eliminated. A second approach is based on singular value analysis by exclude small singular values;

the process model can be adjusted. Lastly ill-condition can be removed by adjusting and MPC design parameter, the move suppression matrix R.

3.2.5 Dynamic optimization

The MPC controller must compute a set of MV adjustment that will drive the process to the desired steady-state operating point without violating constraints.

3.2.6 Dynamics optimization (Performed control calculation)

There are 3 basic types of MPC which are Hard, Soft and Setpoint approximation ***(Refer Appendix A: The three basic type of constraint)*** Hard constraints should not be violeted in the future, but Soft constraints (middle) may be violated in the future, but the violation is penalized in the objective function. Setpoint approximation of constraint (bottom) penalizes deviations above and below the constraint. Shades areas show violations penalized in the dynamic optimization.

3.2.7 Output and input trajectory

A setpoint, zone, reference trajectory or funnel is basic option to specify future CV. ***(Refer Appendix A : Four options for specifying future CV behavior)*** is an option to drive the CVs to a fixed setpoint, with deviations on both sides penalized in the objective function. This is particularly important when the internal model differs significantly from the process. Several of the MPC algorithms use move suppression factors for this purpose. One way to implement zone control is to define upper and lower soft constraints.

3.3 Gantt chart

Figure 3.3 and 3.4 show the process plan or Gantt chart for 2 semesters to complete the MPC project.

3.3.1 Gantt Chart for First Semester and Second semester

Table 3.1: Gantt chart for FYP 1

Gantt chart for FYP semester1					
	Months				
Activities	JAN	FEB	MAR	APRIL	MAY
1. Literature Review					
2. Model Development of CO2 removal unit.					
-Steady state model					
-Dynamic model					
3. Hysys Tutorial					
4. Plant Model simulation					
5. Report writing					10/5/2010

This project starts with literature review it took almost 4 months of study through the literature. According to the literature review the author has not yet found the topic of MPC related with carbon dioxide removal unit. However, other literature help the author understand the concept of MPC. During research through the journal and books the author has studies HYSYS tutorial and try to simulate the plant according to data and information from (SIMS, 2007). Two month of try and error to obtain the plant model simulation some part of the equipment not converges. However, the plant model simulation of this project will continue doing during June 2006 semester break.

Table 3.2: Gantt chart for FYP 2

Gantt chart for FYP semester2					
Activities	Months				
	JUL	AUG	SEP	OCT	NOV
1. Plant Testing.					
2. MPC Design.					
3. Simulation and MPC Implementation.					
4. Comparison with Base Layer Control.					
5. Report Writing.					

According to Gantt chart, the project activities have to be completed on time and during early second semester have to start the plant testing. The steady state simulation and dynamic simulation plant model have been done during semester break. And early of the 2nd semester July and August 2010, around 2 months start doing the plant testing, it take time to run because of some erroneous and un-converge of model. However, MPC design and MPC implementation target to finish on October 2010. After complete the project of MPC design and implementation lastly prepare and writing dissertation report and compare the project with the existing Base layer control for 2 months and 1 month respectively.

3.4 **Requirement tool:**

Two main tools of Aspen HYSYS version 2006 and MATLAB are required in this project.

- **Aspen HYSYS**

Aspen HYSYS has been used to simulate the process design and to get the steady state model of gas sweetening process and carbon dioxide removal unit. According to carbon dioxide removal unit by aqueous solution of alkanolamine the Amine Properties Package model have been simulated. Within Amine Properties Package models, Kent Eisenberg or Li- Mather are available (Lars, 2007).

The Aspen HYSYS simulate is used to gain steady state model, dynamic model, step testing of project and it is used to compare the performance of PI controller and MPC controller in MPC comparison method.

- **MATLAB**

MATLAB software has been use to calculate the dynamics model in constraint MPC.
It is used to find the FOPTD (First Order Plus Time Delay) model parameter which then be used for MPC installation.

This chapter discussed about the methodology for modeling of CO_2 removal and work plan which has been posted in Gantt chart for first semester and second semester. The next chapter will be chapter 4 which the result of the project is discussed and compared the project with other sources of information related.

CHAPTER 4

RESULT AND DISCUSSION

In this chapter explain the details of project and discuss the result, which consisted of the process description and modeling (steady state modeling and dynamic modeling) of the project, step testing, system identification and lastly discuss the process model of MPC with other related literature.

4.1 Process Description and Process Flow Diagram

CO_2 removal unit process consisted with many types of equipment. The target of this process is to remove the concentration of CO_2 or remove CO_2 out from sour gas to sweet gas. The sour gas enters to the FWKO separator Tank to remove or knock some water and heavy particle out of the gas. The liquid particle will drop to the bottom of FWKO separator tank and some of vapor particle goes to the top of the FWKO separator. The gas from the top of the FWKO separator enters to the DEA Contractor column at the bottom stage. In the DEA Contractor column consisted of 20 trays and at each tray, physical sorption and chemical sorption occurred. Lean amine is loaded at the top of the DEA Contractor in counter flow contact with gas from the bottom of the DEA Contractor. The Lean amine solvent absorbs CO_2 out of the gas. The treated gas leave the top of DEA Contractor as sweet gas while the outlet of DEA Contractor as rich amine (Rich DEA).

The Rich DEA enter to the Flash tank and drop down from the tank to the heat exchanger (L/R HEX) before enter into the stripper column (Regenerator).

The Regen Feed from heat exchanger (L/R HEX) enters to the stripping section (Regenerator) and it being regenerated via the heating from the reboiler near of the Regenerator. The heat strip CO_2 out of rich amine and exists at the top of Regenerator while lean amine accumulate at the bottom of Regenerator.

The lean amine leaves the Regenerator and pass through heat exchanger (L/R HEX) before mix with MAKEUP H_2O to purify the gas from amine solution. Then it needs to cool before pump and recycle back to the top of the absorber (DEA Contractor).

Figure4.1 below shows a simplified process flow diagram or process of the project.

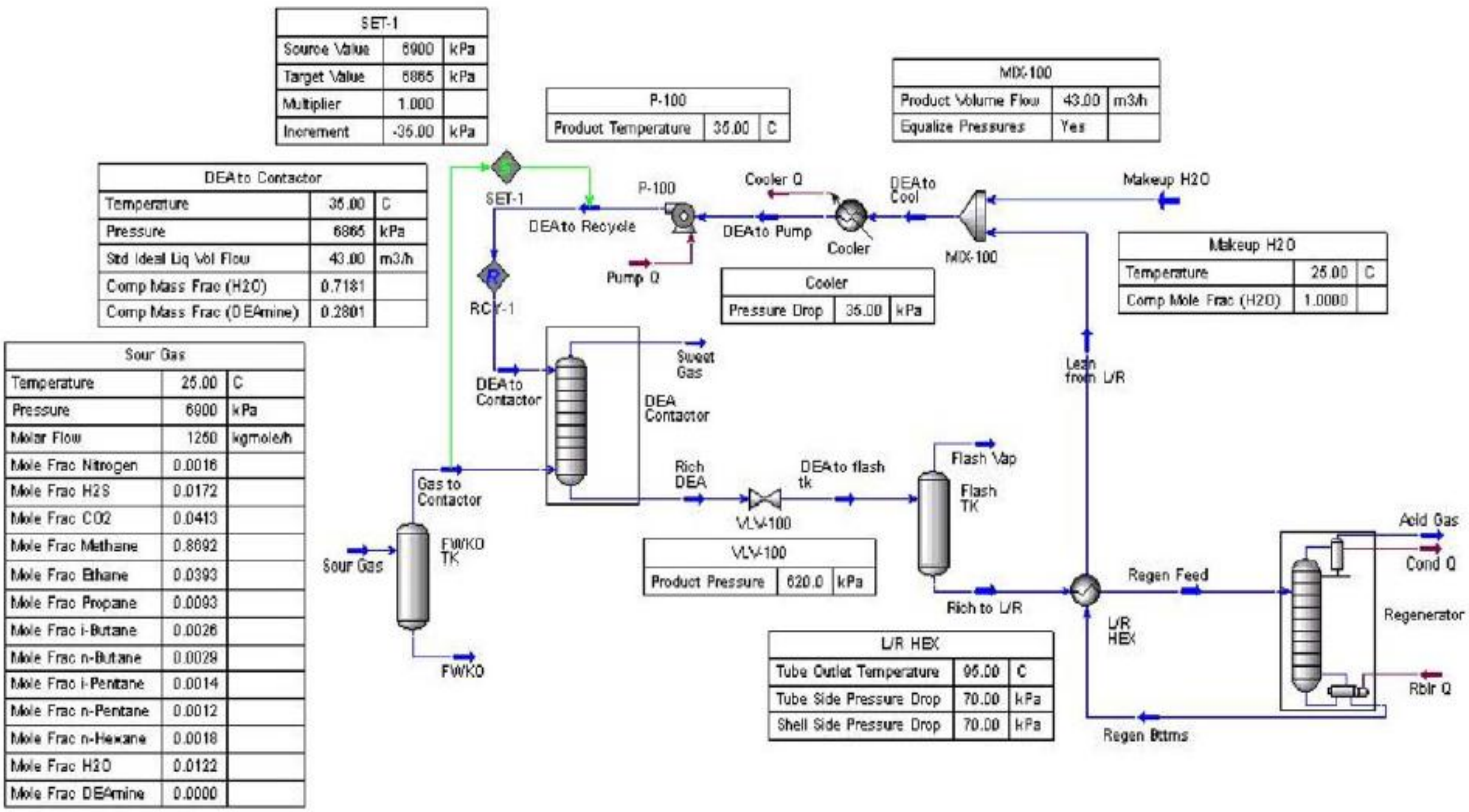

Figure4.1: The overall steady state plant of CO2 removal unit by aqueous alkanolamine

4.2 Modeling

The project of Modeling and predictive control of CO_2 removal unit by aqueous alkanolamine is simulated under Aspen HYSYS 2006, by using Amine property package as simulation basis. It consisted of 2 main column; absorber column/Contractor and stripper column/ Regenerator, separators, pump, heat exchanger, mixer, cooler, and vessel. The initial components of feed gas or sour gas into the separator are as hydrocarbon from C_1 until C_7 and as well as N_2 CO_2 H_2S and DEA Diethanolamine as absorbent to catch CO_2 from the process. The project consisted of 2 modeling which are steady state modeling and dynamic modeling. The details of these modeling are discussed below:

4.2.1Steady state Modeling

The first thing to set up the steady state modeling is selecting the property package that suit the model project. According to this project which deals with aqueous alkanolamine or amine solution the appropriate property package for this modeling is amine property package and use the Li-Mather/Non-Ideal Thermodynamic model as basis. It can predict the behavior of amine hydrocarbon- water systems.

Install stream line and equipments:

The first stream line is the Sour Gas Material Stream and the second main stream line is the DEA to contractor steam line. Details of these 2 streams are shown below in table5 and 6 respectively. Others stream line properties are shown in appendix A3.

Table 4.1: Sour Gas material stream and properties

Sour Gas Material Stream		Sour Gas Material Stream	
N_2	Mole Fraction 0.0016	nC4	Mole Fraction 0.0029
CO_2	Mole Fraction 0.0413	iC_5	Mole Fraction 0.0014
H_2S	Mole Fraction 0.0172	nC_5	Mole Fraction 0.0012
C_1	Mole Fraction 0.8692	nC_6	Mole Fraction 0.0018
C_2	Mole Fraction 0.0393	nC_7	Mole Fraction 0.0072
C_3	Mole Fraction 0.0093	H_2O	Mole Fraction 0.005
iC_4	Mole Fraction 0.0026	DEA amine	Mole Fraction 0.000
SOUR GAS			
Temperature		86.0000 F	
Pressure		1000.0000 psia	
Molar Flow		25 MMSCFD	

Table 4.2: DEA to contractor material stream

DEA TO CONTRACTOR	
Temperature	95 F
Pressure	995 psia
Std Ideal Liq Vol Flow	190 USGPM
CO2 Mass Fraction	0.0018
Water Mass Fraction	0.7187
DEA Mass Fraction	0.2795

Adding main equipments of the process which are V-100 Separator, DEA contractor, Flash TK separator, Regenerator, pump and recycle operation to the aspen HYSYS 2006. Before proceed to Dynamic Modeling make sure all equipment and stream are as steady state condition.

4.2.2Dynamic Modeling

Since the steady state modeling has been converged and stable the second part of the modeling is to convert steady state modeling to Dynamic modeling. The step and details procedure to complete dynamic modeling are as follows.

Converting from Steady State

To complete the dynamics simulation, valve will be installed and pressure flow will be added to selected stream. Equipments will be implemented such as the tray sizing section and all unit operations will sized.

Adding Controllers

Some equipment will be installed and define as manually with appropriate controllers such as pressure transmitter, level transmitter and flow transmitter.

Preparing the Dynamics Simulation

This is the last step to set up the dynamic simulation. The data book or work book that has been shown above at figure 4.2 will be set up. Variables in process are changed and dynamic will be observed.

4.3 Step testing

The step testing is a procedure which planed to choose the possible move and manipulated variables are determined. In this project a 2x2 constrained MPC scheme is developed for CO_2 removal unit. There are 2 minipulated variables and 2 controlled variables the bottom stage of regenerator at tray 17, mole fraction of CO_2 in sour gas1 are controlled variables (CV) and 2 possible manipulated variables (MV) which are the percent opening of over head flow PIC-100 and percent opening of reboiler duty TIC-100.

The initial OP of the process was set as 22.23% for PIC-100 and 52.58% for TIC-100. The manipulated variable set as manual mode, however all others controllers which are not manipulated variables set as auto mode. In this project there are 9 controllers; 2 pressure controllers, 2 temperature controllers, 4 level controllers and 1 flow controller. Before run step testing make sure all the controllers are stable and not fluctuated as can see and check in table detail of each controller.

After all of above controllers are stable, do start step testing for PIC controller which opening as 22.33%. Beside that TIC-100 controller is set as manual mode. Let the process run or moving and show the relation of manipulated variable, control variable and see the response move in strip chart. It shows in term of graph moving of the process. Repeat step testing until reach the target move and repeat the same things for the PIC-100 controller until reach the target move.

The steps testing are applied with 8 step input moves. It starts from the initial 22.23% for PIC and 52.58% for TIC-100 during step testing make sure the process reach the steady state before do the next input move. The step input move of PIC-100 and TIC-100 are shown in table 4.3 below:

Table 4.3: Step Input Moves for PIC-100 and TIC-100 controller

Step input	PIC-100 OP%		TIC-100 OP%	
	Initial	Final	Initial	Final
1	22.23	25.23	52.58	55.58
2	22.23	27.23	52.58	58.58
3	22.23	19.23	52.58	49.58
4	22.23	16.23	52.58	46.58
5	22.23	22.23	52.58	52.58

The graphs below are step testing for each input move PIC-100 and TIC-100

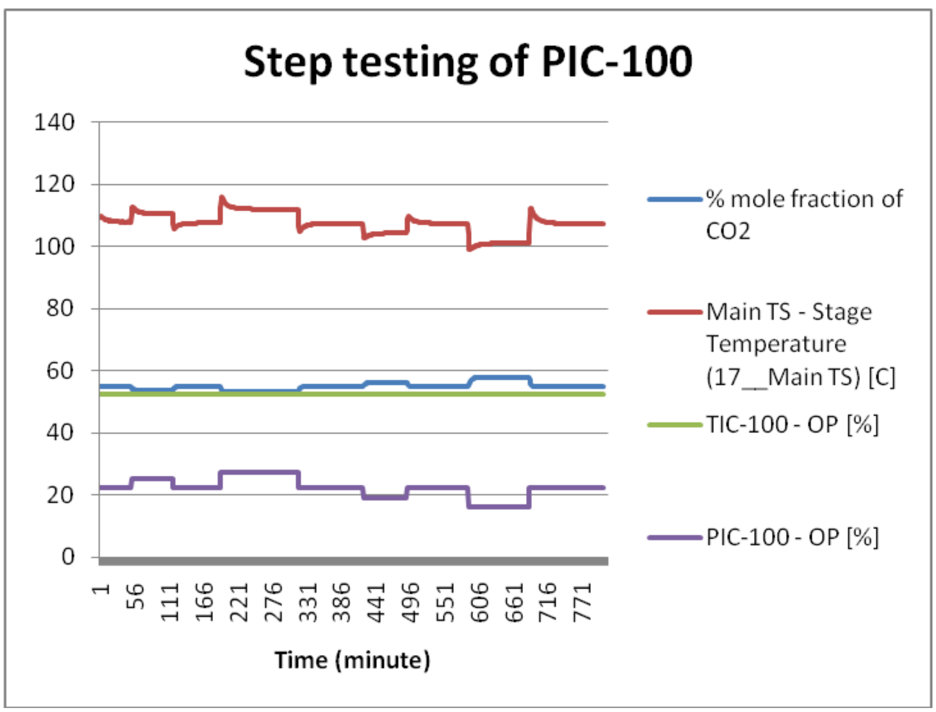

Figure4.2: Step testing of PIC-100 input move

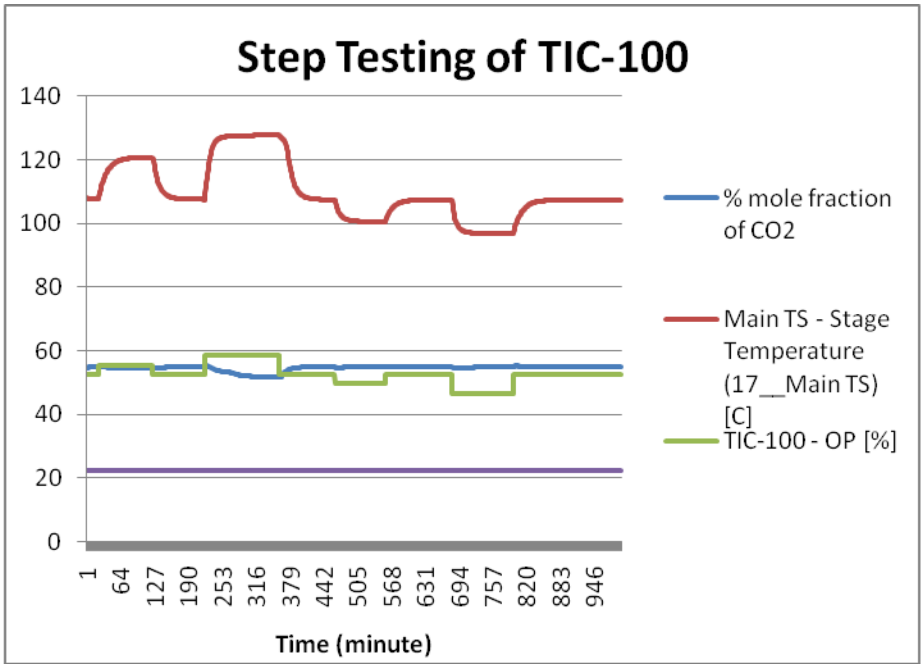

Figure4.3: Step testing of TIC-100 input move

From figure 4.3 the step changes input PIC-100 (OP) is plotted. Each input test is move from the original 22.23% and increases each move by 3%. The first step moves after 20 minute, wait until it reach the steady state point and return to the original condition. Repeat 4 steps move until 400 minute then decrease each move by 3% for 4 steps move. From the graph the long duration is observed to ensure each steps move reach the steady state condition. Meanwhile the control variable or output moves (Main stage temperature at tray no.17 and percentage mole fraction of CO_2) are measured for each input move.

From figure 4.4 the step changes input of TIC-100 (OP) is plotted. The first input move starts at 52.58% of TIC-100 (OP) which is the original point for the second input move or second step testing. Each step increase each move by 2% , the first step moves after 45 minute and wait until it reach the steady state point then return to the original condition. Repeat 4 step moves until 700 minute then decrease each move by 2% for another 3 step move. For this second input move of TIC-100(OP) each move took long duration to reach steady state; it was because the real time factor for the process is low; it is about 0.5-2.50 minute. However this step testing for second input move is done with the long period of 29 hours to complete.

From step testing of first input move and second input move the data collection is recorded in historical data and save it as .csv file. The data of each input move will be used for next step to find the FOPTD (First Order Plus Time Delay) model parameter by using MATHLAB tool for system identification method.

4.4 System identification

The data collected from Aspen HYSYS during the step testing is performed by using MATLAB System Identification Toolbox. The system identification is a step which calculates the mathematical model of dynamic system which measure the input and output of the model.

At the MATLAB tool the variable u1, y1 and y2 are imported to the MATLAB tool

u1 = data(:,6);

y1 = data(:,3);

y2 = data(:,2);

After all variable imported then call function **ident** to open the system identification tool.

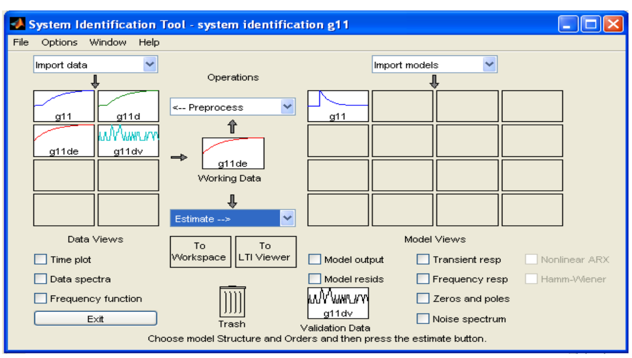

Figure 4.4: System identification in MATHLAB tool

The system identification method is shown in the figure above; firstly import time domain data, second it need to remove means of the data before considering the transfer function and lastly select range of the process. The propose of system identification on MATLAB tool is to find the transfer function of each input and output by using the FOPTD (First Order Plus Time Delay) model below:

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

Where

g_{ij} is transfer function of output and input.

$k_{p,ij}$ is process gain of the process.

$\tau_{p,ij}$ is time constant.

$\tau_{s,ij}$ is time delay.

The transfer function of the process is calculated; since it is a project of 2x2 MPC therefore it will be 4 transfer functions which are shown in the table 4.4 below:

Table 4.4: Model parameter of the process

Transfer function	**Model parameter**		
	Kp	τ_p **(min)**	τ_s**(min)**
g_{11}	**-0.35**	**1.799**	**0**
g_{12}	**-0.32**	**18.93**	**6.1**
g_{21}	**0.99**	**0.026**	**0**
g_{22}	**3.25**	**12.18**	**2.5**

In this model parameter of the process the kp, τ_p and τ_s are shown in the table above. The process gain value of g_{11} and g_{12} transfer functions are negative value of -0.35 and -0.32 respectively. The negative value of the process gain shows the reverse of the process when increasing in input (manipulated variable) the output (controlled variable) decrease and vice versa. The transfer functions of g_{12} , and g_{22} have delayed in response while the delayed in response of g_{11} and g_{21} are zero. After get all value then installs and adds the value of model parameter of the process in MPC controller in ASPEN HYSYS tool.

4.5 Install MPC

MPC controller is installed, in MPC setup enable MPC modifications-MPC control setup as 2x2 inputs and outputs. While control interval of MPC is 1 minute and the process mode type of MPC is defined as first order model. The connection of MPC controller is connected to the input and output which already define from previous step. The controlled variables of main stage temperature at tray no.17 and percentage mole fraction of CO_2 are connected in the process variable sources, meanwhile the manipulated variable of percent opening of over head flow (PIC-100) and percent opening of reboiler duty (TIC-100) are connected as output target object. In the parameters of MPC controller the PV minimum and PV maximum of main stage temperature at tray no.17 and percentage mole fraction of CO_2 are required; the PV minimum are 80^0C and 0 and PV maximum are 150 ^{0}C and 100 respectively.

The MPC controller has been installed, the next step is then comparing the performance of PI controller with MPC controller by using disturbance rejection. This test is compared in term of the capability to reach and maintain the steady stage of the controlled variable with the noise active variance 5%, 10%, 15% and 20% respectively. The first test by MPC controller are set as PIC-100 and TIC-100 as off mode while MPC controller is set as auto mode which can refer to the figure below.

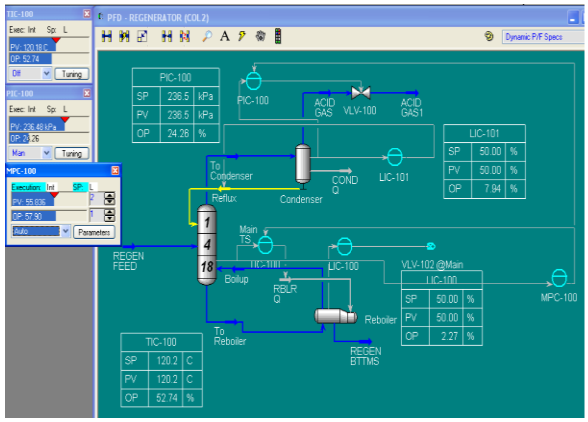

Figure 4.5: Install MPC controller

4.6 Compare MPC and PI controller by disturbance rejection

After finished install MPC, the disturbance rejection method is introduced to test the existing PI controllers which are PIC-100 and TIC-100 with noise active variance 5%, 10%, 15% and 20% respectively. While testing PI controller the PIC -100 and TIC-100 are set as auto mode and MPC controller is set as off mode. Meanwhile, during testing MPC controller, MPC is set as auto mode while PI controllers are set as off mode.

From the graph below in figure 4.8 and figure 4.9 show the comparison of ability to maintain the set point of each controlled variable. Figure 4.8 shows the set point of the process for Percentage mole fraction of $CO_{2.}$ The red line in figure 4.8 represents the set point of process of Percentage mole fraction of CO_2 (52.98). The green line in figure 4.9 represent the set point of main stage temperature at tray no.17
which is the value where MPC controller and PI controller have to achieve to compare the performance of MPC controller and PI controller.

The result of both testing in MPC disturbance rejection and PI disturbance rejection are show in figure below:

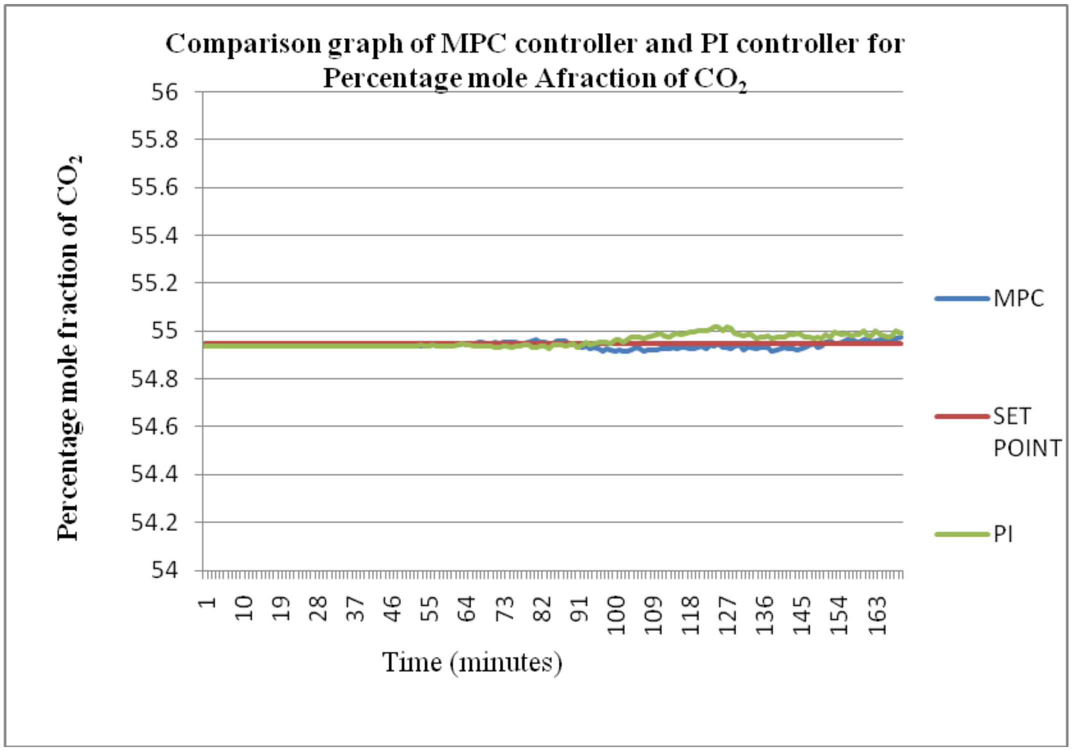

Figure 4.6: Comparison graph of MPC controller and PI controller for Percentage mole fraction of CO_2

According to graph above in figure 4.8 shows the performance of both MPC controller and PI controller to achieve the set point of Percentage mole fraction of CO_2 . The result from time 0 the MPC controller and PI controller are at steady state point which disturbance rejection is not starts yet. The disturbance rejection of 5% noise is introduced at time 50 minute then follows with 10%, 15% and 20% of noise variance during time 80, 110 and 140 minutes respectively. During starting of 5% noise the PI controller and MPC controller are in range of set point which still not able to compare the performance of it. However, at 80 minute 10% of noise is introduced to the process, the graph move of PI controller and MPC controller seem fluctuate at this point, the PI controller and MPC controller try to achieve and reach the set point. At time100 minute, the observations of MPC controller seem better than the graph of PI controller according to its achievement and maintain at the set point. At time 120 to 140 minute the PI controller graph is out of set point range which can conclude thatt the performance to achieve set point of MPC controller is better than PI controller.

Next is comparison graph of MPC controller and PI controller for main stage temperature at tray no.17 shows below:

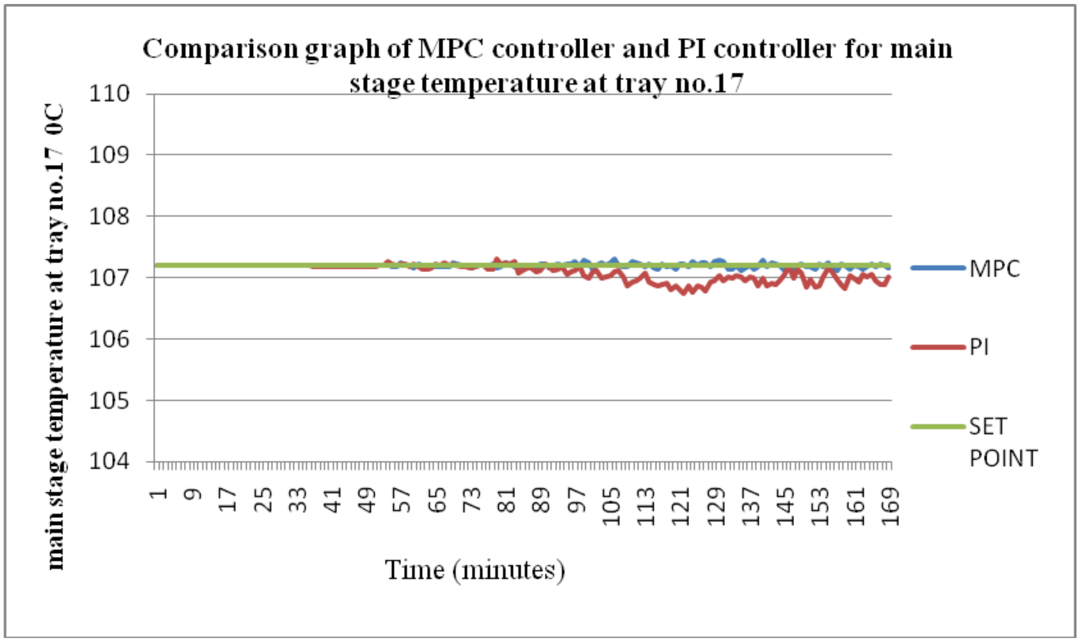

Figure 4.7: Comparison graph of MPC controller and PI controller for main stage temperature at tray no.17

The disturbance rejection step for figure 4.9 above is similar with above procedure of figure 4.8. The result of the graph at each disturbance is introduced noise at feed as 5%,10%, 15% and 20% in 50, 80, 110, and 140 respectively. The green line of graph in figure 4.9 is a set point of main stage temperature at tray 17 which is 107.2 ^{0}C.

The graph above clearly sees that, at time 80 minute which 10% noise is introduced PI controller out of range and far from the set point meanwhile the MPC controller is achieve the set point of the process. The performance of MPC is higher than PI according to its achievement and maintain at set point.

According to the figure 4.8 and 4.9, MPC is a better Performance than PI controller to achieve and maintain at set point of Percentage mole fraction of CO_2 and set point of main stage temperature at tray no.17. The achievement and maintain at set point in process impact the value of CO_2 release, it means that the higher performance to achieve set point help process to control the CO_2 release and percent opening of reboler duty at TIC-100. Therefore, develop Modeling and Predictive Control in amine adsorption technology is good practice to reduce carbon dioxide emissions of natural gas manufacturer and minimize energy use of reboiler duty than existing PI controller.

However this project has been done by Bedelbayev et al. in the title of Dynamic Modeling to Minimize Energy Used for CO_2 Capture in Power Plant by Aqueous Monoethanolamine. The project is applied for CO_2 removal unit at absorber by step testing and at the end MPC also minimize energy used in the process.

CONCLUSION

Modeling and predictive control for carbon dioxide removal unit by aqueous solution is a project concerns on how to reduce energy consumption and remove carbon dioxide emission. According to its disadvantage to the process, equipment plant and environment (Global warming), many technologies nowadays are available to remove and eliminate carbon dioxide such as by chemical absorption process. Due to that, the related topic of this project is to reduce carbon dioxide of natural gas manufacturer by applying the Modeling and Predictive control in amine adsorption technology.

The performance of MPC controller is compared with PI controller in term of disturbance rejection. The result of MPC controller is better performance to maintain the set point for percentage mole fraction of CO_2 and main stage temperature at tray no.17 than PI controller. Therefore, develop Modeling and Predictive Control in amine adsorption technology is good practice to reduce carbon dioxide emissions of natural gas manufacturer and minimize energy use of reboiler duty than existing PI controller.

REFERENCES

Abedini R., Abedini A. b, Zanganeh I. a., 2010, Investigation of mixed amine solution in gas sweetening plant. Journal of Natural Gas Science and Engineering 2: 36-40

Ahmada A.L., Sunarti A.R., Lee K.T. a, Fernando W.J.N., 2010, CO_2 removal using membrane gas absorption. International Journal of Greenhouse Gas Control 4:495–498

Barry B. and Lili L., 2000, “A comparison of physical solvents for acid gas removal”, Texas; U.S.A.

Beldelbayev A., Greer T., Lie B., 2009, “Model based control of absorption tower for CO_2 capture”, Norway.

Camacho, Bordons C., 1999, Model Predictive control, London, Springer.

Camacho E.F and Bordons C., 2004, Model Predictive Control. Second edition, London, Springer.

Dale E. Seborg, Thomas F. Edgar, Duncan A. Mellichamp, 2003, Process Dynamics and Control. Second Edition, United States of America., John Wiley & Sons,Inc.

Georgios A. Florides , Paul Christodoulides, 2009, Global warming and carbon dioxide through sciences. Environment International 35: 390–401

Gotevorg, 2007, “Aspen HYSYS Simulation of CO_2 Removal by Amine Absorption from a gas based power plant”, Telemark University College, Norway

Haitao Huang, James B. Riggs, 2002, Comparison of PI and MPC for control of a gas recovery unit. Journal of Process Control 12: 163-173.

Huang Biao, Kadali Ramesh, 2008, Dynamic Modeling, Predictive Control and Performance Monitoring. Canada: Springer.

Mandala B. P., Guhab M., Biswasb A. K., Bandyopadhyaya S. S., 2001, Removal of carbon dioxide by absorption in mixed amines: modeling of absorption in aqueous MDEA=MEA and AMP=MEA solutions. Chemical Engineering Science56: 6217–6224

Porfirio C.R., Almeida E., Neto, Odloak, 2003, Multi-model predictive control of an industrial C3/C4 splitter. Control Engineering Practice 11:765-779

Qin S.J. and Badgwell T.A. (2002). A survey of industrial model predictive control technology. Control Engineering Practice 11 (2003), 733-764

Sepideh Z.,Gray T., Thomas F. Edgar, 2009, Dynamic Modeling to Minimize Energy Use for CO_2 Capture in Power Plant by Aqueous Monoethanolamine. Department of chemical engineering, 6105-611

Tennyson R.N. and Schaff R.P., Guidelines can help choose proper process for gas treating plant. Oil and Gas Journal, Vol 75, No2, pp 78-86.

Youchang Xiaoa, Bee Ting Lowa, Seyed Saeid Hosseinia,Tai Shung Chunga,b, Donald Ross Paulc, 2010, The strategies of molecular architecture and modification of polyimide-based membranes for CO_2 removal from natural gas—A review. Progress in Polymer Science 34 :561–580

.APPENDIXES

Appendix A: Type of constraint

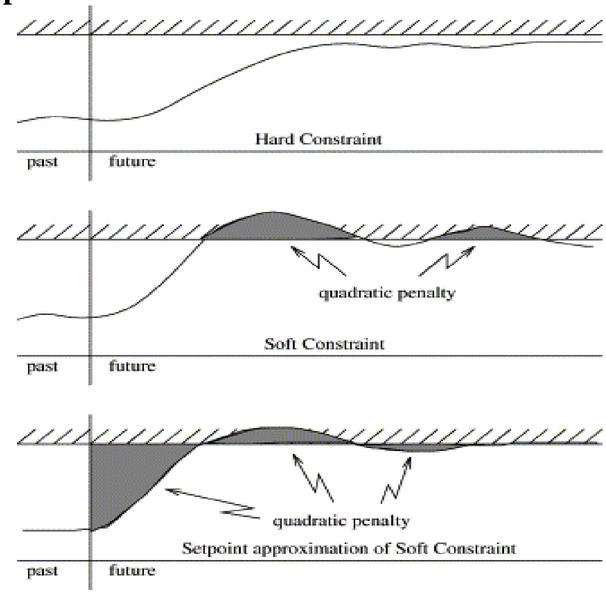

Figure A1: The three basic type of constraint

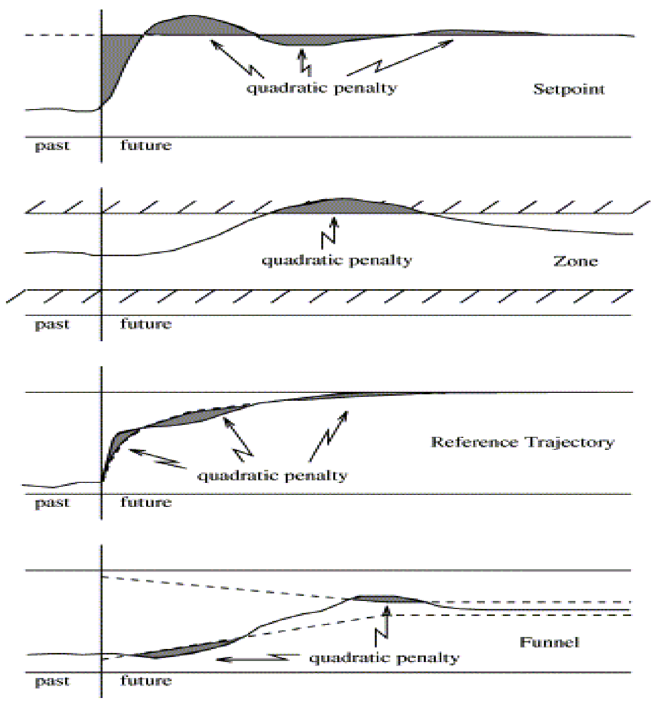

Figure A2 Four options for specifying future CV behavior

Workbook - Case (Main)

Name	DEA TO CONT	SOUR GAS	GAS TO CONTA	FWKO	SWEET GAS	RICH DEA	DEA TO FLASH 1
Vapour Fraction	0.0000	0.9906	1.0000	0.0000	1.0000	0.0000	0.0011
Temperature [F]	95.22	86.00	86.00	86.00	95.22	141.0	140.9
Pressure [psia]	995.0	1000	1000	1000	994.0	1006	90.00
Molar Flow [lbmole/hr]	4156	2745	2718	44.51	2554	4319	4319
Mass Flow [lb/hr]	9.790e+004	5.342e+004	5.223e+004	1527	4.561e+004	1.045e+005	1.045e+005
Liquid Volume Flow [barrel/day]	6553	1.052e+004	1.039e+004	147.8	9824	7118	7118
Heat Flow [Btu/hr]	-4.938e+007	1.619e+007	1.608e+007	-1.453e+005	1.562e+007	-4.891e+007	-4.891e+007

Name	FLASH VAP	RICH TO L/R	REGEN FEED	LEAN FROM L/R	REGEN BTTMS	ACID GAS	MAKEUP H20
Vapour Fraction	1.0000	0.0001	0.0113	0.0000	0.0000	1.0000	0.0000
Temperature [F]	140.9	141.0	198.4	196.5	261.6	178.5	70.00
Pressure [psia]	90.00	44.13	33.78	21.50	35.79	26.50	21.50
Molar Flow [lbmole/hr]	4.724	4312	4381	4176	4173	199.9	31.27
Mass Flow [lb/hr]	92.40	1.044e+005	1.060e+005	9.868e+004	9.874e+004	7162	563.3
Liquid Volume Flow [barrel/day]	17.17	7096	7210	6607	6611	588.0	38.65
Heat Flow [Btu/hr]	2.841e+004	-4.889e+007	-4.314e+007	-4.034e+007	-3.402e+007	1.004e+006	-4.624e+005

Name	DEA TO COOL	DEA TO PUMP	DEA TO RECY	REGEN-BTTMS1	Nitrogen Blanket	FWKO-1	FLASH VAP-1
Vapour Fraction	0.0000	0.0000	0.0000	0.0000	1.0000	0.0023	1.0000
Temperature [F]	195.8	90.99	95.22	263.6	195.8	86.00	140.9
Pressure [psia]	20.69	15.65	995.0	35.79	21.50	986.5	89.99
Molar Flow [lbmole/hr]	4158	4156	4156	4173	3.719e-003	44.51	4.724
Mass Flow [lb/hr]	9.802e+004	9.791e+004	9.791e+004	9.874e+004	8.708e-002	1527	92.40
Liquid Volume Flow [barrel/day]	6563	6554	6554	6611	6.888e-003	147.8	17.17
Heat Flow [Btu/hr]	-4.034e+007	-4.975e+007	-4.938e+007	-3.382e+007	18.05	-1.453e+005	2.841e+004

Name	RICH TO VALVE	RICH TO PUMP	REGEN BTTMS	REGEN BTTMS-	DEA TO VALVE	** New **
Vapour Fraction	0.0000	0.0000	0.0000	0.0007	0.0000	
Temperature [F]	141.0	140.9	263.6	262.3	195.8	
Pressure [psia]	99.83	90.00	525.3	35.79	21.50	
Molar Flow [lbmole/hr]	4312	4312	4173	4173	4158	
Mass Flow [lb/hr]	1.044e+005	1.044e+005	9.874e+004	9.874e+004	9.802e+004	
Liquid Volume Flow [barrel/day]	7096	7096	6611	6611	6563	
Heat Flow [Btu/hr]	-4.889e+007	-4.889e+007	-3.382e+007	-3.388e+007	-4.034e+007	

Material Streams | Compositions | Energy Streams | Unit Ops

FigureA3: Project’s work book of CO_2 removal unit by aqueous alkanolamine.