UNIVERSITI MALAYSIA PAHANG

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EVALUATION OF EFFECTIVENESS PARAMETERS ON GAS DEHYDRATION PLANT

WINNIE LO CHEN SING

Thesis submitted in partial fulfillment of the requirements for the award of the degree of Bachelor of Chemical Engineering (Gas Technology)

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SUPERVISOR'S DECLARATION

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Special Dedication to my supervisor, my family members, my friends, my fellow colleague and all faculty members for all your care, support and believe in me.

ACKNOWLEDGEMENTS

I would like to express my eternal gratitude and sincere appreciation to my supervisor, Associate Professor Dr. Abdurahman Hamid Nour, and Sir Abdul Halim for their invaluable guidance, empowering support and profound advice throughout this research project.

My heartfelt gratitude also goes to my colleagues from University Malaysia Pahang who was respondents for this study, for their full support and co-operation to make this study possible.

I would like to express my appreciation to all my friends and family for the endless support and love on me.

ABSTRACT

Dehydration of Natural Gas to controlled water content is necessary in order to avoid gas hydrates and minimize corrosion. This study describes the dehydration on liquid desiccant dehydration unit. The key task was to identify the optimum parameters used in Triethylene Glycol dehydration unit. Absorption dehydration involves the use of Triethylene Glycol to remove water vapor from the gas stream. Water was removed to meet a water dew point requirement of sale gas contract specification range from 32.8 to $117 \text{ kg}/10^6$ standard m³. The evaluation of effectiveness parameters on gas dehydration plant was carried out by using ASPEN HYSYS simulation. ASPEN HYSYS was used for steady state simulation, design, performance monitoring and optimization of oil and gas production, gas processing and petroleum refining industries. Peng-Robinson equation of state was chosen in the system. The simulation was carried out to determine the effect of important parameters such as column operating pressure, number of theoretical stages of column and gas flow rate. A technically optimized dehydration process has been proposed based on the simulated data. Results showed that, when the gas flow rate increases, the water absorption rate will decrease. Same condition obtained as increasing pressure. However, water absorption rate increases with the increasing number of theoretical stages. The optimum condition of gas dehydration was at 2000kPa with 1000m³/h gas flow rate and 4 column theoretical stages.

Key words: Dehydration, hydrates, Triethylene Glycol, Aspen Hysys, absorption, optimization.

ABSTRAK

Penyahhidratan gas asli kepada kandungan air yang terkawal adalah diperlukan untuk mengelakkan pembentukkan hidrat gas dan meminimumkan pengaratan. Kajian ini telah menerangkan pengyahhidratan dengan menggunakan bahan pengering cecair dalam unit pengyahhidratan. Tujuan utama kajian ini adalah untuk mengenal pasti parameter optimum yang digunakan dalam unit pengyahhidratan yang menggunakan Triethylene Glycol. Penyahhidratan penyerapan melibatkan penggunaan Triethylene Glycol untuk memisahkan wap air dari aliran gas. Wap air dipisah untuk memenuhi keperluan titik embun air yang ditetapkan oleh specifikasi kontrak jualan gas yang merangkai dari 32.8 ke 117 kg/ 10^6 standard m³. Penilaian parameter yang keberkesanan pada unit penyahhidratan dilaksanakan dengan menggunakan simulasi ASPEN HYSYS. Biasanya, ASPEN HYSYS digunakan dalam keadaan simulasi yang stabil, reka bentuk, pemantauan penampilan, dan pengoptimuman pengeluaran minyak dan gas ,serta industry pemprosesan gas dan penapisan petroleum. Persamaan Peng-Robinson adalah dipilih dalam sistem ini. Simulasi ini dijalankan untuk menentukan kesan-kesan parameter penting seperti tekanan operasi kolum, bilangan dulang kolum, dan kadar aliran gas. Process penyahhidratan yang dioptimumkan secara teknikal telah dicadangkan berdasarkan data-data simulasi. Sebagai keputusan, semasa kadar aliran gas meningkat, kadar penyerapan air akan menurun. Keadaan yang sama berlaku jika tekanan kolum meningkat. Akan tetapi, kadar penyerapan air akan meningkat dengan pertambahan bilangan dulang kolum. Keadaan penyahhidratan gas yang optimum berlaku pada 2000kPa dengan 1000m³/h kadar aliran gas dan 4 dulang kolum.

Kata kunci: Penyahhidratan, hidrat, Triethylene Glycol, Aspen Hysys, penyerapan,

pengoptimuman.

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LIST OF SYMBOLS

| W _r | Water removed |
|---------------------------|--|
| \mathbf{W}_{i} | Water content of inlet gas |
| Wo | Water content of outlet gas |
| Q_{G} | Gas flow rate |
| L | Liquid molar flow rate |
| Х | Mole fraction in liquid phase |
| F | Feed molar flow rate |
| l | Refers to liquid phase |
| Κ | Equilibrium constant |
| \overline{H}_i | Fractional molar enthalpy of component I in liquid phase |
| H_i^+ | Fractional molar enthalpy of component i in ideal gas state |
| φ_i^v | Fugacity coefficient of gas phase |
| γ_i | Activity coefficient |
| Р | Pressure |
| V | Molar volume |
| Т | Temperature |
| R | Universal gas constant (8.314 J.K ⁻¹ .mol ⁻¹) |
| T _c | Critical temperature |
| P _c | Critical pressure |

α Alpha function, function of reduced temperature

LIST OF ABBREVIATIONS

| CO_2 | Carbon dioxide |
|--------|--|
| H_2S | Hydrogen sulfide |
| TEG | Triethylene Glycol |
| SI | International System of unit |
| PFD | Process Flow Diagram |
| BTEX | Emission of Aromatic, e.g. Benzene, toluene, ethylbenzene, xylenes |
| VOC | Volatile Organic Compounds |
| | |

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CHAPTER 1

INTRODUCTION

1.0 INTRODUCTION

1.1 BACKGROUND OF STUDY

Gas dehydration is widely used in natural gas treatment plant as a common process and it removes water that is associated with natural gases in vapor form. The natural gas industry has recognized that dehydration is necessary to ensure smooth operation of gas transmission lines as water and hydrocarbon can form gas hydrates which may block valves and pipelines.

Several methods have been developed to dehydrate gases on an industrial scale. The three major methods of dehydration are direct cooling, absorption of water in glycol, and adsorption of water by solid. In absorption processes, the absorption/stripping cycle is used for removing large amounts of water, and adsorption is used for cryogenic systems to reach low moisture contents and does not involve any chemical reaction. For commercial dehydration purpose, high absorption efficiency, and easy and economic regeneration should be possessed by those dehydrating agent. They should be noncorrosive and non-toxic, no operation problems when used in high concentration, no interaction with the hydrocarbon portion of the gas, and no contamination by acid gases. When optimizing the design of dehydration facilities, the impact of number of trays in the contactor, liquid desiccant circulation rate through the contactor, temperature of the reboiler in the regenerator, amount of stripping gas used, and operating pressure of the regenerator should be considered.

1.2 PROBLEM STATEMENT

Natural gas is a mixture of many components which is classified into 3 major groups, that is hydrocarbons containing Hydrogen and Carbon, inert elements and trace compounds. Natural gas hydrates are solids that formed from natural gas hydrocarbons and water. The water molecules have a honeycomb structure with a molecule of one of the natural gas components occupying each void. Since these solids are denser than water ice, their formation is favored at higher pressure. Natural gas hydrates may form and interfere with the passage of natural gas through valves and pipes. These may block pipeline flow and control systems. Natural gas in transit needs to be dehydrated to a controlled water content in order to avoid gas hydrates and to minimize corrosion.

1.3 RESEARCH OBJECTIVES

- 1.3.1 To investigate the effective parameters on Natural Gas Dehydration Plant.
- 1.3.2 To evaluate the optimum parameters of triethylene glycol dehydration unit by using Aspen HYSYS.

1.4 SCOPE OF STUDY

The scope of study was to study the optimum parameters of triethylene glycol, which were number of theoretical stages, pressure, and gas flow rate. Optimization of parameters will be made based on the performance of drying agent which is triethylene glycol. Absorption dehydration involves the use of a triethylene glycol to removal water vapor from the gas.

1.5 SIGNIFICANCE OF STUDY

Dehydration to the gas will be subjected to prevent hydrate formation and corrosion from condensed water. The latter consideration is especially important in gas streams containing CO_2 or H_2S where the acid gas components will form an acid with the condensed water.

Natural gases either from natural production or storage reservoirs contain water, which condense and form solid gas hydrates to block pipeline flow and especially control systems. Natural gas in transit to market should be dehydrated to a controlled water content to avoid hydrate as well as to minimize the corrosion problems.

CHAPTER 2

LITERATURE REVIEW

2.0 LITERATURE REVIEW

2.1 INTRODUCTION

For hundreds of years, natural gas has been known as a very useful substance. The Chinese discovered a very long time ago that the energy in natural gas could be harnessed, and used to heat water. In the early days of the natural gas industry, the gas was mainly used to light streetlamps, and the occasional house. However, with much improved distribution channels and technological advancements, natural gas is being used in ways never thought possible. According to the Energy Information Administration, energy from natural gas accounts for 23% of total energy consumed in the United States, making it a vital component of the nation's energy supply. Natural gas is used across all sectors, in varying amounts. The industrial sector accounts for the greatest proportion of natural gas use in the United States, with the residential sector consuming the second greatest quantity of natural gas.

However, natural gases either from natural production or storage reservoirs contain water, which condense and form solid gas hydrates to block pipeline flow and especially control systems. Natural gas in transit to market should be dehydrated to a controlled water content to avoid hydrate as well as to minimize the corrosion problems.

2.2 INTRODUCTION OF DEHYDRATION

Natural gas contains many contaminants which the most common impurity is water. Most natural gas will be near water saturation at the temperature and pressure of production. Dehydration of natural gas is the removal of the water that is associated with natural gases in vapor form. The natural gas industry has recognized that dehydration is necessary to ensure smooth operation of gas transmission lines. Removal of water from the gas stream reduces the potential for corrosion, gas hydrates formation and freezing in the pipeline. Unless gases are dehydrated, liquid water may condense in pipelines and accumulate at low points along the line, reducing its flow capacity. Water is removed to meet a water dew point requirement of a sale gas contract specification range from 32.8 to $117 \text{ kg}/10^6$ std m³(Gandhidasan P. et al., 2001). Several methods have been developed to dehydrate gases on an industrial scale.

2.3 THEORY OF ABSORPTION

Absorption is a physical or chemical process in which atoms, molecules, or ions enter some bulk phase, means gas, liquid or solid material. This is a different process from adsorption, since molecules undergoing absorption are taken up by the volume, not by the surface. (McMurry,J & McDonald,A , 2003). Absorption dehydration involves the use of liquid desiccant to remove water vapor from the gas. The liquid that is most desirable to use for commercial dehydration purposes should possess high absorption efficiency, and easy and economic regeneration. It should be non-corrosive and non-toxic, no operation problems when used in high concentration, no interaction with the hydrocarbon portion of the gas, and no contamination by acid gases. There are numbers of liquids that can be used to absorb water from natural gases such as calcium chloride, lithium chloride, and glycols. Triethylene glycol (2-2'-(1,2-ethanediylbis(oxy))ethanol) is a stable, noncorrosive chemical with high flash point. It is a straight-chain dihydric alcohol aliphatic compound terminated on both ends by a hydroxyl group. It is a clear, practically colorless and odorless, hygroscopic liquid at room temperature. It is used as a dehydrating agent for natural gas; a solvent and lubricant in textile dyeing and printing; a plasticizer, a raw material for the production of polyester resins and polyols; a humectants; a constituent of hydraulic fluids; a selective solvent for aromatics (Huntsman, 2007). The structure of triethylene glycol is shown as follow:



Figure 2.1: Straight-chain dihydric alcohol aliphatic compound - TEG

2.4 PREVIOUS STUDY ON LIQUID DESICCANT DEHYDRATION

As said by Honerkamp et al. (1983), there are a variety of ways for separating water from gas, but the method most commonly employed in the petroleum industry is the use of glycol or other liquid desiccant dehydrators. Glycol readily mixes with water and has a lower vapor pressure than water. Thus, it is well known to pass wet gas into contact with glycol wherein the water in the gas is absorbed by the glycol. The gas is then passed in an upward column wherein the glycol/water mixture is then heated to drive off the water, the dry or anhydrous glycol then being recirculated to contact wet gas in a continuous process so that water is extracted from a gas stream.

According to Smith,R.S. (1996), liquid desiccant system are relatively simple to operate and easy to maintain. However, it typically unable to produce treated gases with extremely low level of moisture. Solid desiccant systems are often used to provide gas with very low levels of moisture; however these plants can be more complex and expensive to operate than liquid desiccant systems. They also present a higher risk of downstream damage by failure of automatic regeneration switching valves. So, there is a continuing need for a relatively simple liquid desiccant gas dehydration system that produces gas with the low moisture content normally associated with solid desiccant systems.

As stated by Gandhidasan,P et al (2001), the glycols have proved to be the most effective liquid desiccants in current use since they have high hygroscopicity, low vapor pressure, high boiling points and low solubility in and of natural gas. Triethylene Glycol has gained nearly universal acceptance as the most cost effective of the glycols due to superior dew pint depression, operating cost and operation reliability. However, there are several operating problems with glycol dehydrators. Suspended foreign matter may contaminate glycol solution. Overheating may produce both low and high boiling decomposition products. Resultant sludge may collect on heating surfaces and causing loss in efficiency. High concentrated glycol solutions tend to become viscous at low temperature and therefore hard to pump. Glycol lines may solidify completely at low temperature when the plant is not operating. Furthermore, there are substantial environmental problems due to fugitive emission, soil contamination, and fluid disposal problems.

From Natural Gas Transmission and Processing Handbook (2006), the amount of water to be removed in a TEG system is calculated from the gas flow rate, the water content of incoming gas, and the desired water content of outgoing gas. By assuming the inlet gas is saturated with water, the water removal rate can be determined as $W_r = \frac{Q_G(W_i - W_o)}{24}$, where W_r is water removed, lb/hr; W_i is water content of inlet gas, lb/MMscf; W_o is water content of outlet gas, lb/MMscf; and Q_G gas flow rate, MMscfd.

The glycol circulation rate is determined on the basis of the amount of water to be removed and is usually between 2 and 6 gallons of TEG per pound of water removed, with 3 gallons TEG/lb water being typical. Higher circulation rates provide little additional dehydration while increasing reboiler fuel and pumping requirements. Problems can arise if the TEG recirculation rate is too low. Therefore, a certain amount of overcirculation is desired. An excessive circulation rate may overload the reboiler and prevent good glycol regeneration. The heat required by the reboiler is directly proportional to the circulation rate. So, an increase in circulation rate may decrease the reboiler temperature, decreasing lean glycol concentration, and decrease the amount of water that is removed by the glycol from the gas. To lower the dew point of the gas, the circulation rate can be increased and the reboiler temperature must remain constant (Saeid M. et. al., 2006).

From Saeid M. et. al (2006), usually the glycol absorber contains 6-12 trays that provide an adequate contact area between the gas and the glycol. With increasing number of trays, the greater the dew point depression for a constant glycol circulation rate and lean glycol concentration. On the contrary, specifying more trays with the same TEG concentration, a lower circulation rate is required. By specifying more trays, fuel savings can be realized because the heat duty of the reboiler is directly related to the glycol circulation rate. Also, calculated tray efficiency values depend on the TEG/water equilibrium data used.

Since absorption is an exothermic process, increasing input TEG temperature decreases water absorption rate. In real dehydration plants, temperature of TEG entering the absorption tower is kept at 5-6°C more than entering gas temperature (Kasiri N. and Hormozdi Sh., 2005). Water absorption rate as a function of TEG temperature is represented in the following figure:



Figure 2.2: Effect of TEG temperature on water absorption rate

Increasing number of equilibrium stages increases water absorption rate as well as manufacture and maintenance costs (Kasiri N. and Hormozdi Sh., 2005). As number of stages increases, glycol circulation rate decreases. Percentage of water removal as a function of number of equilibrium stages in absorber is represented in the following figure:



Figure 2.3: Effect of number of equilibrium stages in contactor on percentage of water removal

At a constant temperature, the water content of the inlet gas decreases with an increasing pressure. Less water is removed if the gas is dehydrated at a higher pressure (Mohamadbeigy Kh., 2008). At lower pressure, wall thickness required is lesser. Therefore, an economic trade-off exists between operating pressure and contactor cost.



Figure 2.4: Effect of pressure in contactor on the water content of gas stream

However, there are also some operating problems associated with the absorber, which are insufficient dehydration, foaming, and hydrocarbon solubility in glycol. From Saeid M. et. al.(2006), lean glycol purity plays a main role in the rate of water removal. A minimum lean glycol concentration is needed to achieve a specified dew point depression. Higher water concentrations in the lean glycol results in poor dehydration. Temperature of the inlet gas dictates the amount of water fed into the unit. A lower inlet gas temperature will require less water to be removed by the glycol. Lean glycol temperature at the top of the absorber will affect the water partial pressure at the top stage, where high TEG temperature may cause high moisture content of the outlet gas. Reboiler temperature can therefore be increased up to 400°F above which glycol degradation starts.

From Handbook of Natural Gas Transmission and Processing (2006), due to poor heat transfer, lean glycol results in too warm temperature and poor dehydration and insufficient dew point depression can be resulted. Also, glycol vaporization losses to the product gas may be higher with increased lean glycol temperature. Poor heat transfer and the resulting high lean glycol temperature may be caused by fouled heat exchangers, undersized heat exchangers, or over circulation. Exchangers may be fouled by deposits such as salt, solids, coke, or gum. In the case of undersized exchangers, additional heat exchangers may be required.

From Saeid M. et. al. (2006), pump reliability is enhances by limiting the lean glycol temperature from 180°F to 200°F and ensuring good filtration. Pump wear, leakage, and failures increase if the glycol becomes dirty or hot. It is possible that the seals on the glycol balance pumps wear out and contaminate the lean glycol by the rich glycol. This increases the water content of the lean glycol and may cause the gas no longer to be dried to pipeline specifications. Excessively high glycol circulation rates can lead to many problems. If the unit is over circulating the glycol, the lean glycol may have insufficient heat exchangers to be cooled properly, and the resulting hot lean glycol may not achieve the desired water removal rate. A high circulation rate may not allow adequate residence time in the phase separator for the hydrocarbons to be removed, which may lead to hydrocarbon deposits, glycol losses, foaming, and emissions. Excessive glycol circulation rates can also result in increased sensible heat requirements in the reboiler. Also, as emissions are proportional to the circulation rate, over circulation results in greater VOCs emissions. However, under circulating the glycol provides an insufficient quantity of glycol in the absorber for the quantity of water to be removed and results in wet sales gas.

2.5 ABSORPTION MODEL DESCRIPTION

There are four sets of equations, mass balance, equilibrium relations, sum of mole fractions of each phase and heat balance are used to describe tray columns treatment. Material balance of component i on tray j, for liquid phase is as follows (Kasiri.N, and Hormozdi.Sh., 2005) :

$$L_{j}x_{i,j} - L_{j+1}x_{j+1} - F'_{i,j} = 0$$
[1]

where L is liquid molar flow rate, x is mole fraction in liquid phase, F is feed molar flow rate and the superscript l refers to liquid phase. If the impact of entrainment factor in vapor phase is considered, the following equation will be obtained:

$$(1 + E_{j}^{l}) L_{j} x_{i,j} - L_{j+1} x_{i,j+1} - E_{j-1}^{l} L_{j-1} x_{l,j-1} - F_{i,j}^{l} = 0$$
[2]

Equilibrium relation for component *i* at tray *j* is given by:

$$K_{i,j} = y_{i,j} / x_{i,j}$$
 [3]

where K is equilibrium constant. Sum of mole fraction of each phase are shown as follows:

$$\sum_{i=1}^{N} y_{i,j} = 1 , \sum_{i=1}^{N} x_{i,j} = 1$$
[4]

Energy balance equation is as follows:

$$L_{j+1}H_{Lj+1} + V_{j-1}H_{Vj-1} + F_jH_{Fj} - L_jH_{Lj} - V_jH_{Vj} - Q_j = 0$$
[5]

Real enthalpy of components is calculated by combination of ideal gas enthalpy and residual enthalpy of gases and liquids. Ideal gas enthalpy is evaluated by:

$$H^{ig}{}_{i} = a' + b'T + c'T^{2} + d'T^{3} + e'T^{4} + f'T^{5}$$
[6]

where a',b',c',d',e', and f' could be found in literature (Prausnitz, J.M., 1999). General form of gas and liquid residual enthalpy are as follows:

$$\frac{H-H^{ig}}{RT} = \frac{1}{RT} \int_0^P \left[\forall -T(\frac{\partial \forall}{\partial T})_P \right] dP$$
[7]

$$\left(\frac{\partial Lnf_i^l}{\partial T}\right)_{P,\chi} = -\frac{\overline{H}_i - H_i^+}{RT^2}$$
[8]

where \overline{H}_i and H_i^+ are fractional molar enthalpy of component *i* in liquid phase and ideal gas state respectively. With regards to the type of applied equilibrium relations, liquid and gas residual enthalpy could be extracted by equations of state and activity model relations. Peng-Robinson equation is the most common equation of state applied for hydrocarbons (Kasiri.N. and Hormozdi.Sh., 2005). Wilson activity model is utilized for prediction of liquid phase treatment. The following equation could be applied to evaluate fugacity coefficient of gas phase (φ_i^v) from PR equation state and activity coefficient (γ_i)from Wilson activity model:

$$\gamma_i \varphi_i^v P = x_i \gamma_i P_i^{sat} \varphi_i^{sat}$$
[9]

For the purpose of analysis, thermodynamic behavior of the dehydration system was represented by means of Peng-Robinson equation of state. The models guarantee good phase equilibrium predictions over a wide range of temperature and pressure (Polak, L., 2009).

The Peng-Robinson equation is written as:

$$P = \frac{RT}{(V-b)} - \frac{a.\propto}{V(V+b) + b(V-b)}$$
[10]

where

$$a = 0.45724 \frac{R^2 T_c^2}{P_c}$$

$$b = 0.07780 \frac{RT_c}{P_c}$$

$$P = \text{pressure}$$

$$V = \text{molar volume}$$

$$T = \text{temperature}$$

$$R = \text{universal gas constant (8.314 \text{ J.K}^{-1}.\text{mol}^{-1})}$$

$$T_c = \text{critical temperature}$$

$$P_c = \text{critical pressure}$$

$$\alpha = \text{alpha function, function of reduced temperature } T_r = T/T_c$$
$$= [1 + (0.37464 + 1.54226\omega - 0.26992\omega^2)(1 - (\frac{T}{T_c})^{0.5})]^2$$
$$\omega = -\log\left(\frac{P}{P_c}\right) - 1$$

The Peng-Robinson equation is universally used for hydrocarbons and related components over a wide range of temperature and pressure.

CHAPTER 3

MATERIALS AND METHODS

3.0 RESEARCH METHODOLOGY

3.1 INTRODUCTION

This chapter will discuss the research methodology that was used in the study. This project will be carried on based on the three main stages which will be shown in research procedures and the study involves the simulation of parameters performance via ASPEN HYSYS. Many researches have been done and the problems have been identified. In order to solve the problem and upgrading the recent technology in dehydration, since there are rooms to be improved. This research entitled as "Evaluation of The Effectiveness Parameters on Gas Dehydration Plant" and aims to overcome and accomplish to the objectives of Chapter 1 by using Aspen Hysys simulation method.

3.2 RESEARCH DESIGN

3.2.1 HYSYS Simulation Approach of Dehydration

ASPEN HYSYS is usually used for modeling of a dehydration process. It was used for steady state simulation, design, performance monitoring and optimization of oil and gas production, gas processing and petroleum refining industries (HYSYS User Guide, 2005). HYSYS simulator is the main tool in this analysis. Gas composition and some other relevant assumptions are the main inputs for HYSYS.

In HYSYS, all necessary information pertaining to pure component flash and physical property calculations is contained within the Fluid Package, therefore choosing the right Fluid Package is essential. For the given composition of natural gas flowing through the dehydration unit, different Fluid Packages were checked, and Peng-Robinson equation of state was chosen, as an ideal model for process calculations (HYSYS User Guide, 2005) because it is suitable to handle system containing hydrocarbon, and water.

Material streams were used to simulate the material traveling in and out of the simulation boundaries and passing between unit operations. For the material stream, the main properties and composition has to be defined. The parameters necessary are the temperature, pressure, molar flow, and composition. Energy streams are used to simulate the energy travelling in and out and it allows user to define stream parameters, view objects to which the stream is attached and specify dynamic information. The main parameter for energy streams is heat flow (HYSYS User Guide, 2005).

Separator is a unit to separate feed into constituent parts, with one or multiple feeds, one vapor and one liquid product stream. Every separator may be provided with some common features and the user can choose between various heater types, which determine the way in which heat is transferred to the vessel operation (HYSYS User Guide, 2005).

3.2.2 Details on Feed Natural Gas

Natural Gas from crude oil named MASILA was proposed in this research. Composition of gas is needed to analyze any gas dehydration process because gas properties are highly influenced by the composition of gas. Mole fraction in percent of typical gas fields mixed natural gas is given as below:

| Component | Mole % |
|------------|--------|
| Azote N2 | 4.31 |
| CO2 | 0.51 |
| Methane | 84.62 |
| Ethane | 5.65 |
| Propane | 2.93 |
| Iso-Butane | 0.61 |
| Butane | 0.63 |
| Iso-C5 | 0.20 |
| C5 | 0.19 |
| C6+ | 0.15 |
| Total | 100.00 |
| Mol. Wt | 19.24 |
| 1000 T/y | 4.618 |

Table 3.1: Table of gas composition

(Sources: Adapted from GasCities, www.docstoc.com)

3.2.3 Liquid Desiccant Dehydration Unit



Figure 3.1: Typical glycol dehydration unit

Figure above shows process flow diagram of Glycol dehydration unit. The glycol solution enters at the top of the absorber tower and absorbs water as it progresses toward the bottom of the column. Dry gas exits at the top of the contactor and may be used for cooling the incoming lean glycol. The rich stream flows to a separator where gaseous hydrocarbons that were absorbed along with some of the water in the contactor are liberated. Then, the glycol flows to the stripper where it is regenerated by boiling off the water and returned to the contactor. For processes requiring gas with very low water dew points, a stripping vapor will most likely be needed to aid the regeneration process.

An investigation of required parameters for dehydration units design will be present. Effectiveness parameters such as trays number of absorber, absorber's pressure and natural gas flow rate are studied to obtain optimum condition. In addition to the design parameters, there are other factors that influence the residual water content of the sales gas. First, the temperature of the inlet gas will dictate the total amount of water fed to the unit. Lower plant temperatures will require less water to be removed by the glycol. Second, lean glycol temperature at the top of the contactor will affect the water partial pressure at the top stage as high glycol temperatures will result in high water content in the overhead gas.

3.2.4 Details on TEG Dehydration Unit

Triethylene glycol has been used to dehydrate sweet and sour natural gas over a wide range of operating conditions. The system sizing involves specifying the minimum glycol mass flow, specific glycol circulation rate, and number of theoretical stages. The following table shows the TEG Dehydration Unit specification, which it is used for simulation and modelling of dehydration unit.



Figure 3.3: TEG Absorption Column

| Specifications | Natural Gas, in | TEG,in |
|----------------------------------|---------------------------|---------------------------|
| Vapor Fraction | 1 | 0 |
| Temperature (°C) | 35 | 35 |
| Pressure (kPa) | 2000,4000,6000,8000,10000 | 2000,4000,6000,8000,10000 |
| Gas Flow $(\frac{m^3}{h (gas)})$ | 1000,2000,3000,4000,5000 | 1000,2000,3000,4000,5000 |
| Number of Theoretical | 2,4,6,8 | 2,4,6,8 |
| Stages of Absorber | | |

 Table 3.2: Inlet stream used in TEG Dehydration Unit

3.3 RESEARCH PROCEDURES



Figure 3.4: Project Flow Chart

3.4 DETAILED SIMULATION STEPS

To start a simulation case, from the file menu, select New and then Case. Click the New Case icon in the toolbar. The Simulation Basis manager appears:

| omponent Lists - Component List - | 1 | Databank Selection HYSYS Databa Aspen Propertie | nks is | | | | |
|--------------------------------------|------------|---|-------------|----------------|----------------|------------------|--|
| | L | Delete | | | | | |
| | | Сору | Imp | ort | | | |
| | | ⊻iew | Exp | iort | | | |
| | | <u>R</u> efresh | Re-i | mp <u>o</u> rt | | | |
| Componente | Eluid Dkao | Hupothatiasla | Oil Managor | Beactions | Component Maps | Liser Properties | |

Figure 3.5: Starting a new case

Then, from the tools menu, select Preferences. The Session Preferences property view appears. The HYSYS default session settings were save in a Preference file called HYSYS.pfr.



Figure 3.6: Setting HYSYS in a Preference file

To create a new unit set, select Variables tab in the session preferences property view and select the Units page if it is not already selected. In Available Unit Sets group, select SI to make it the active set. Click the Clone button and a new unit set named NewUser appears.

| variables | ~~~ | | Jets | | | | |
|-----------|------|---------------------|------------|----------|---------------|-----|--------------------|
| Units | Eur | oSI d | | | | | Clone |
| Formats | SI | | | | | | Delete |
| | Uni | t Set Na <u>m</u> e | e SI | | | | View <u>U</u> sers |
| | Disp | lay Units | | 111.5 | | | |
| | Aci | litu | | Unit | ma KOH/ | | <u></u> iew |
| | Act | Gas Flow | 1 | - | ACT m3/ | | Add |
| | Act | Vol. Flow | | 1 | | n | |
| | Act | ual Liquid F | Flow | | m3/ | s | Delete |
| | Act | Actual Mass Density | | | kg/m3 | 3 | AI |
| | Ang | jle – | | | deg | 3 | Z+ |
| | API | Fire Equal | tion Consl | tan | Btu/hr-ft1.64 | 1 | |
| | Are | а | | <u> </u> | m. | 2 | |
| | Are | aPerVolum | e | | m2/m3 | 3 🗾 | |
| | | | r | | 1 | | 1 11 |

Figure 3.7: Creating new unit set

In the Display Unit group, select the unit of the specification.

| Variables | Ava | ilable Unit | Sets | | | | |
|-----------|------|---------------------|----------|-----------|--------------|----------|--------------------|
| Inits | Eur | oSI Id | | | | | Clone |
| Formats | Ne | wUser | | | | | Delete |
| | Uni | t Set Na <u>m</u> e | e Ne | wUser | | | View <u>U</u> sers |
| | Disp | ay Units | | | | | |
| | | | | Unit | | - | View |
| | Aci | dity | | | mg KOH/ | 9 | |
| | Act | . Gas Flow | | | ACT_m3/ | h - | A <u>d</u> d |
| | Act | . Vol. Flow | | 6 | m3/ | h | |
| | Act | ual Liquid I | Flow | | m3/s | s | Delete |
| | Act | ual Mass E | ensity | | kg/m | 3 | |
| | Ang | gle | | | de | 9 | 2+ |
| | API | Fire Equa | tion Cor | istan | Btu/hr-ft1.6 | 4 | |
| | Are | a | | | m | 2 | |
| | Are | aPerVolum | e | - E | m2/m | 3 🗾 | |
| | | | Files | Bergungen | Eutomainus | Oillined | Fran Cinima |

Figure 3.8: Selecting Unit used

To add a Fluid Package, click the Fluid Packages tab on the Simulation Basis Manager Property view. Click the Add button, and the property view of the new Fluid Package appears.



Figure 3.9: Adding a Fluid Package

On the Set Up tab, select <none> in the Property Package Selection list and type Peng Robinson. In the Property Package Filter, select the EOSs button, which is Equations of State.

| Kabadi-Danner | ~ | Property Package Filter |
|----------------|---|-------------------------|
| Margules | | All Types |
| MBŴR | | C EOSs |
| NBS Steam | | C Activity Models |
| | | C Chao Seader Models |
| Peng-Robinson | | C Vapour Press Models |
| PR-Twu PRSV | = | C Miscellaneous Types |
| Sour SRK | | |
| Sour PR | | |
| SBK-TIMU | - | Launch Property Wizard |

Figure 3.10: Selecting Property Package

To create a component list in the Fluid Package property view, select Component List-1 in the Set Up tab.

| Add Component | Selected Lomponents | | -Lomponents Avai | ilable in the | Component Library | | |
|----------------------------|--|--------------------------|------------------------------------|----------------------|--------------------|----------------------------|---|
| Components Traditional | | | <u>M</u> atch | | | View Filters | |
| Hypothetical | | | C Sim Name | (● Fu | ill Name / Synonym | C Formula | |
| | | < <u>A</u> dd Pure | Methane Ethane Propane | C1 C2 C3 | | CH4 C2H6 C3H8 | Ê |
| | | <- <u>S</u> ubstitute-> | i-Butane n-Butane i-Pentane | i-C4 n-C4 i-C5 | | C4H10 C4H10 C5H12 | |
| | | <u>H</u> emove> | n-Pentane n-Hexane n-Heptane | n-C5 C6 C7 | | C5H12 C6H14 C7H16 | |
| | iPentane i-C5 C5H12 n-Pentane n-C5 C5H12 n-Hexane C6 C6H14 n-Hexane C7 C7H16 Sort List n-Octane C8 C8H18 n-Decane C1 C1H22 n-Dctane C1 C1H22 | C8H18 C9H20 C10H22 | | | | | |
| | | View Component | n-C11 n-C12 n-C13 | C11 C12 C13 | | C11H24 C12H26 C13H28 | - |
| | | | Show Synony | yms | Cluster | | |
| Selected Compor | nent by Type | | | | | | |

Figure 3.11: Selecting the component

| < <u>A</u> dd Pure |
|-------------------------|
| <- <u>S</u> ubstitute-> |
| |
| <u>R</u> emove> |
| S <u>o</u> rt List |
| ⊻iew Component |
| |

Figure 3.12: Completed component list

Close the Fluid Package view to return to the Simulation Basis Manager Property view.

| urrent Fluid Packages | | Flowsheet - Fluid Pkg Ass | sociatio <u>n</u> s | |
|----------------------------------|-------------------|---|---------------------|---|
| Basis-1 NC: 12 PP: Peng-Robinson | ⊻iew | Flowsheet | Fluid Pkg To Use | |
| | | Case (Main) | Basis-1 | |
| | Delete | | | |
| | Сору | | | |
| | lucast 1 | Default Fluid Pkg | Basis-1 | • |
| | Import | -Fluid Pkg for New Sub-Fl | owSheets | |
| | Export | Use Default Fluid P Use Parent's Fluid F | kg ⊃kg | |
| Components Fluid Pkgs Hypothetic | als 🚽 Oil Manager | Reactions Component Ma | aps User Properties | |

Figure 3.13: Back to Simulation Basis Manager

Next, enter the simulation environment. A PFD view will appear. Save the case.



Figure 3.14: Entering the Simulation Environment.

Then, click the Workbook icon to access the Workbook property view. Feed steam is installed. Data required, included the properties and compositions, and are entered.

| Worksheet | Stream Name | Wet Gas | | |
|---|-------------------------------|-----------------|---|--|
| Conditions | Vapour / Phase Fraction | 0.9990 | | |
| Conditions | Temperature [C] | 35.00 | | |
| Folgentes Composition K Value User Variables Notes Cost Parameters | Pressure [kPa] | 8000 | | |
| | Molar Flow [kgmole/h] | 1.774e+004 | | |
| | Mass Flow [kg/h] | 3.399e+005 | | |
| | Std Ideal Lig Vol Flow [m3/h] | 1000 | | |
| | Molar Enthalpy [kJ/kgmole] | -7.754e+004 | | |
| | Molar Entropy [kJ/kgmole-C] | 147.7 | | |
| | Heat Flow [kJ/h] -1.375e+009 | | | |
| | Liq Vol Flow @Std Cond [m3/h] | <empty></empty> | | |
| | Fluid Package | Basis-1 | | |
| | Utility Type | | | |
| | | |) | |
| Worksheet Att | achments Dynamics | | | |

Figure 3.15: Entering Data on Feed

Then, install the necessary unit operations for processing the gas. Click on the Workbook icon to reach the Workbook property view and click on the Unit Ops tab. To add a unit, click Add UnitOps button. The UnitOps property view appears and select unit needed.

| Categories | Available Unit Operations | Add |
|---|---|--------|
| All <u>Unit</u> Ops Vesels Heat Transfer Equipment Piping Equipment Solids Handling Reactors Prebuilt Columns Sub-Flowsheets Logicals Extensions User Ops Electrolyte Equipment Refinetry Ops Ustream Ops | 3 Phase Separator 3 Stripper Crude 4 Stripper Crude 4 Stripper Crude 4 Stripper Crude Absorber ACM Oper Adjust Aspen Hydraulics Sub-Flow Balance Black Oil Translator Boolean And Boolean CountDown Boolean Latch Boolean OffDly Boolean OffDly Boolean OffDly Boolean OffD | Gancel |

Figure 3.16: Selecting Unit Operations

Next, label the inlet and outlet stream and enter the details in the input expert property. Firstly, an absorber is added to the system.

| Design | Column Name Absorber | Sub-Flowsheet Tag COL1 | |
|--|---|--|--|
| Connections Monitor Specs Summary Subcooling Notes | Top Stage Injet TEG.in Optional Inlet Streams Stream Inlet Stage I << Stream >: Bottom Stage Inlet Wet Gas Stage Numbering © Top Down © Bottom Up | 1 Num of Stage: n = P \$000 kPa Pn \$000 kPa | O <u>v</u> hd Vapour Outlet Dry Gas Optional Side Drgws Stream Type Oraw Stage Stream Stream |
| Dening Russ | Edit Trays | hant Destaurance I Elemetreet | |

Figure 3.17: Creating the inlet and outlet stream of absorber

Then, insert a reboiler and enter the details as follow:

| Design | | | |
|----------------|-----------------------|-------------------|---|
| Connections | | | |
| Parameters | Delta <u>P</u> | Duty | - |
| User Variables | 0.0000 kPa | 6.50000e+006 kJ/h | |
| Notes | G | × | |
| | _ (| | |
| | - | ~ | |
| | | | |
| | | | |
| | | | |
| | | | |
| Design Rating | Worksheet Performance | Dynamics | |

Figure 3.18: Entering data of reboiler

After absorption, soluble gases such as CO_2 are first released from the glycol in a flash tank. So, a separator is added to the system.

| Design | Name Flash Drum | |
|--|--|-------------------------|
| Connections Parameters User Variables | Injets Reboiled Control Contro | Vapour Outlet Flash Gas |
| Notes | | |
| | Energy (Optional) | ļ |
| | | |

Figure 3.19: Connect streams for separator

Glycol is needed to be heated before it flows to a distillation column. So, install the gas exchanger by access the Object Palette. The Heat Exchanger property view appears. The inlet and outlet stream are attached and parameters for example pressure drop and number of shells in Series and Parallel, and Tube Passes per Shell need to be entered.



Figure 3.20: Entering the connections and data required

In general, HYSYS has a number of pre-built column templates that can be installed and customized by changing attached stream names, number of stages, and default specifications. From the Simulation tab, install distillation column and enter the details in the input expert property as shown below. The regeneration column has a reboiler and condenser. Regeneration of the glycol by atmospheric distillation at about 200°C achieves about 1 weight % water in regenerated glycol. The regenerated glycol is pumped back to the absorption column and heat exchanged against the rich glycol and cooling water.

| Design | Column <u>N</u> ame T-101 | Sub-Flowsheet Tag COL2 | |
|---|--|----------------------------|---|
| Connections Monitor Specs Specs Summary Subcooling | Condenser Energy Stream | | Delta P Steam _ |
| Notes | Injet Streams Stream Injet Stage Rich TEG,in 2_Mair << Stream >> | | Optional Side Draws Stream Type Draw Stage << Stream >> |
| | Stage Numbering Top Down C Bottom Up Edit Trays | | ta P Bottoms Liguid Outlet |
| Design Param | eters Side Ops Rating Worksh | neet Performance Flowsheet | Reactions Dynamics |
| Delete 1 | Column Environment Bu | n Beset | nverged III Indate Outlete III Japanes |

Figure 3.21: Entering data of distillation column

A pump is added after regeneration column. Its duty and adiabatic efficiency is entered.

| Desi | gn | Ľ | <u>N</u> ar | me P-100 | | | |
|---|--------|------------------|-----------------|----------|---------------------------------------|----------|---------|
| Connection Parameters Curves Links | ons | Injet Pump.in | <u>-</u> | | Outlet Pump,out | • | |
| Notes | 5163 | | Energy Q-103 | |) Fluid <u>P</u> ackage Basis-1 | <u> </u> | |
| _ Design | Rating | Worksheet | Performance | Dynamics | | | |
| Delete | | | | IOK - | | 🗖 🔽 On 🗖 | lanored |

Figure 3.22: Entering data of pump

Before glycol recycles back to the absorber, a recycle block is used to control that the recycled glycol flow equals the glycol flow to the absorption column.

| Worksheet | Name | Cool TEG | TEG,in |
|-------------|--------------------------------|----------------|-------------|
| HOIKSHOOK | Vapour | 0.0000 | 0.0000 |
| ditions | Temperature [C] | 90.00 | 90.00 |
| erties | Pressure [kPa] | 8041 | 8041 |
| | Molar Flow [kgmole/h] | 1.456e+006 | 1.455e+006 |
| position | Mass Flow [kg/h] | 2.623e+007 | 2.621e+007 |
| | Std Ideal Lig Vol Flow [m3/h] | 2.628e+004 | 2.626e+004 |
| | Molar Enthalpy [kJ/kgmole] | -2.810e+005 | -2.810e+005 |
| | Molar Entropy [kJ/kgmole-C] | 68.95 | 68.95 |
| | Heat Flow [kJ/h] | -4.091e+011 | -4.088e+011 |
| | | - | |
| Connections | Parameters Worksheet Monitor L | Iser Variables | |
| Delete | Continue | | |

Figure 3.23: Connect stream in and stream out of the recycle block

Lastly, after all the unit operations are set up, click the run button to begin calculations and make sure the whole system is converged. The information displayed on the Monitor page is updated with each iteration. Repeat the simulation by manipulating different operating pressure, gas flow rate, and different desiccant Percentage of water removal is gained.

3.5 CONCLUSION

This chapter discussed about effective parameters that we used for optimization of dehydration conditions of natural gas. The parameters, for example pressure, number of trays, and gas flow rate are studied.

CHAPTER 4

RESULTS AND DISCUSSIONS

4.1 INTRODUCTION

This chapter presents all the results obtained with appropriate parameters that involved in the simulation by using the method described in Chapter 3. In achieving the objectives of this study, survey on the best software for natural gas dehydration simulation has been done, which are Aspen Hysys, Chemcad, and Pro II. Aspen Hysys is selected because of its thermodynamic package for dehydration process and also its user-friendly properties. The main objective is to determine the most effective parameters in Triethylene Glycol Dehydration process. The screening processes of effect of number of theoretical stages, natural gas volumetric flow rate, and absorber pressure on percentage of water removal were done. The raw data for all runs are shown in Appendix.

4.2 ANALYSIS OF PERCENTAGE OF WATER REMOVAL VERSUS GAS FLOW RATE AT VARIOUS PRESSURE AND NUMBER OF THEORETICAL STAGES IN ABSORBER

In order to calculate the percentage of water removal in each run, the molar flow of water in wet gas stream and dry gas stream are obtained. Below are equation used for the calculation of percentage of water removal in each run with different parameters:

Percentage of Water Removal (%)

 $=\frac{Wet \ Gas \ Molar \ Flow-Dry \ Gas \ Molar \ Flow}{Wet \ Gas \ Molar \ Flow} \ x \ 100\%$

The results of the calculations are shown in the Table and Graph below:







Figure 4.2: Percentage of Water Removal versus Gas Flow Rate at 4 Theoretical Stages in Absorber



Figure 4.3: Percentage of Water Removal versus Gas Flow Rate at 6 Theoretical Stages in Absorber



Figure 4.4: Percentage of Water Removal versus Gas Flow Rate at 8 Theoretical Stages in Absorber

4.3 DISCUSSIONS

Figures 4.1, 4.2, 4.3 and 4.4 were drawn based on Table 4.1, 4.2, 4.3 and 4.4 respectively. Figure 4.1, 4.2, 4.3 and 4.4 showed the changes of percentage of water removal at various gas flow rates, pressure, and number of theoretical stages. From these figures, the average water removal efficiency with natural gas flow of 1000m³/h was the highest and then followed by natural gas flow of 2000m³/h, 3000m³/h, 4000m³/h, and 5000m³/h. The percentage of water removal decreased when the gas flow rate was increased. This attributed by the contact between constant glycol and gas flow was reducing as the gas flow was increasing. Thus, less water molecules dissolved in Triethylene Glycol (TEG). These results can be compared by those results which with decreasing TEG flow rate.

According to Kasiri, N. and Hormozdi, Sh. (2005), increasing TEG flow rate increased water absorption rate. Also, with reference to Mohamadbeigy,K. (2008), higher TEG flow rate increased water dehydration efficiency. In the other word, decreasing TEG flow rate decreased water absorption rate. In present work, natural gas flow was increased and TEG flow rate kept constant. The condition can be said was the same with decreasing TEG flow rate and constant natural gas flow. Thus, the statement from N. Kasiri, Hormozdi, Sh. (2005) and Mohamadbeigy,K. (2008) proved the present result.

Results of above figures shown that, at constant flow rate, the percentage of water removal decreased with increasing pressure. With increasing pressure, there were less contact between natural gas and TEG. Therefore, less amount of water content was absorbed and removed by TEG. From Figure 4.1, the result showed that pressure of 2000kPa had highest dehydration efficiency, followed by 6000kPa, 4000kPa, 8000kPa and 10000kPa. From Figure 4.2, the dehydration efficiency was also highest with pressure of 2000kPa, and then followed by 4000kPa, 6000kPa, 8000kPa, and 10000kPa. There was same condition with the results shown at Figure 4.3. However, from Figure 4.4, which was at column of 8 theoretical stages, pressure with 4000kPa showed the highest dehydration efficiency, followed by 2000kPa, 8000kPa and 10000kPa. In this case, pressure of 2000kPa was considered as the optimum pressure for dehydration.

As mentioned by Mohamadbeigy.K. (2008), the water content of the inlet gas decreases with increasing pressure and thus less water must be removed if the gas was dehydrated at a higher pressure. It can be explained by at higher pressure, less water will be removed. At lower pressure, less wall thickness of absorber was required. Thus, consideration of economic existed between operating pressure and absorber cost. The statement proved the present results.

In present work, the percentage of water removal increased by number of theoretical stages in absorber. At constant pressure, the dehydration efficiency was nearly close among all number of theoretical stages except for theoretical stages of 2. Its dehydration efficiency was slightly lower. There were obvious observations that the efficiency of water removal were increasing by comparing Figures 4.1 and 4.2, which were of 2 theoretical stages and 4 theoretical stages respectively. When we further increased the number of theoretical stages by looking at Figures 4.3 and 4.4, which were of 6 theoretical stages and 8 theoretical stages, the efficiency of water removal were further increased.

According to Erik L. (2002), increasing number of theoretical stages will only reduce the water content slightly. This has been proved by comparison of figures above. As Mohamadbeigy.K. (2008) carried out his studies; increasing number of trays will increase the percentage of water removal and at the same time allow the gas to approach equilibrium with the lean glycol at a lower glycol circulation rate. As mentioned by Pezhman K. and Roya H. (2011), in reboiler temperature of 195°C and 204°C, three and four equilibrium stages are sufficient to reach 96% absorption efficiency. So it can conclude that absorber with theoretical stages of four was the optimum number of theoretical stages that should be used since the difference of water removal of 4, 6, and 8 theoretical stages are very small.

Absorption is a method that removes water from the gas stream by counter current contact with Triethylene Glycol in a tray type contactor tower. The feed gas must be cleaned to remove all liquid water, heavy hydrocarbon, wax, and other impurities. These impurities can be removed by using a separator. Natural Gas entered the unit at the bottom of the absorber and rised through the tower where it contacted with the TEG solution flowing downward across the trays. Natural gas transferred its water vapour to TEG through the contact. Dried gas exited from the top of the contactor. The gas required water concentration specification. The water rich TEG was removed from the bottom of the absorber and passed through the reflux condenser coil and flashed off most of the soluble gas in the flash tank. The water-rich TEG flowed through the rich-lean heat exchanger to the regenerator. TEG circulated in a closed system, where the water was boiled from the TEG at atmospheric pressure. The temperature of regeneration process was under 204°C because TEG can degrade at temperature above 204°C. The regenerated TEG was then recirculating to the contacting tower. The reflux ratio of the column used was 0.5 and the overhead vaporization rate was 100kmol/h.

In Aspen Hysys, Peng-Robinson (PR) equation of state has been used. This fluid package is simple and has only one adjustable binary parameter for each binary component pair. According to Erik L., this is not enough to fit vapour/liquid equilibrium data accurately. An activity model with parameters fitted to new TEG/Water equilibrium data was more accurate. This is important in absorber design calculations but for flow sheet development and comparison, Peng-Robinson equation is satisfactory. The actual loss of glycol from a dehydration plant is due to mechanical losses from the absorption column, the flash tank or the regeneration column.

During the running of simulation, convergence problems were happened. The main problems mainly occurred at stripper and recycle block. At first, the stripper cannot be converged because of too high operating pressure that has been inserted. This problem can be simply solved by using atmospheric pressure in the system. The problem of recycle block convergence was because there was excessive water content in lean glycol or means by under circulation of glycol. This resulted in poor dehydration. However, this can be solved by adjusting the overhead reflux ratio in stripper.

CHAPTER 5

CONCLUSIONS AND RECOMMENDATIONS

5.1 CONCLUSIONS

The project is an evaluation on effectiveness parameters on natural gas dehydration plant using liquid desiccant which is triethylene glycol. This present study determined the optimum parameters used in liquid desiccant dehydration unit. Studies have been done on absorption process which play important role in dehydrating natural gas. In design of a natural gas dehydration unit, the effectiveness parameters such as number of theoretical stages in absorber, natural gas flow rate, and operating pressure were investigated with economic considerations.

The process simulation is a useful tool which can lead to determine optimum conditions for maximum efficiency. In this study, the selection for suitable software with considering of specific thermodynamic package for dehydration process has carried out. Aspen Hysys was selected as the simulator while Peng-Robinson was selected as fluid package. The results showed that increasing gas flow rate and operating pressure decreases dehydration efficiency which is in contrast with the behaviour of process as number of theoretical stages increased. In natural gas dehydration unit, water absorption rate decreases with the increasing gas flow rate but increases with increasing number of equilibrium stages. As pressure increases, water absorption rate will decrease because of less contact between natural gas and triethylene glycol. Absorber with 1000m³/h gas flow rate, 2000kPa and number of theoretical stages of 4 was considered optimum in this study.

5.2 **RECOMMENDATIONS**

- Consider the effect of high carbon dioxide composition in the feed. High quantities of CO₂ in the feed can accelerate corrosion in the regenerator.
- Consider the emission of aromatic (BTEX) and other volatile organic compounds, VOC's from the regenerator and flash drum. Most plant feed contained small quantities of aromatic hydrocarbons that are quite soluble in TEG, which are primarily comprised of benzene, toluene, ethylbenzene, or xylenes and will be carried to the flash tank where small fraction is released along with other volatile organic compounds. The remaining VOC's and aromatics will travel to regenerator and removed as volatile gases. This may cause serious environmental impact (Michael W.H., et. al. (1993).
- ↔ Other parameters such as TEG flow rate and temperature can be studied.

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APPENDIX A

SIMULATION DATA

APPENDIX A.1

BASE DEHYDRATION UNIT OPERATING PARAMETERS

| Inlet gas temperature | 35°C |
|--------------------------------|-----------------------|
| Inlet gas pressure | 6000kPa |
| Inlet gas flow | 5000m ³ /h |
| Inlet gas composition | |
| Methane | 0.8462 |
| Ethane | 0.0565 |
| Propane | 0.0293 |
| i-Butane | 0.0061 |
| n-Butane | 0.0063 |
| i-Pentane | 0.0020 |
| n-Pentane | 0.0019 |
| n-Hexane | 0.0015 |
| Nitrogen | 0.0431 |
| CO_2 | 0.0051 |
| H_2O | 0.0020 |
| Lean glycol temperature | 107°C |
| Rich glycol flash pressure | 4000kPa |
| Regenerator pressure | 110kPa |
| Equilibrium trays in contactor | 6 |
| | |

APPENDIX A.2

SIMULATION DATA WITH DIFFERENT CONDITIONS

| | | Percentage of Water Removal | | | | | | |
|-----------------------------------|---------|-----------------------------|---------|---------|----------|--|--|--|
| Gas Flow Rate (m ³ /h) | 2000kPa | 4000kPa | 6000kPa | 8000kPa | 10000kPa | | | |
| 1000 | 98.7 | 99.45 | 99.65 | 99.85 | 99.8 | | | |
| 2000 | 99.7 | 99.7 | 99.9 | 83.7 | 76.3 | | | |
| 3000 | 83.2 | 77.25 | 80.15 | 70.65 | 62.9 | | | |
| 4000 | 83.2 | 70.8 | 73.85 | 60.75 | 55.15 | | | |
| 5000 | 79.8 | 60 | 63.45 | 50.95 | 46.5 | | | |

| Table 4.1: Analysis of Water Removal on Gas Flow Rates and Pressures | at 2 |
|--|------|
| Theoretical Stages in Absorber | |

| | Pressure | | | | | | |
|-----------------------------------|----------|---------|---------|---------|----------|--|--|
| Gas Flow Rate (m ³ /h) | 2000kPa | 4000kPa | 6000kPa | 8000kPa | 10000kPa | | |
| 1000 | 99.95 | 100 | 100 | 100 | 100 | | |
| 2000 | 100 | 99.95 | 100 | 85.25 | 79.35 | | |
| 3000 | 88.2 | 82.65 | 89.85 | 70.7 | 63 | | |
| 4000 | 82.05 | 87.2 | 73.15 | 60.8 | 55.25 | | |
| 5000 | 75.85 | 75.65 | 63.1 | 51 | 46.6 | | |

Table 4.2: Analysis of Water Removal on Gas Flow Rates and Pressures at 4Theoretical Stages in Absorber

| | | Pressure | | | | | | |
|-----------------------------------|---------|----------|---------|---------|----------|--|--|--|
| Gas Flow Rate (m ³ /h) | 2000kPa | 4000kPa | 6000kPa | 8000kPa | 10000kPa | | | |
| 1000 | 100 | 100 | 100 | 100 | 100 | | | |
| 2000 | 100 | 100 | 100 | 87.1 | 79.35 | | | |
| 3000 | 87.8 | 83.5 | 86.8 | 70.7 | 63.05 | | | |
| 4000 | 81.9 | 87.6 | 74.75 | 60.8 | 55.4 | | | |
| 5000 | 77.6 | 75 | 63.05 | 51 | 46.6 | | | |

Table 4.3: Analysis of Water Removal on Gas Flow Rates and Pressures at 6Theoretical Stages in Absorber

| | Pressure | | | | | |
|-----------------------------------|----------|---------|---------|-----------------|------|--|
| Gas Flow Rate (m ³ /h) | 2000kPa | 4000kPa | 6000kPa | 0kPa 8000kPa 10 | | |
| 1000 | 99.95 | 100 | 100 | 100 | 100 | |
| 2000 | 99.95 | 100 | 100 | 88.1 | 79.9 | |
| 3000 | 87.6 | 92.85 | 89.15 | 70.75 | 63.2 | |
| 4000 | 76.15 | 88.2 | 74.3 | 60.85 | 55.4 | |
| 5000 | 77.6 | 75.2 | 62.95 | 51 | 46.6 | |

Table 4.4: Analysis of Water Removal on Gas Flow Rates and Pressures at 8Theoretical Stages in Absorber

APPENDIX B

EXAMPLE OF CALCULATIONS

APPENDIX B.1 CALCULATION FOR WATER DEHYDRATION EFFICIENCY

For simulation run at 4000kPa, 5000m³/h gas flow rate and 6 theoretical stages,

Percentage of Water Removal (%) = $\frac{Wet \ Gas \ Molar \ Flow - Dry \ Gas \ Molar \ Flow}{Wet \ Gas \ Molar \ Flow} \ x \ 100\%$

 $=\frac{177.38-44.28}{177.38} \ x \ 100\%$ = 75%

✤ All other simulation runs are calculated using the same method and equations and the results are shown in Tables 4.1, 4.2, 4.3, and 4.4.

APPENDIX C

EXAMPLE OF SIMULATION WORKBOOK (4000KPa, 5000m³/h, 6 theoretical stages)

APPENDIX C.1 SIMULATION WORKBOOK

| 1 | | Case Name: | Case Name: RUN5-4000KPA,35C,5000M3H,6TS.HSC | | | | | | |
|----------------------------|----------------------------|-------------------|---|---------------------------------|-------------|-------------|--|--|--|
| 3 | asnen Burlingtor | Unit Set: NewUser | | | | | | | |
| 4 | USA | | Date/Time: Thu Oct 27 19:17:04 2011 | | | | | | |
| 6 7 8 | Workbook: Case (Main) | | | | | | | | |
| 9 | gent bauts | e (Brisedar | Material Stream | Material Streams Fluid Pkg: All | | | | | |
| 11 | Name | Wet Gas | Dry Gas | Rich TEG | Reboiled | Flash Gas | | | |
| 12 | Vapour Fraction | 0.9990 | 1.0000 | 0.0000 | 0.0000 | 1.0000 | | | |
| 13 | Temperature (C) | 35.00 | 18.75 | 16.32 | 50.14 | 50.14 | | | |
| 14 | Pressure (kPa) | 8000 | 4000 | 4000 | 4000 | 4000 | | | |
| 15 | Molar Flow (kgmole/h) | 8.869e+004 | 8.855e+004 | 6 408 -+ 004 | 6 4080+004 | 0.0000 | | | |
| 10 | Liquid Volume Flow (Kg/n) | 5000 | 4997 | 58.64 | 58 64 | 0.0000 | | | |
| 18 | Heat Flow (k.l/b) | -6.877e+009 | -6.828e+009 | -4.978e+008 | -4.913e+008 | 0.0000 | | | |
| 19 | Name | Flash out | Rich TEG,in | Lean TEG,out | Pump,in | Steam | | | |
| 20 | Vapour Fraction | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.9997 | | | |
| 21 | Temperature (C) | 50.14 | 75.00 | 106.9 | 81.54 | 101.4 | | | |
| 22 | Pressure (kPa) | 4000 | 4000 | 110.0 | 110.0 | 110.0 | | | |
| 23 | Molar Flow (kgmole/h) | 1142 | 1142 | 1002 | 1002 | 140.0 | | | |
| 24 | Mass Flow (kg/h) | 6.408e+004 | 6.408e+004 | 6.151e+004 | 6.151e+004 | 2574 | | | |
| 25 | Liquid Volume Flow (m3/h) | 58.64 | 58.64 | 55.91 | 55.91 | 2.732 | | | |
| 26 | Heat Flow (kJ/h) | -4.913e+008 | -4.864e+008 | -4.420e+008 | -4.4/0e+008 | -3.264e+007 | | | |
| 27 | Name | Pump,out | Hot Dry Gas | 0.0000 | 0.0000 | | | | |
| 28 | Vapour Fraction | 78.58 | 19.61 | 0.000 | 60.00 | | | | |
| 30 | Pressure (kPa) | 2.352e+004 | 4000 | 2.352e+004 | 2.352e+004 | | | | |
| 31 | Molar Flow (kgmole/h) | 1002 | 8.855e+004 | 1002 | 1002 | | | | |
| 32 | Mass Flow (kg/h) | 6.151e+004 | 1.697e+006 | 6.151e+004 | 6.151e+004 | | | | |
| 33 | Liquid Volume Flow (m3/h) | 55.91 | 4997 | 55.91 | 55.91 | | | | |
| 34 | Heat Flow (kJ/h) | -4.452e+008 | -6.824e+009 | -4.486e+008 | -4.486e+008 | | | | |
| 35 | | | Compositions | | Fluid Pk | g: All | | | |
| 37 | Name | Wet Gas | Dry Gas | Rich TEG | Reboiled | Flash Gas | | | |
| 38 | Comp Mole Frac (Methane) | 0.8462 | 0.8475 | 0.0024 | 0.0024 | 0.8115 | | | |
| 39 | Comp Mole Frac (Ethane) | 0.0565 | 0.0566 | 0.0003 | 0.0003 | 0.0566 | | | |
| 40 | Comp Mole Frac (Propane) | 0.0293 | 0.0293 | 0.0002 | 0.0002 | 0.0273 | | | |
| 41 | Comp Mole Frac (i-Butane) | 0.0061 | 0.0061 | 0.0000 | 0.0000 | 0.0042 | | | |
| 42 | Comp Mole Frac (n-Butane) | 0.0063 | 0.0063 | 0.0000 | 0.0000 | 0.0044 | | | |
| 43 | Comp Mole Frac (i-Pentane) | 0.0020 | 0.0020 | 0.0000 | 0.0000 | 0.0011 | | | |
| 44 | Comp Mole Frac (n-Pentane) | 0.0019 | 0.0019 | 0.0000 | 0.0000 | 0.0007 | | | |
| 45 | Comp Mole Frac (Nitrogen) | 0.0013 | 0.0013 | 0.0000 | 0.0000 | 0.0007 | | | |
| 47 | Comp Mole Frac (CO2) | 0.0051 | 0.0051 | 0.0007 | 0.0007 | 0.0106 | | | |
| 48 | Comp Mole Frac (TEGlycol) | 0.0000 | 0.0000 | 0.2880 | 0.2880 | 0.0000 | | | |
| 49 | Comp Mole Frac (H2O) | 0.0020 | 0.0005 | 0.7067 | 0.7067 | 0.0053 | | | |
| 50 | Name | Flash out | Rich TEG,in | Lean TEG,out | Pump,in | Steam | | | |
| 51 | Comp Mole Frac (Methane) | 0.0024 | 0.0024 | 0.0000 | 0.0000 | 0.0197 | | | |
| 52 | Comp Mole Frac (Ethane) | 0.0003 | 0.0003 | 0.0000 | 0.0000 | 0.0026 | | | |
| 53 | Comp Mole Frac (Propane) | 0.0002 | 0.0002 | 0.0000 | 0.0000 | 0.0014 | | | |
| 54 | Comp Mole Frac (i-Butane) | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | | | |
| 55 | Comp Mole Frac (n-Butane) | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0001 | | | |
| 57 | Comp Mole Frac (I-Pentane) | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | | | |
| 58 | Comp Mole Frac (n-Hexane) | 0.0000 | 0.0000 | 0.0000 | 0.0000 | 0.0000 | | | |
| 59 | Comp Mole Frac (Nitrogen) | 0.0017 | 0.0017 | 0.0000 | 0.0000 | 0.0138 | | | |
| 60 | Comp Mole Frac (CO2) | 0.0007 | 0.0007 | 0.0000 | 0.0000 | 0.0061 | | | |
| 61 | Comp Mole Frac (TEGlycol) | 0.2880 | - 0.2880 | 0.3282 | 0.3282 | 0.0003 | | | |
| 62 | Comp Mole Frac (H2O) | 0.7067 | 0.7067 | 0.6718 | 0.6718 | 0.9559 | | | |
| 63 64 65 66 67 | | | | | | | | | |
| 68 69 | | | | | | | | | |

| 1 | | | | | Case Name: RUN5-4000KPA,35C,5000M3H,6TS.HSC | | | | | |
|----------|--|--------------|----------------|-------------------------------------|---|-------------------------|--------|------------|-------------|--|
| 2 | aspentech UNIV MALAYSIA Burlington, MA USA | | | Unit Set: NewUser | | | | | | |
| 4 | | | | Date/Time: Thu Oct 27 19:17:04 2011 | | | | | | |
| 6 | Workl | (continued) | | | | | | | | |
| 9 | | | | Co | mpositions (contir | nued) Fluid Pkg: | | | | |
| 10 | Name | | Pump,out | | Hot Dry Gas | Cool TEG | TEG,in | | | |
| 12 | Comp Mole Frac (Methane) | | 0.00 | 00 | 0.8475 | 0.0000 | | 0.0000 | | |
| 13 | Comp Mole Frac (Ethane) | | 0.00 | 00 | 0.0566 | 0.0000 | | 0.0000 | | |
| 14 | Comp Mole Frac (Propane) | | 0.00 | 00 | 0.0295 | 0.0000 | | 0.0000 | | |
| 16 | Comp Mole Frac (n-Butane) | | 0.00 | 00 | 0.0063 | 0.0000 | | 0.0000 | | |
| 17 | Comp Mole Frac (i-Pentane) | 1 | 0.00 | 00 | 0.0020 | 0.0000 | | 0.0000 | | |
| 18 | Comp Mole Frac (n-Pentane |) | 0.0000 | | 0.0019 | 0.0000 | | 0.0000 | | |
| 19 | Comp Mole Frac (n-Hexane) | | 0.0000 | | 0.0015 | 0.0000 | | 0.0000 | | |
| 20 | Comp Mole Frac (Nitrogen) | | 0.0000 | | 0.0051 | 0.0000 | | 0.0000 | | |
| 22 | Comp Mole Frac (TEGlycol) | | 0.32 | 82 | 0.0000 | 0.3282 | | 0.3283 | | |
| 23 | Comp Mole Frac (H2O) | | 0.67 | 0.6718 0.000 | | 0.6718 | | 0.6717 | 11000 | |
| 24 25 | 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 | | | | Energy Streams | | | Fluid Pkg: | All | |
| 26 | Name | (1.1/6) | Q-100 | 00 | Q-101 | Q-102 | Q-103 | 8000+006 | | |
| 27 | Heat Flow | (KJ/N) | 6.5000+0 | 00 | Unit Ops | 1.4556+007 | | .00000000 | | |
| 29 30 | Operation Name | Ope | ration Type | | Feeds | Products | | Ignored | Calc Level | |
| 31 32 | Absorber | Absorber | TE | | G,in t Gas | Dry Gas Rich TEG | | No | 2500 | |
| 33 | E-100 | Heater | | Ric O-1 | h TEG | Reboiled | | No | 500.0 | |
| 35 | Flash Drum | Separato | Separator | | poiled | Flash out | | No | 500.0 | |
| 36 37 | F 404 | Uset Eve | Evehanger | | in TEG,out | Flash Gas Pump,in | | No | 500.0 | |
| 38 | E-101 | Heat Exc | Heat Exchanger | | sh out | Rich TEG,in | | NO | 500.0 | |
| 39 | E-102 | Heat Exc | Heat Exchanger | | np,out | Cool TEG Hot Dry Gas | | No | 500.0 | |
| 40 | | | | Rich TEG.in | | Steam | | | | |
| 42 | T-101 | Distillation | | Q-102 | | Lean TEG,out Q-101 | | No | 2500 | |
| 44 | P-100 | Pump | 990 - C | Pump,in | | Pump,out | | No | 500.0 | |
| 45 | RCY-1 | Recycle | Paquela | | DI TEG | TEG.in | | No | 3500 | |
| 47 | Korr | recycle | | 1 000 | 1120 | 120,00 | | | | |
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